THE UNIVERSITY of TULSA

Department of Petroleum Engineering

Fluid Flow Projects

Seventy Second Semi-Annual Advisory Board Meeting Brochure and Presentation Slide Copy

March 25, 2009

Tulsa University Fluid Flow Projects
Seventy Second Semi-Annual Advisory Board Meeting Agenda
Wednesday March 25, 2009

Tuesday,
March 24, 2009

Tulsa University High-Viscosity Oil Projects
Advisory Board Meeting
University of Tulsa – Allen Chapman Activity Center (ACAC) - Gallery
440 South Gary
Tulsa, Oklahoma
8:00 – 11:30 a.m.

Tulsa University High-Viscosity Oil Projects and Tulsa University Fluid Flow Projects Workshop Luncheon
University of Tulsa – Allen Chapman Activity Center – (ACAC) – Chouteau C
440 South Garry
Tulsa, Oklahoma
11:30 – 1:00 p.m.

Tulsa University Fluid Flow Projects Workshop
University of Tulsa – Allen Chapman Activity Center (ACAC) - Gallery
440 South Gary
Tulsa, Oklahoma
1:00 – 3:00 p.m.

Tulsa University High-Viscosity Oil Projects, Tulsa University Fluid Flow Projects and Tulsa University Paraffin Deposition Projects
Tour of Test Facilities
University of Tulsa North Campus
2450 East Marshall
Tulsa, Oklahoma
3:30 – 5:30 p.m.

Tulsa University High-Viscosity Oil Projects and Tulsa University Fluid Flow Projects
Reception
University of Tulsa – Allen Chapman Activity Center (ACAC) – Alcove Room
2933 East 6th Street
Tulsa, Oklahoma
6:00 – 9:00 p.m.

Wednesday, March 25, 2009

Tulsa University Fluid Flow Projects
Advisory Board Meeting
University of Tulsa – Allen Chapman Activity Center (ACAC) - Gallery
440 South Gary
Tulsa, Oklahoma
8:00 a.m. – 5:00 p.m.

Tulsa University Fluid Flow Projects and Tulsa University Paraffin Deposition Projects
Reception
University of Tulsa – Allen Chapman Activity Center (ACAC) – Atrium Reading Room
2933 East 6th Street
Tulsa, Oklahoma
5:30 – 9:00 p.m.

Thursday, March 26, 2009

Tulsa University Paraffin Deposition Projects
Advisory Board Meeting
University of Tulsa – Allen Chapman Activity Center (ACAC) - Gallery
440 South Gary
Tulsa, Oklahoma
8:00 a.m. – 1:00 p.m.

Tulsa University Fluid Flow Projects
Seventy Second Semi-Annual Advisory Board Meeting Agenda
Wednesday March 25, 2009

8:00 a.m.	**Breakfast – Allen Chapman Activity Center - Gallery**	
8:30	**Introductory Remarks**	Cem Sarica
	Executive Summary	Cem Sarica
9:00	**TUFFP Progress Reports**	
	Liquid Entrainment in Annular Two-Phase Flow in Inclined Pipes	Kyle Magrini
	Modeling of Gas-liquid Flow in an Upward Vertical Annulus	Tingting Yu
10:30	**Coffee Break**	
10:45	**TUFFP Progress Reports**	
	Modeling of Oil-Water Pipe Flow	Anoop Sharma
	Slug Flow Evolution in Gas-Oil-Water Flow in Hilly-Terrain Pipelines	Gizem Ersoy-Gokcal
12:15 p.m.	**Lunch – Allen Chapman Activity Center – Chouteau C**	
1:15 p.m.	**TUFFP Progress Reports**	
	Effects of High Oil Viscosity on Slug Liquid Holdup in Horizontal Pipes	Ceyda Kora
	CFD Modeling of Drift Velocity of Viscous Two-Phase Horizontal Flow	Cem Sarica/ Abdel Al-Sarkhi
2:30	**Coffee Break**	
2:45	**TUFFP Project Reports**	
	Investigation of Slug Length for High Viscosity Oil-Gas Flow	Eissa Alsafran
	High Pressure – Large Diameter Multiphase Flow Loop	Scott Graham
3:45	**TUFFP Business Report**	Cem Sarica
4:00	**Open Discussion**	Cem Sarica
4:30	**Adjourn**	
5:30	**TUFFP/TUPDP Reception – McFarlin Library – Faculty Study**	

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Appendices

Executive Summary

Progress on each research project is given later in this Advisory Board Meeting Brochure. A brief summary of the activities is given below.

- *"Investigation of Gas-Oil-Water Flow"*. Three-phase gas-oil-water flow is a common occurrence in the petroleum industry. The ultimate objective of TUFFP for gas-oil-water studies is to develop a unified model based on theoretical and experimental analyses. A three-phase model has already been developed. There are several projects underway addressing three-phase flow.

- *"Oil-Water Flow in Pipes"*. Our three-phase model requires knowledge on oil/water interaction. Moreover, oil-water flow is of interest for many applications ranging from horizontal well flow to separator design. The objectives of this study are to assess performance of current models by checking them against experimental data and improving existing models through better closure relationships or developing new models if necessary.

  After completion of several experimental oil-water flow studies, efforts are concentrated on improvement of the modeling. A new modeling approach based on energy minimization is being developed. Comparisons with experimental data and other oil-water flow models prove that the newly developed model performs the best.

- *"High Viscosity Oil Two-phase Flow Behavior"*. Oils with viscosities as high as 10,000 cp are produced from many fields around the world. Current multiphase flow models are largely based on experimental data with low viscosity fluids. The gap between lab and field data may be three orders of magnitude or more. Therefore, current mechanistic models need to be verified with higher liquid viscosity experimental results. Modifications or new model developments are necessary.

  An earlier TUFFP study conducted by Gokcal showed that the performances of existing models are not sufficiently accurate for high viscosity oils. It was found that increasing oil viscosity had a significant effect on flow behavior. Mostly, intermittent flow (slug and elongated bubble) was observed in his study. Based on his results, this study focuses on slug flow.

  Drift velocity measurements for a horizontal pipe configuration made last fall indicated that drift velocity decreases with increasing liquid viscosity. Drift velocity measurements were completed for the entire range of upward inclination angles for a viscosity range of 200 – 1200 cp, and a drift flux model for horizontal flow was developed before the Spring 2008 Advisory Board Meeting. Since the spring Advisory Board Meeting, significant progress has been made on several fronts of this study. At the Fall 2008 Advisory Board Meeting, the translational velocity and slug frequency correlation development results were presented. Moreover, it was reported that slug lengths followed a log-normal distribution, and the average slug length decreases as the liquid viscosity increases. Currently, a new study to investigate slug length is underway. Dr. Eissa Al-Safran is working on this project as part of his sabbatical assignment with TUFFP.

  In another collaborative study with Dr. Abdel Al-Sarkhi of King Fahd Petroleum and Minerals University (KFPMU), drift velocity for high viscosity oils is being investigated using a CFD approach. A presentation on the results will be made at this Advisory Board Meeting.

  One of the important closure relationships for slug flow is slug liquid holdup. A current experimental study focuses on the investigation of slug liquid holdup. During this period, the feasibility of various experimental measurement methods was studied. A reliable technique was developed and will be implemented soon to conduct an extensive experimental campaign during the Summer of 2009.

- *"Droplet Homo-phase Interaction Study"*. There are many cases in multiphase flow where droplets are entrained from or coalesced into a continuous homophase. For example, in annular mist flow, the liquid droplets are in dynamic equilibrium with the film on the walls, experiencing both entrainment and coalescence. Very few mechanistic models exist for entrainment rate and coalescence rate. Understanding the basic physics of these phenomena is essential to model situations of practical interest to the industry. Droplet homo-phase covers a broad range of possibilities.

  A past sensitivity study of multiphase flow predictive models showed that, in stratified and annular flow, the variation of droplet entrainment fraction can significantly affect the predicted pressure gradient. Although better entrainment fraction correlations were proposed, a need was identified to experimentally investigate entrainment fraction for inclined pipes. The current study investigates entrainment fraction for various inclination angles. The 3-in. ID severe slugging facility is being utilized. A new device to measure entrainment fraction has been designed and constructed. Several tests have been performed following the Fall 2008 Advisory Board Meeting. The results show the dependency of entrainment fraction on the pipe inclination angle.

- *"Simplified Transient Flow Studies"*. TUFFP's simplified transient flow studies project proposal ranked #5 in our recent questionnaire. Therefore, it will be launched as a separate project as soon as we identify an appropriate graduate student or Research Associate.

- *"Low Liquid Loading Gas-Oil-Water Flow in Horizontal and Near Horizontal Pipes"*. Low liquid loading exists widely in wet gas pipelines. These pipelines often contain water and/or hydrocarbon condensates. Small amounts of liquid can lead to a significant increase in pressure loss along a pipeline. Moreover, the existence of water can significantly contribute to corrosion and hydrate formation problems. Therefore, understanding the flow characteristics of low liquid loading gas-oil-water flow is of great importance in the transportation of wet gas.

  In a previous study, a large amount of data was collected on various flow parameters, such as flow pattern, phase distribution, onset of droplet entrainment, entrainment fraction, and film velocity. The results revealed a new flow phenomenon.

  This study has been delayed due to insufficient performance of the student. A new graduate student has already been identified to take over the project and started his graduate studies in January 2009. Moreover, Dr. Yuxing Li, a visiting professor from China, joined TUFFP to spend his sabbatical with us. He has started to work with us on this project. We hope to accelerate the project with his involvement.

- *"Multiphase Flow in Hilly-Terrain Pipelines"*. Three-phase flow in hilly terrain pipelines is a common occurrence. The existence of a water phase in the system poses many potential flow assurance and processing problems. Most of the problems are directly related to the flow characteristics. Although two-phase gas-liquid flow has been investigated extensively, there are very few studies addressing multiphase gas-oil-water flow in hilly-terrain pipelines. The general objectives of this project are to thoroughly investigate and compare existing models, and develop closure relationships and predictive models for three-phase flow of gas-oil-water in hilly-terrain pipelines.

  Since the Fall 2008 Advisory Board Meeting, a significant amount of data has been acquired to capture three-phase flow characteristics of a hilly-terrain unit. Data analysis and model development are currently underway. Analysis of the data shows variation of in-situ water cut along the hilly-terrain section. Detailed progress is reported in this Advisory Board Meeting.

- *"Up-scaling Studies"*. One of the most important issues that we face in multiphase flow technology development is scaling up of small diameter and low pressure results to large diameter and high pressure conditions. Studies with a large diameter facility would significantly improve our understanding of flow characteristics in actual field conditions. Therefore, our main objective in this study is to investigate the effect of pipe diameter and pressures on flow behavior using a larger diameter flow loop.

  This project is one of the main activities of TUFFP and a significant portion of the TUFFP budget is being used for this facility. Facility construction is currently underway. In Mid-December, we applied for all the applicable building permits that we need from the City of Tulsa and are awaiting approval at this time. We are currently in the process of securing bids for both the concrete work and the structural steel fabrication. Our goal is to have all the concrete work done by the end of March and begin assembling the structural steel in April. The flow-loop itself will be constructed during the Summer of 2009. The Sundyne Gas compressor is on location as well as the 500KVA diesel generator that will provide electricity for the compressor and liquid pumps. Equipment that still needs to be purchased is as follows: Separation System, Pumps, Liquid Tanks, Surge Tank for gas, Flow-Meters, and Instrumentation.

- *"Gas-Liquid Flow in an Upward Vertical Annulus."* TUFFP has not conducted any study on this topic since Caetano's pioneering work in 1985. This project was initiated to improve our predictions for multiphase flow. A new mechanistic model has already been developed. The new model is an extension of the TUFFP Unified Model to annulus flow. The new model performs better than the original Caetano and the current TUFFP Unified model.

- "*Unified Mechanistic Model"*. TUFFP maintains, and continuously improves upon the TUFFP unified model. Since the last Advisory Board Meeting no significant change has been made in the model.

Current TUFFP membership stands at 16 (15 industrial companies and MMS). Tenaris and Petronas terminated their membership for 2009. Scandpower Petroleum Technology (SPT) has joined TUFFP in 2009. Efforts continue to further increase the TUFFP membership level. Drs. Cem Sarica and Holden Zhang visited PetroChina and Chinese National Offshore Oil Company (CNOOC) in Beijing in June 2008. It is anticipated that CNOOC will join TUFFP in 2009. A detailed financial report is provided in this report. We thank our members for their continued support.

Several related projects are underway. The related projects involve sharing of facilities and personnel with TUFFP. The Paraffin Deposition consortium, TUPDP, is into its third phase with 11 members. The Center of Research Excellence (TUCoRE) initiated by Chevron at The University of Tulsa funds several research projects. TUCoRE activities in the area of Heavy Oil Multiphase Flow have resulted in a new Joint Industry Project (JIP) to investigate Heavy Oil Multiphase Flow in more detail. The JIP currently has three members.

Fluid Flow Projects

72nd Fluid Flow Projects
Advisory Board Meeting

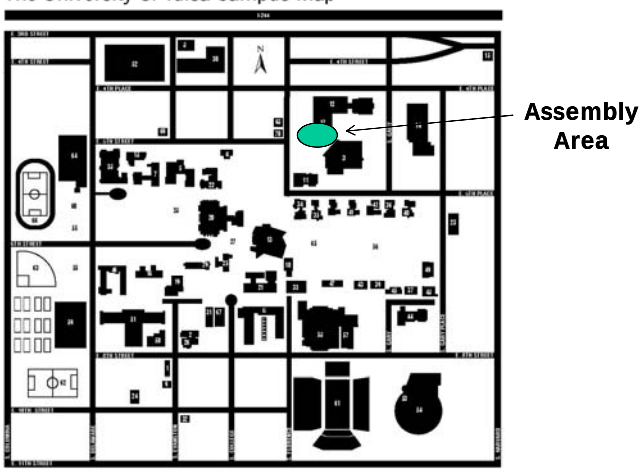
Welcome

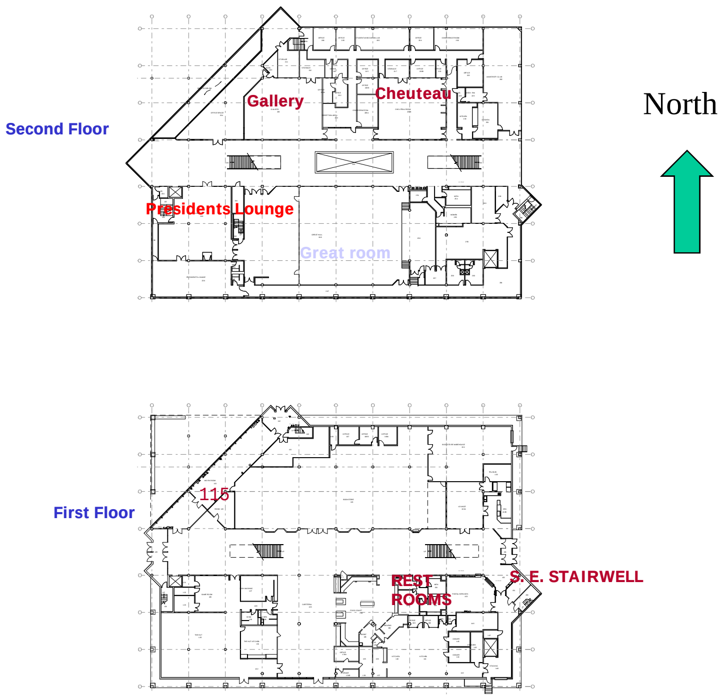
Safety Moment

- Emergency Exits
- Assembly Point – Grassy Area to Northwest
- Tornado Shelter
  - Room 115, Southeast Emergency Stairwell
  - Lower Level Restrooms
- Campus Emergency
  - Call 9-911
  - Campus Security, ext. 5555 or 918-631-5555
- Rest Rooms

Fluid Flow Projects

The University of Tulsa Campus Map





Introductory Remarks

- 72nd Semi-Annual Advisory Board Meeting
- Handout
  - Combined Brochure and Slide Copy
- Sign-Up List
  - Please Leave Business Card at Registration Table

Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Team

- Research Associates
  - Cem Sarica (Director)
  - Holden Zhang (Associate Director)
  - Jim Brill (Director Emeritus)
  - Polat Abduvayt
  - Mingxiu (Michelle) Li
  - Eissa Al-Safran, (On Sabbatical Leave)
  - Yuxing Li (On Sabbatical Leave)

Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Team …

- Project Coordinator
  - Linda Jones
- Project Engineer
  - Scott Graham
- Research Technicians
  - Craig Waldron
  - Brandon Kelsey
- Computer Manager and Web Master
  - James Miller

Team …

- TUFFP Research Assistants
  - Gizem Ersoy (Ph.D.) – Turkey
  - Kiran Gawas (Ph.D.) – India
  - Ceyda Kora (MS) – Turkey
  - Kyle Magrini (MS) – USA
  - Anoop Sharma (MS) – India
  - Tingting Yu (MS) - PRC

Guests

- Jep Bracey, BHP
- Prof. Jing Gong, China University of Petroleum at Beijing
- David Fouchard, Nalco

Agenda

- 8:30 Introductory Remarks and Executive Summary
- 9:00 Progress Reports
  - Liquid Entrainment in Annular Two-Phase Flow in Inclined Pipes
  - Modeling of Gas-liquid Flow in an Upward Vertical Annulus
- 10:30 Coffee Break

Agenda …

- 10:45 Progress Reports
  - Modeling of Oil-Water Pipe Flow
  - Slug Flow Evolution in Gas-Oil-Water Flow in Hilly-Terrain Pipelines
- 12:15 Lunch – Chouteau-C
- 1:15 Progress Reports
  - Effects of High Oil Viscosity on Slug Liquid Holdup in Horizontal Pipes
  - CFD Modeling of Drift Velocity of Viscous Two-Phase Horizontal Flow

Agenda …

- 2:30 Coffee Break
- 2:45 Progress Reports
  - Investigation of Slug Length for High Viscosity Oil-Gas Flow
  - High Pressure – Large Diameter Multiphase Flow Loop

Agenda ...

- 3:45 TUFFP Business Report
- 4:00 Open Discussion
- 4:30 Adjourn
- 5:30 TUFFP/TUPDP Reception (McFarlin Library – Faculty Study)

Other Activities

- March 24, 2009
  - TUHOP Meeting
  - TUFFP Workshop
    - Excellent Presentations
    - Beneficial for Everybody
  - Facility Tour
- March 26, 2009
  - TUPDP Meeting

Fluid Flow Projects

Executive Summary
of Research Activities

Cem Sarica

Advisory Board Meeting, March 25, 2009

Droplet Homo-phase Studies

- Significance
  - Better Predictive Tools Lead to Better Design and Practices
- General Objective
  - Development of Closure Relationships
- Past Study
  - Earlier TUFFP Study Showed
    - Entrainment Fraction (FE) is Most Sensitive Closure Parameter in Annular Flow
    - Developed New FE Correlation
      - Utilizing In-situ Flow Parameters
      - Limited Data, Especially for Inclined Flow Conditions

Droplet Homo-phase Studies …

- Current Study
  - Liquid Entrainment in Annular Two-Phase Flow in Inclined Pipes
  - Objectives
    - Acquire Data for Various Inclination Angles for 3-in. ID Pipe Using Severe Slugging Facility
      - Existing Data are for 1 and 1 ½ in.
    - Develop a New Closure Relationship

Droplet Homo-phase Studies …

- Status
  - New Dimensionless Groups are Proposed to Correlate Entrainment Fraction
  - Facility Modifications are Completed
  - Experimental Study is Underway
    - Almost Half of the Tests Completed
    - Entrainment Fraction is Found to Vary with Inclination Angle
  - Will Be Completed by Summer of 2009

Fluid Flow Projects

Liquid Entrainment in Annular Gas-Liquid Flow in Inclined Pipes

Kyle Magrini

Advisory Board Meeting, March 25, 2009

Outline

- Objectives
- Introduction
- Experimental Facility
- Experimental Results
- Summary
- Future Work
- Project Schedule

Objectives

- Acquire Experimental Data of Entrainment Fraction in Two-Phase Gas-Liquid Annular Flow for Inclination Angles of 0^o, 10^o, 20^o, 45^o, 75^o, and 90^o from Horizontal
- Compare Data with Current Correlation and Model Predictions
- Improve Existing Models with New Correlation

Introduction

- Multiphase Flow Mechanistic Models are Tools in Multiphase Design and Applications
  - Pressure Gradient
  - Liquid Holdup
  - Temperature Gradient
  - Etc.

Introduction …

- These Mechanistic Models (e.g. TUFFP Unified Model) Require Closure Relationships
  - Interfacial Friction Factor
  - Droplet Entrainment Fraction
  - Slug Translational Velocity
  - Etc.

Introduction …

- Chen (2005a) Sensitivity Study Showed that for Annular Flow the TUFFP Unified Model and Xiao Model are Most Sensitive to Droplet Entrainment Fraction Compared to Other Closure Relationships

Experimental Facility

- 3 inch Severe Slugging Flow Loop

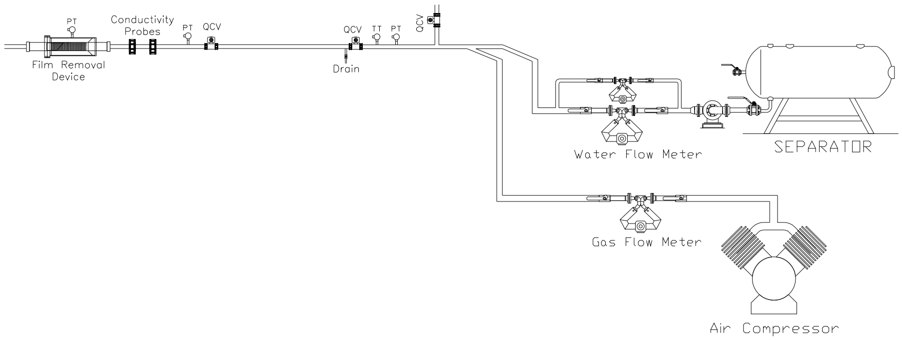
Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Experimental Facility

- Test Section 180 Diameters from Inlet to Ensure Fully Developed Flow
- Installation of Quick Closing Valves to Measure Local Liquid Holdup
- Installation of Conductivity Probes to Measure Wave Characteristics
- Conduct Tests at Horizontal and Inclination Angles of 10º, 20º, 45º, 75º, and 90º
- Measurement of Entrainment Fraction and Deposition Rate

Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Experimental Facility …

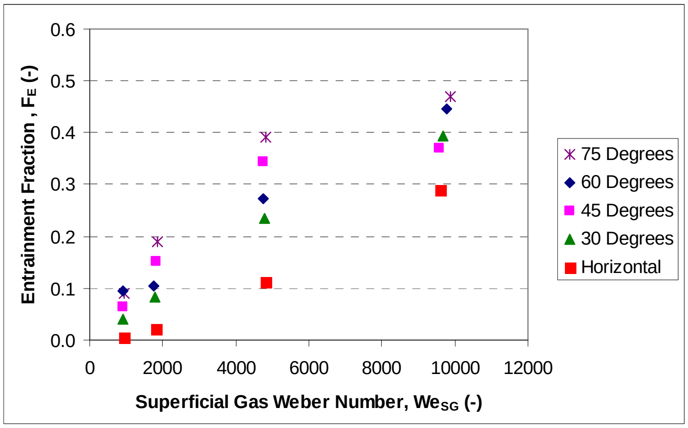


Testing Range

- Inclined Experimental Work
  - Ousaka (1996)
    - I inch Pipe
    - Angles = 0^o, 30^o, 45^o, 60^o, 75^o
    - v_{SL} = 0.06, 0.1, 0.2 m/s
    - v_{SG} = 15, 20, 30, 40 m/s
  - Geraci (2008)
    - 1.5 inch Pipe
    - Angles = 0^o, 30^o, 45^o, 60^o, 85^o
    - v_{SL} = 0.007, 0.011, 0.02, 0.033 m/s
    - v_{SG} = 15, 21 m/s

Testing Range …

- Ousaka (1996)
  - Re_{SL} = 1560



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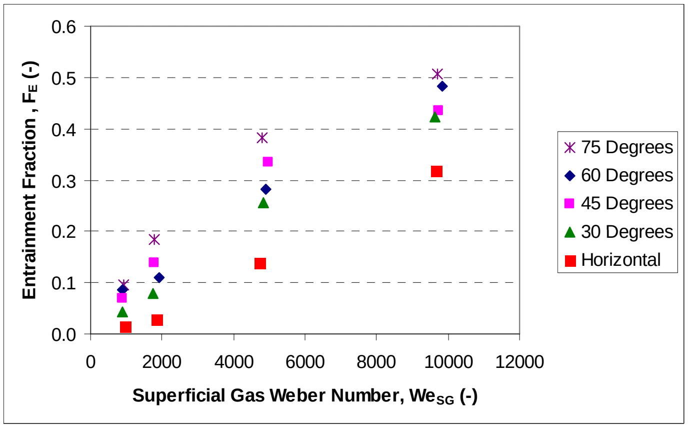
Testing Range …

- Ousaka (1996)
  - Re_{SL} = 2600



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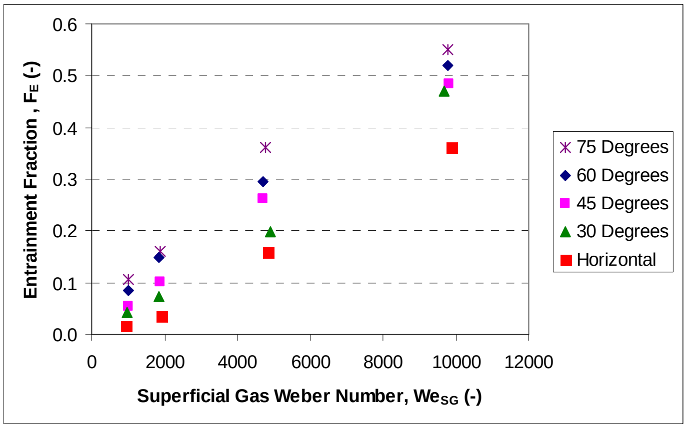
Testing Range …

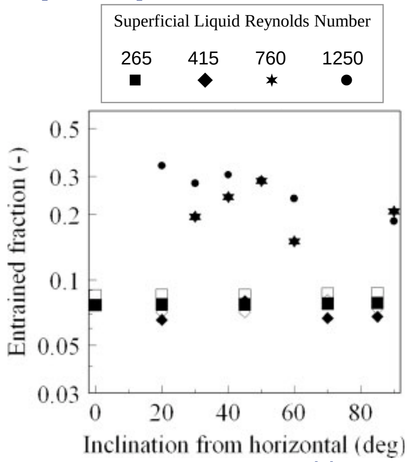
- Ousaka (1996)
  - $Re_{SL} = 5200$



Fluid Flow Projects
Advisory Board Meeting, March 25, 2009

Testing Range …

- Geraci (2008)



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Advisory Board Meeting, March 25, 2009

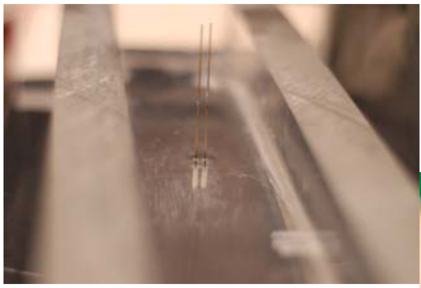
Testing Range …

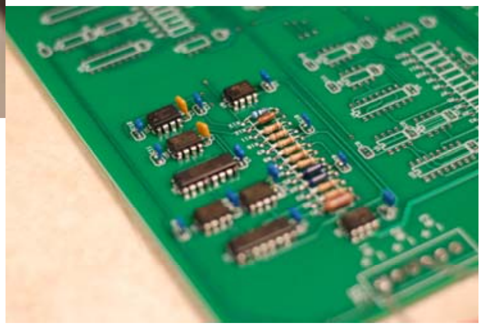
- Superficial Water Velocities Range from 0.005 to 0.05 m/sec
- Superficial Gas Velocities Range from 30 to 80 m/sec
- Maximum Entrainment Fraction will be Measured
- Surface Tension Measurements will be Conducted to Ensure Valid Results

Measurement Techniques

- Conductivity Probe
- Film Removal Device
- Iso-kinetic Probe

Conductivity Probe





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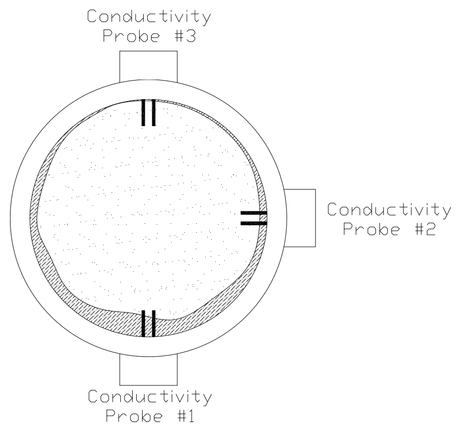
Conductivity Probe …

- Conductivity Probe
  - Measure Wave Characteristics
    - Frequency
    - Wavelength
    - Amplitude
    - Wave Velocity (Celerity)
  - Measure Film Thickness
  - Parallel Wire Configuration

Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Conductivity Probe …

- Configuration



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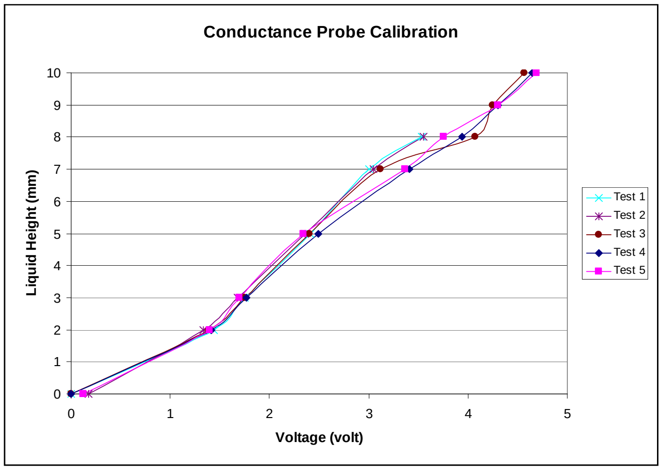
Conductivity Probe …

- Calibration



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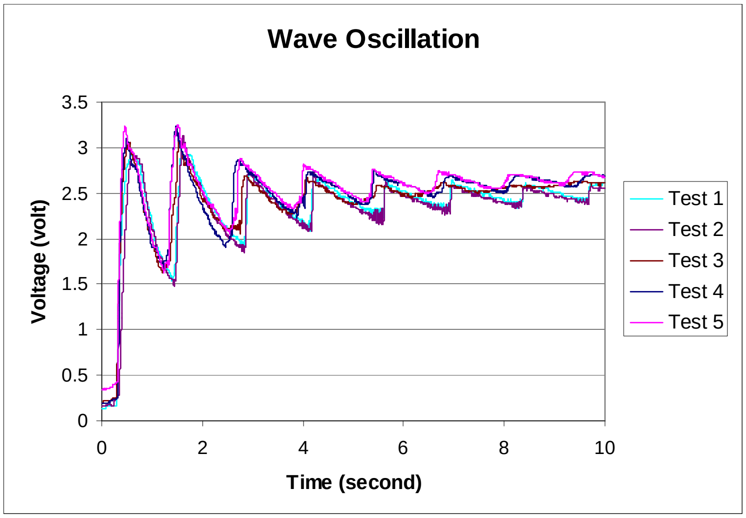
Conductance Probe …



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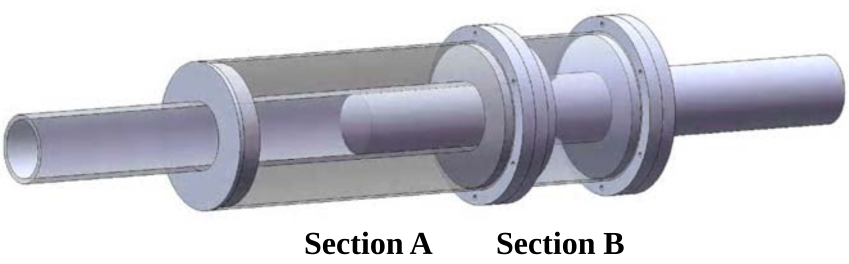
Film Removal Device



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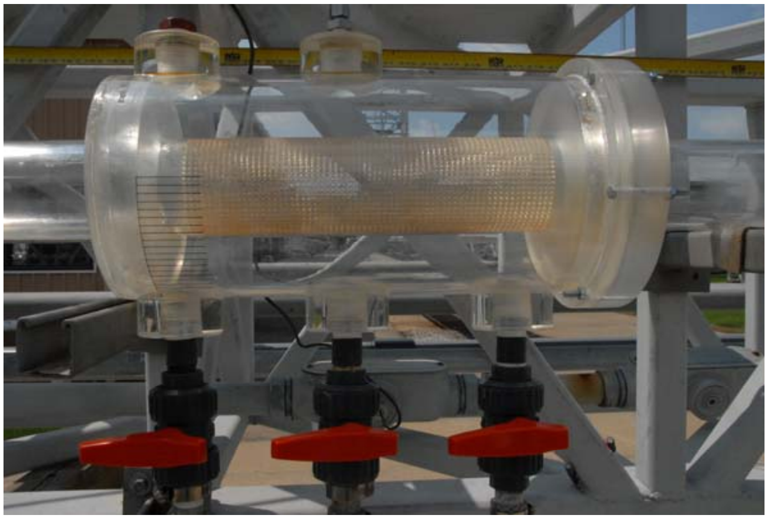
Film Removal Device …



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Film Removal Device …

- Film Removal Device Section A
  - Measurement of Entrainment Fraction
  - Liquid Film is Stripped through Porous Section
  - Film Flow Rate will be Obtained
  - Entrainment Fraction will be Obtained:

$$F_E = 1 - \frac{q_{Film}}{q_{Liquid}}$$

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Film Removal Device …

- Film Removal Device Section B
  - Measurement of Droplet Deposition Rate
  - Liquid Film is Stripped through Porous Section Similar to Section A
  - Film Volume will be Measured Over Time to Determine Deposition Rate

Fluid Flow Projects

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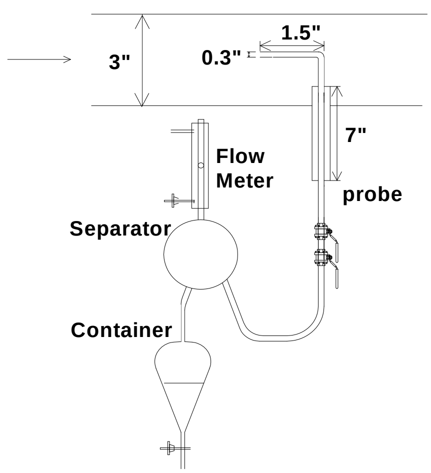
Iso-kinetic Sampling Probe



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Iso-kinetic Sampling Probe …

Iso-kinetic Sampling Probe …

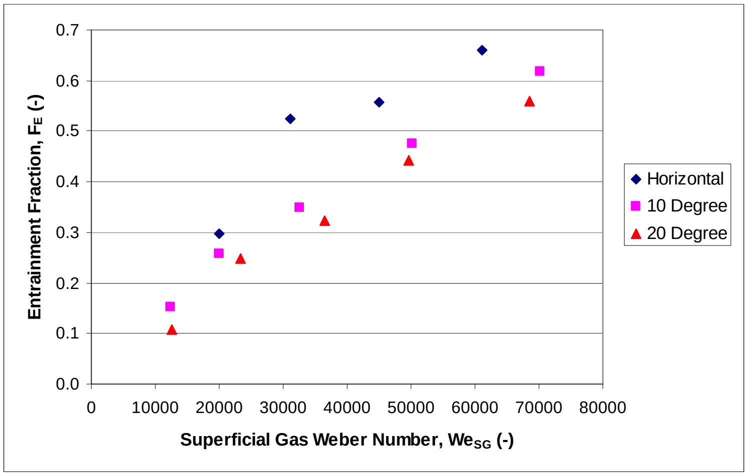
- Iso-kinetic Sampling Probe
  - Entrained Droplets are Sampled Over a Given Length of Time at Various Radial Distances
  - Entrainment Flux Profile is Created
  - Entrainment Fraction is Calculated by Integrating Flux Profile
  - Most Accurate Under Low Liquid Flow Rates
  - Iso-kinetic Conditions Only Reached at Low Gas Flow Rates

Experimental Results

- Experiments Completed at Inclination Angles 0º, 10º, and 20º from Horizontal
- Entrainment Fraction vs. Superficial Gas Weber Number Plotted at Constant Superficial Liquid Reynolds Number
- Results Indicate a Decrease in Entrainment Fraction for Increasing Inclination Angles
- Further Tests Needed to Clarify Effect

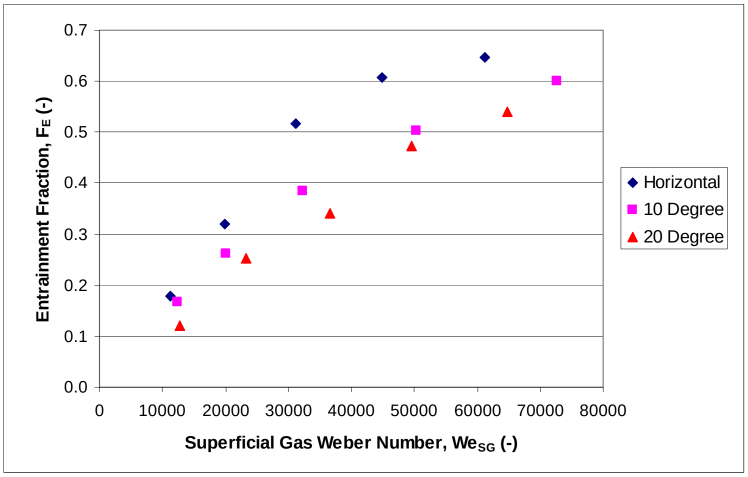
Experimental Results …

- $Re_{SL} = 150$



Experimental Results …

- Re_{SL} = 230



Fluid Flow Projects

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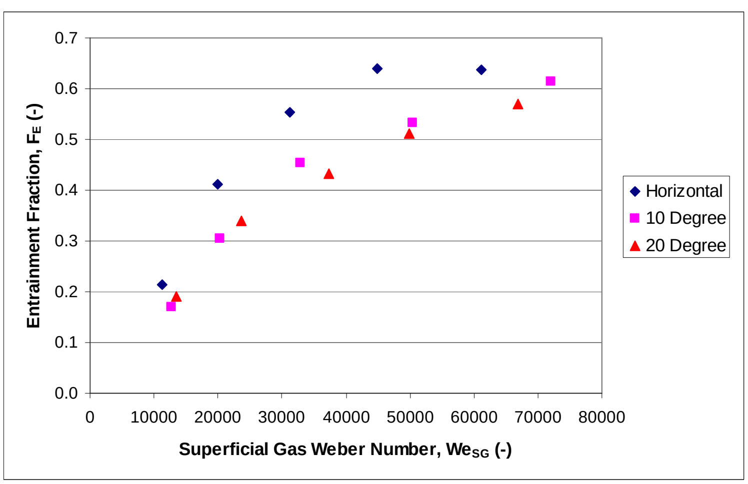
Experimental Results …

- Re_{SL} = 380



Fluid Flow Projects

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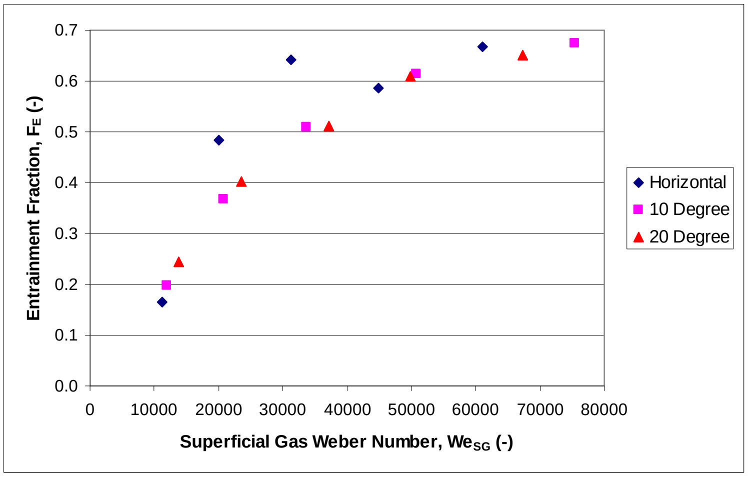
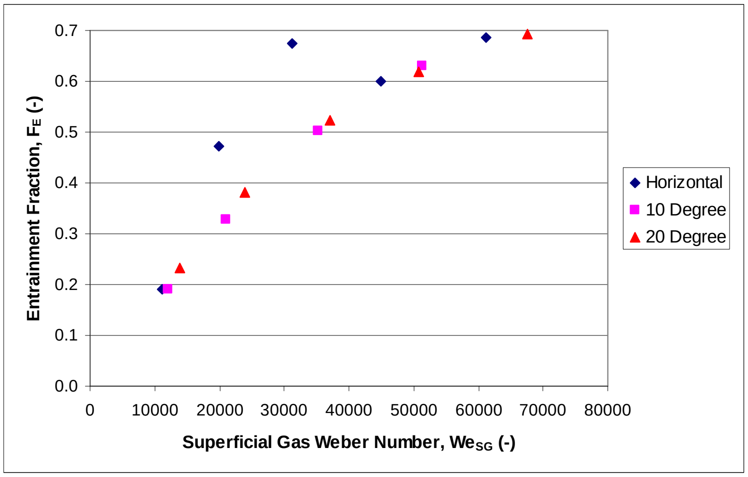
Experimental Results …

- Re_{SL} = 760



Experimental Results …

- Re_{SL} = 1520



Summary

- Testing Range has been Altered to Coincide with Available Entrainment Experiments in Inclined Pipes
- Conductivity Probes Added to Facility to Measure Wave Characteristics
- 75 of the Proposed 150 Experiments have been Completed
- Preliminary Results Indicate Pipe Inclination has Negative Effect on Entrainment at Low Liquid Flow Rates

Future Work

- Acquire Entrainment Data, Liquid Holdup, and Wave Characteristics for Various Flow Rates and Inclination Angles
- Compare Results of Film Removal Device with Those Obtained from Iso-kinetic Probe at Low Flow Conditions
- Validate Existing Models with Experimental Data
- Improve Existing Models with New Correlation

Project Schedule

Literature Review	Ongoing
Facility Construction	Completed
Data Acquisition	April 2009
Model Comparison	April 2009
Final Report	May 2009

Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Questions/Comments

?

Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Liquid Entrainment in Annular Gas-Liquid Flow in Inclined Pipes

Kyle Magrini

PROJECTED COMPLETION DATES:

Literature Review	Completed
Facility Modifications	Completed
Preliminary Correlation Development	Completed
Testing	April 2009
Model and Correlation Validation	April 2009
Final Report	May 2009

Objectives

The objectives of this study are:

- to acquire liquid entrainment data in two-phase gas-water annular flow through pipes from horizontal to near vertical,
- to validate current correlations and models with experimental results, and
- to improve current models, if necessary, or develop a new model.

Introduction

Annular flow usually occurs at high gas velocities and low to medium liquid velocities. The liquid flows as a film along the wall of the pipe and as droplets entrained in the gas core. The interface between the gas core and liquid film is usually very wavy, causing atomization and deposition of liquid droplets. Under equilibrium conditions, the rates at which the droplets atomize and deposit becomes equal, resulting in a steady fraction of the liquid being entrained as droplets, F_E. This critical parameter is crucial to understand and model the behavior of annular flow.

Most multiphase flow prediction models (including the TUFFP unified mechanistic models) are based on a simplified (one-dimensional) two-fluid model in which empirical closure relationships (i.e. interfacial friction factor, interfacial area, droplet entrainment fraction, etc.) are needed. The performance of the multiphase flow models is determined by the accuracy and physical completeness of these closure relationships. The literature reveals that sufficient physics of multiphase flow may not be contained in these empirical closure relationships. Therefore, further refinements of these closure relationships can significantly improve the performance of multiphase mechanistic models.

Chen (2005) conducted a sensitivity study to investigate the influence of individual closure relationships on the predictions of a multiphase mechanistic model. The study showed that in annular flow the variation in droplet entrainment fraction can substantially affect the predicted pressure gradient and liquid hold-up. Thus, the use of an accurate predictive model for entrainment fraction is imperative.

Literature Review

The liquid droplet entrainment phenomenon is very complicated. Various factors, such as pipe size, pipe orientation, velocity, fluid properties, and wave characteristics, control the process. There are several studies devoted to understanding the different aspects of liquid entrainment. Many of these studies were presented at the April 2008 ABM, along with the various correlations found in literature for different pipe orientations. The literature review will be an ongoing task until this study is completed.

Experimental Study

TUFFP's 76.2-mm (3-in.) diameter severe slugging facility (Fig. 1) has been modified for this experimental study. The facility is capable of being inclined from horizontal to vertical. Pressure and temperature transducers have been placed near the test section to obtain fluid properties and flow characteristics that are used in the entrainment fraction correlations. Quick-closing valves have been installed on the facility to measure the local liquid holdup of the flow. Conductivity probes will be installed to obtain information on wave characteristics.

The test section used to obtain entrainment fraction was placed 180d (15.24 m) from the entrance to ensure fully developed flow. Experiments for entrainment fraction will be conducted at inclination angles of 0^o, 10^o, 20^o, 45^o, 75^o, and 90^o from horizontal. Iso-kinetic sampling and liquid film removal will be used to calculate the entrainment fraction.

Test Fluids

Compressed air and Tulsa city tap water are used in this study. The surface tension of the tap water is tested frequently to ensure accurate results.

Testing Range

The testing range presented at the September 2008 ABM has been modified to coincide with two known inclined experiments in the literature. Geraci (2007) conducted entrainment experiments at low superficial liquid velocities (0.007, 0.011, 0.02, and 0.033 m/s). Geraci noted that for the flow conditions studied, pipe inclination had little or no effect on liquid entrainment fraction. His results can be found in Fig. 2. Ousaka (1996) conducted experiments at higher superficial liquid velocities (0.06, 0.1, 0.2 m/s) and noted a strong inclination effect on entrainment (Figs. 3-5). In these figures, the superficial liquid Reynolds number, Re_{SL}, is defined as

$$\mathrm{Re}_{SL} = \frac{v_{SL}\rho_L d}{\mu_L}, \quad (1)$$

and the superficial gas Weber Number, We_{SG}, defined by Ishii and Mishima (1989) as

$$We_{SG} = \frac{\rho_L v_{SG}{}^2 d}{\sigma}\left(\frac{\rho_L - \rho_G}{\rho_G}\right)^{1/3}. \quad (2)$$

To clarify the effect of pipe inclination on liquid entrainment, superficial velocities were chosen within and between the ranges used in these previous two studies. Superficial water velocities range from 0.005 to 0.05 m/sec. Superficial gas velocities range from 30 to 80 m/sec.

Conductivity Measurement

Cousins and Hewitt (1968) identified disturbance waves as the major source of entrained droplets in annular flow. To better understand these disturbance waves, conductivity probes will be installed to measure wave properties and liquid film thickness. The test section will include two conductivity test locations. Each location will consist of three conductivity probes located at the top, bottom, and side of the pipe (Fig. 6). The conductivity probe, consisting of two parallel wires separated by 3 mm, along with a conductivity meter, will generate an analog signal based on the liquid wave height across the probe. From the analog signal of each probe and calibration curve, wave characteristics such as frequency, wavelength, and amplitude will be determined. The wave velocity (celerity) will also be calculated using a cross correlation technique between the probes at the two locations. Using the three conductivity points at each location, the film thickness profile can also be generated.

Film Removal Device

The procedure for measuring entrainment fraction in the test section involves removing the liquid film from the wall of the pipe, while allowing droplets entrained in the gas phase to continue flowing. The entrained liquid flow rate will be calculated by subtracting the liquid film flow rate from the total liquid flow rate. The specially designed test section is shown in Figs. 7 and 8. Section A is similar to the one used by Hay et al. (1996), Azzopardi et al. (1996), Simmons and Hanratty (2001), and Al-Sarkhi and Hanratty (2002). The flow passes through a porous section and the liquid film, traveling at a lower velocity than the gas core, is pushed through the porous section. The high inertia of the droplets in the gas core, flowing close to the gas velocity, prevents them from being removed through the porous section. To ensure no droplets will escape, a long sleeve will be inserted close to where the liquid film dissipates. This sleeve will be moved in and out in the pipe to make sure the liquid film passes under the sleeve and only the gas core passes through the test section.

To ensure the accurate measurement of the entrainment fraction, the test section will be held at constant pressure. The removed liquid film will be collected in an enclosed 6-in. pipe section. A pigging system will be implemented to accurately measure the liquid film volume. The volume of water from the liquid film and time will be measured to determine the film flow rate and entrainment fraction.

The deposition rate can be measured after the liquid film is stripped in Section A of the test section. In Section B of Figs. 7 and 8, the film is once again stripped from the flow through a porous section. The deposition rate of the droplets is calculated based on the accumulation, liquid amount, and stripping area.

Iso-kinetic Sampling Probe

An iso-kinetic sampling probe (Fig. 9) has also been installed in the facility to measure entrainment fraction. The iso-kinetic sampling probe will be inserted into the pipe at various radial distances. The liquid sampled from the gas core will be separated in a small gas-liquid separator and collected in a graduated cylinder. From these measurements, the droplet entrainment flux profile will be determined. The entrainment fraction can be calculated by integrating this flux profile. The iso-kinetic sampling probe works best at low liquid flow rates where a more distinct division between the gas core and liquid film exists. From preliminary testing, the ability to reach iso-kinetic conditions was limited to low gas flow rates. Due to these limitations, iso-kinetic measurements will be limited to low liquid and gas flow rates. The results of the iso-kinetic sampling probe will be used in validating the results obtained from the film removal device at the lower flowing conditions.

Experimental Results

Currently, over 75 of the proposed 150 experiments have been completed, which include results at inclination angles of 0^o, 10^o, and 20^o from horizontal. Figures 10-14 compare the entrainment measurements for the three inclination angles. Each figure is at a constant superficial liquid Reynolds Number and displays the entrainment fraction vs. superficial gas Weber Number for the three inclination angles. From these figures, the pipe inclination seems to have a negative effect on liquid entrainment but is difficult to distinguish. This effect disagrees with Ousaka who found that entrainment fraction increases as the inclination angle increases. However, Ousaka had no data between horizontal and 30^o to compare. More entrainment data at differing inclination angles is needed to clarify this effect. The remaining experimental results will be obtained for inclination angles of 45^o, 75^o, and 90^o from horizontal. Entrainment results will first be obtained at an inclination angle of 90^o from horizontal (vertical). These results will be compared with entrainment data from horizontal to determine if an inclination effect on entrainment fraction is present. Once this effect is verified, entrainment tests at lower inclination angles will be conducted.

Future Tasks

The main tasks for the future are:

- Complete the facility modifications,
- Conduct experiments,
- Validate correlations,
- Modify or develop new correlations.

Nomenclature

d = pipe diameter [m]

F_E = entrainment fraction

Re = Reynolds number

v = velocity [m/s]

We = Weber number

Greek Letters

μ = viscosity [kg/ms]

ρ = density [kg/m^3]

σ = surface tension [N/m]

Subscripts

G = gas phase

L = liquid phase

SG = superficial gas

SL =superficial liquid

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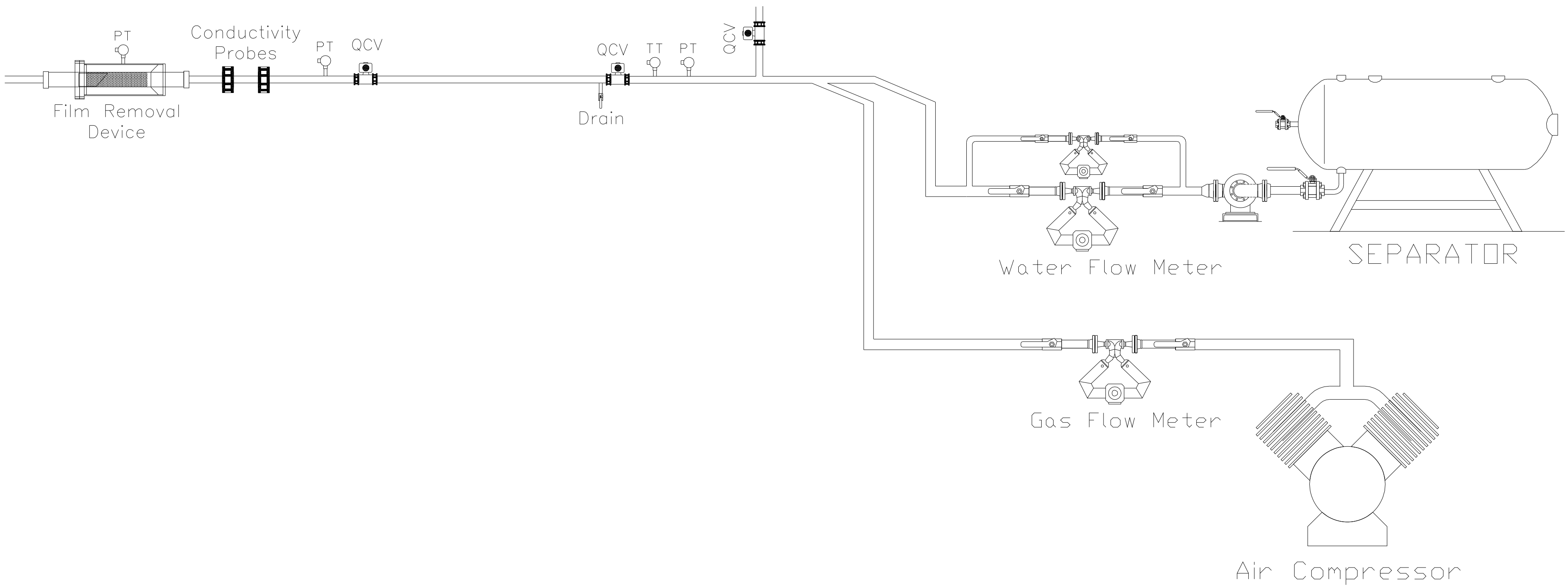


Figure 1. Facility Schematic

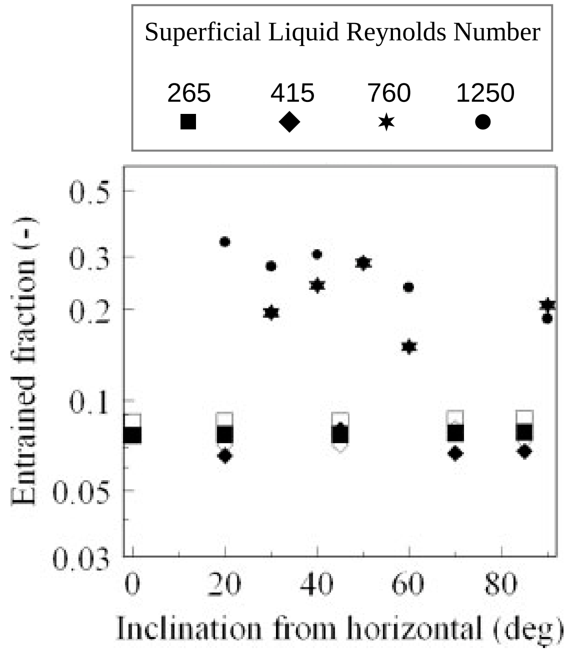


Figure 2. Geraci (2007) Entrained Fraction Variation with Inclination Angle from Horizontal.
Open symbols: v_{SG} = 21.5 m/s (We_{SG} = 4350)
Closed symbols: v_{SG} = 15 m/s (We_{SG} = 2120).
Data indicated by ●, ✶ are from Azzopardi et al. (1997): v_{SG} = 15 m/s (We_{SG} = 2120).

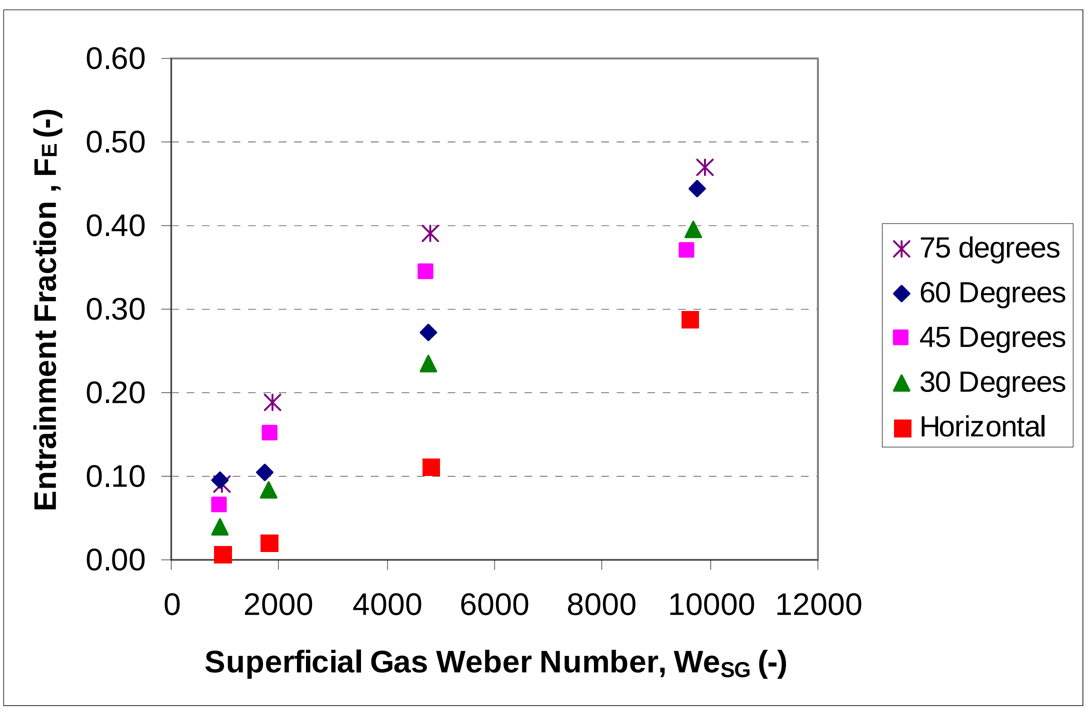


Figure 3. Ousaka (1996) Entrainment Fraction Data at Re_{SL} = 1560 for Various Inclination Angles.

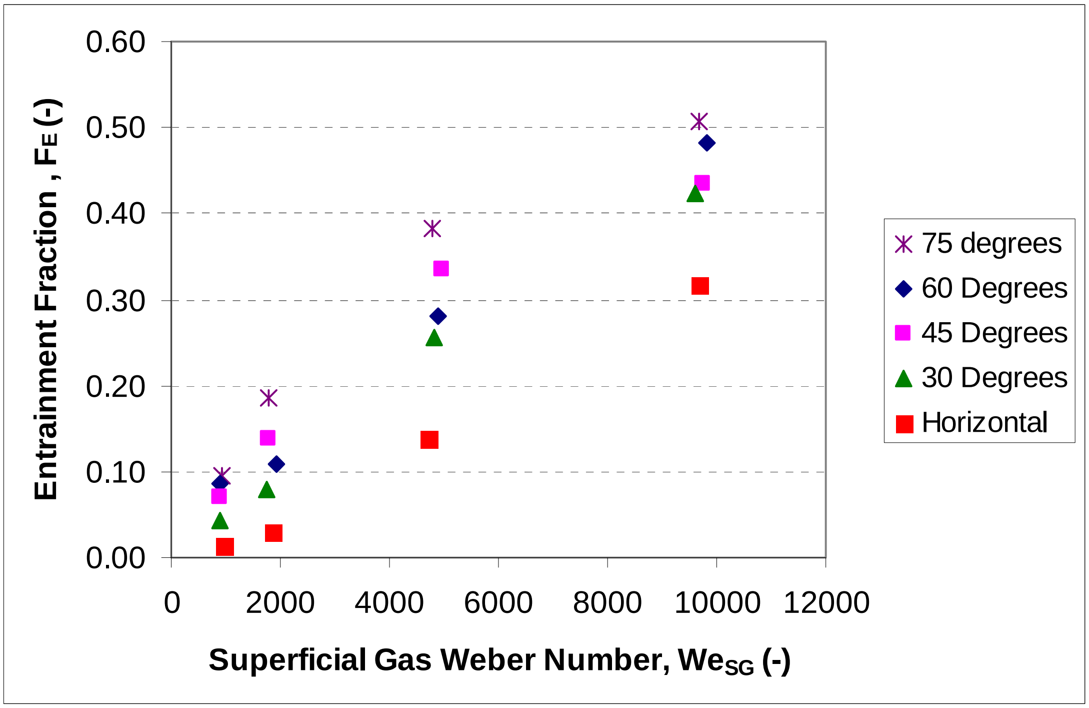


Figure 4. Ousaka (1996) Entrainment Fraction Data at Re_{SL} = 2600 for Various Inclination Angles.

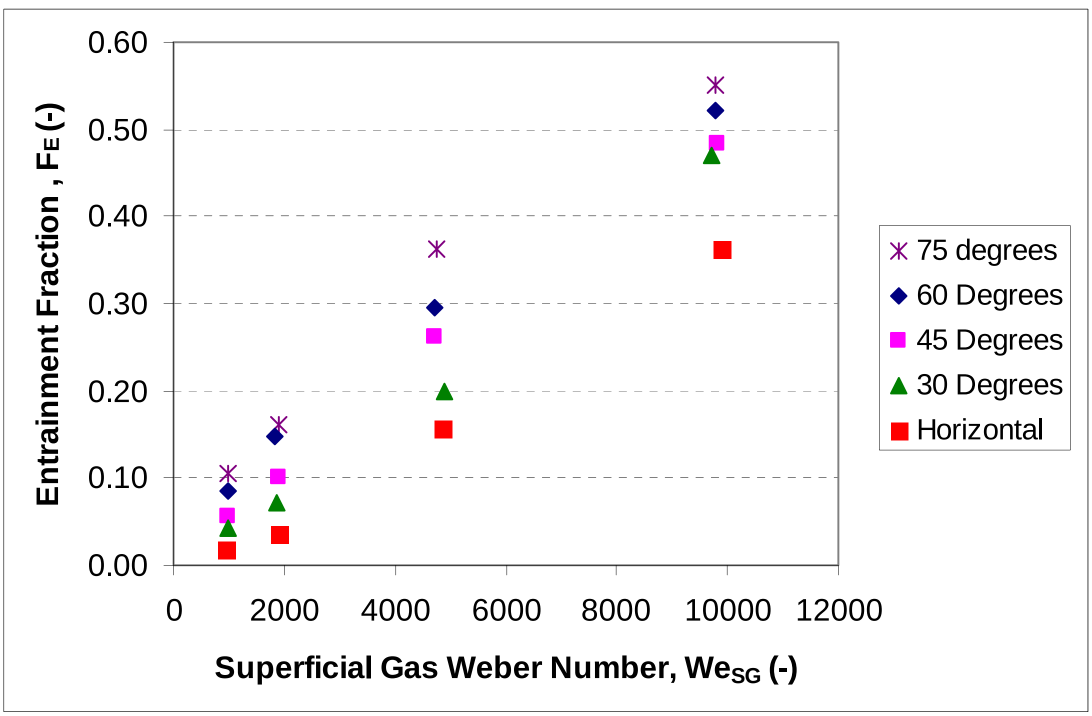


Figure 5. Ousaka (1996) Entrainment Fraction Data at Re_{SL} = 5200 for Various Inclination Angles.

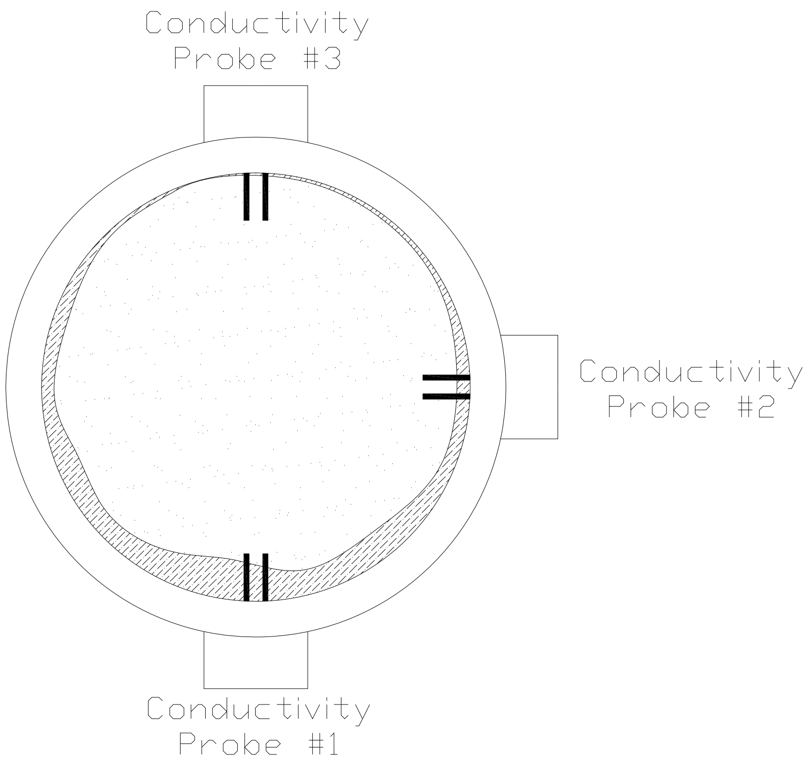


Figure 6. Configuration of Conductivity Probes

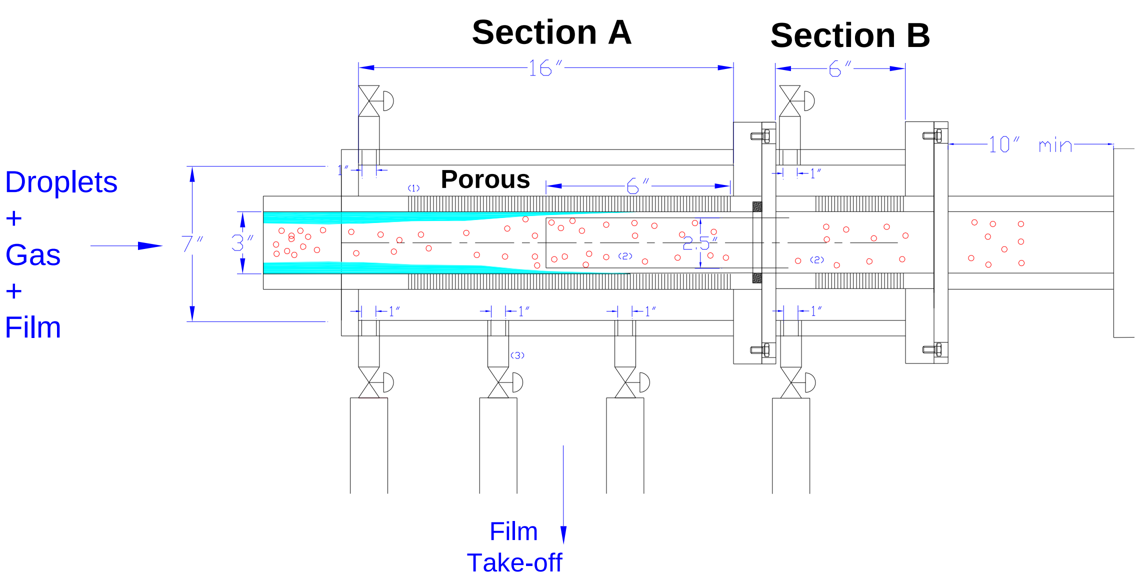


Figure 7. Film Removal Device Schematic

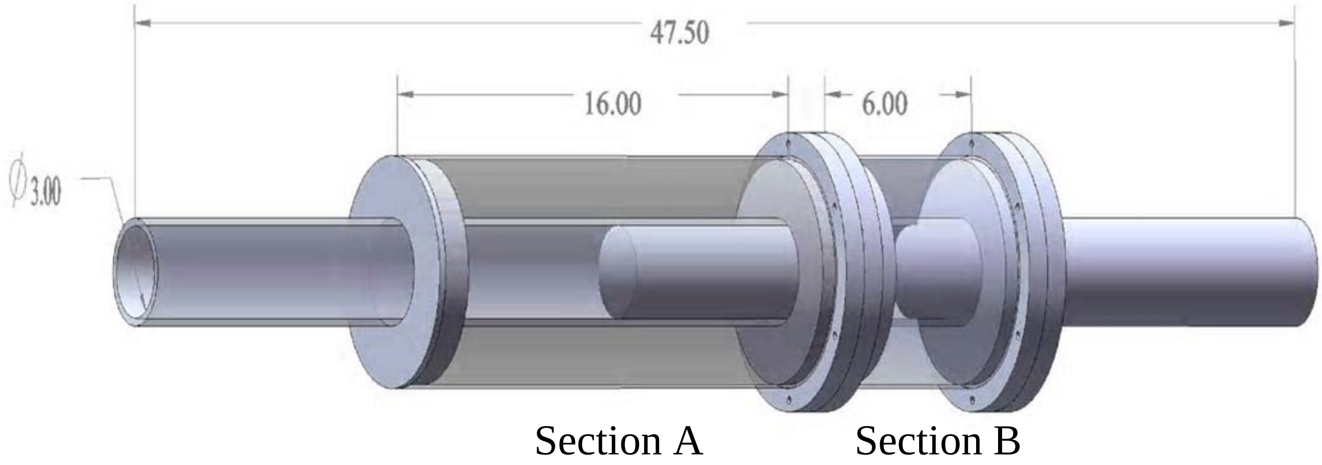


Figure 8. Film Removal Device Drawing

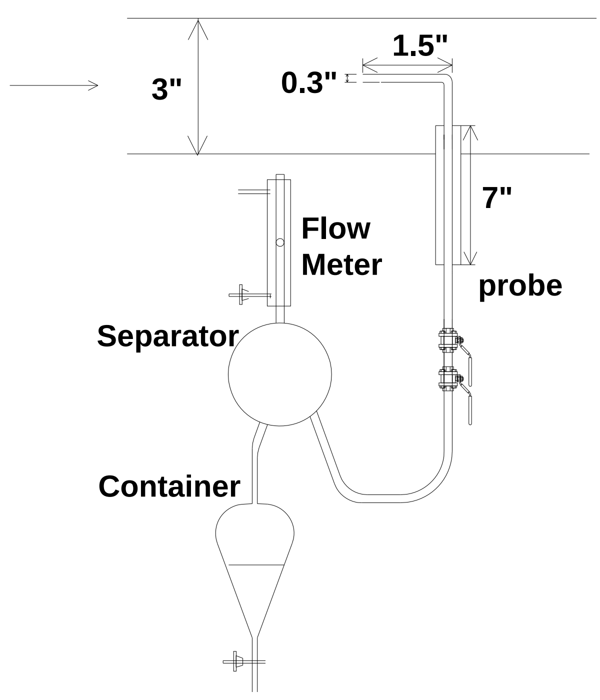


Figure 9. Iso-Kinetic Sampling System

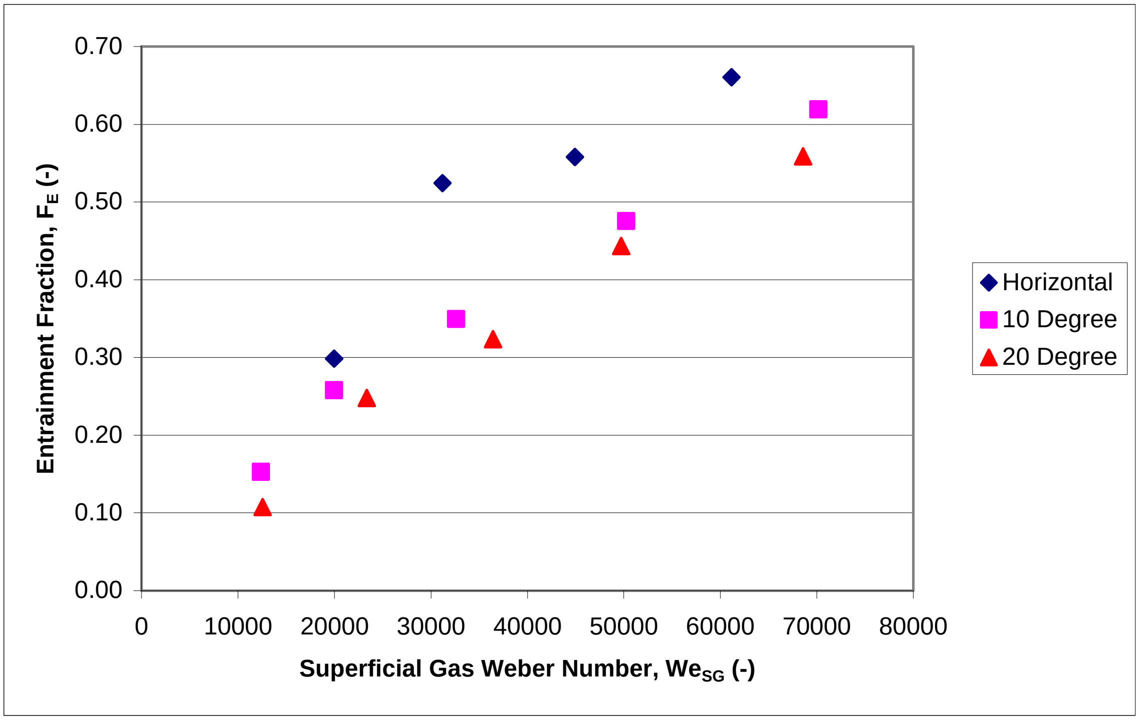


Figure 10. Entrainment Fraction Data at Re_{SL} = 150 for Various Inclination Angles.

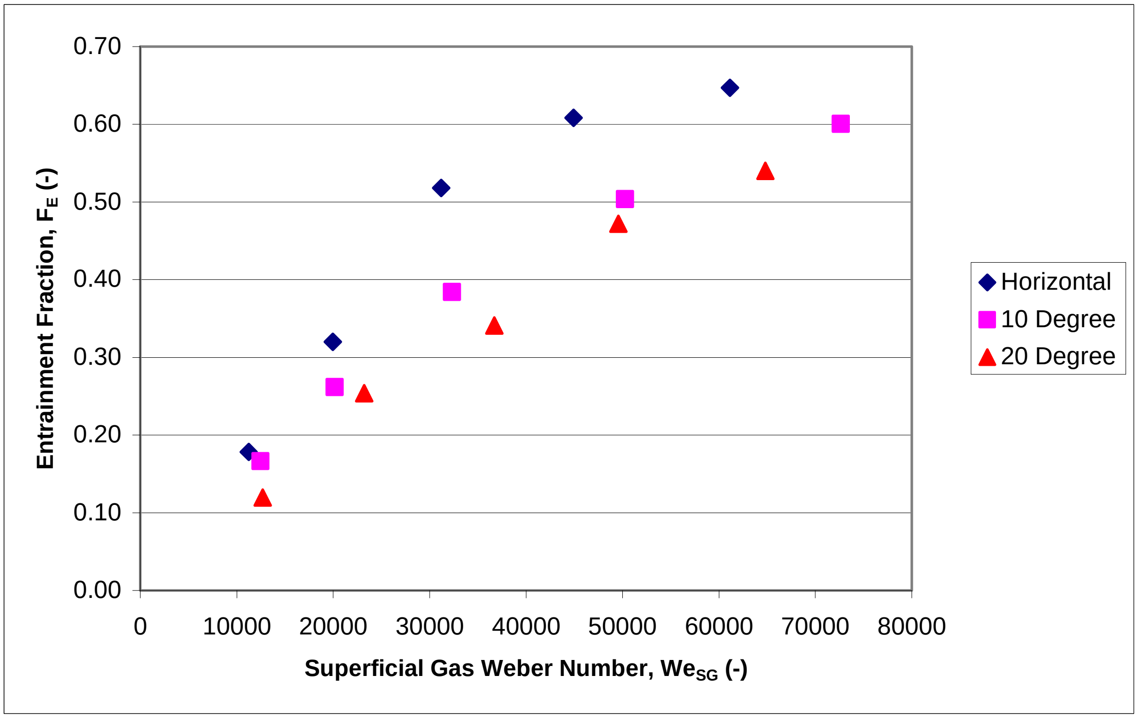


Figure 11. Entrainment Fraction Data at Re_{SL} = 230 for Various Inclination Angles.

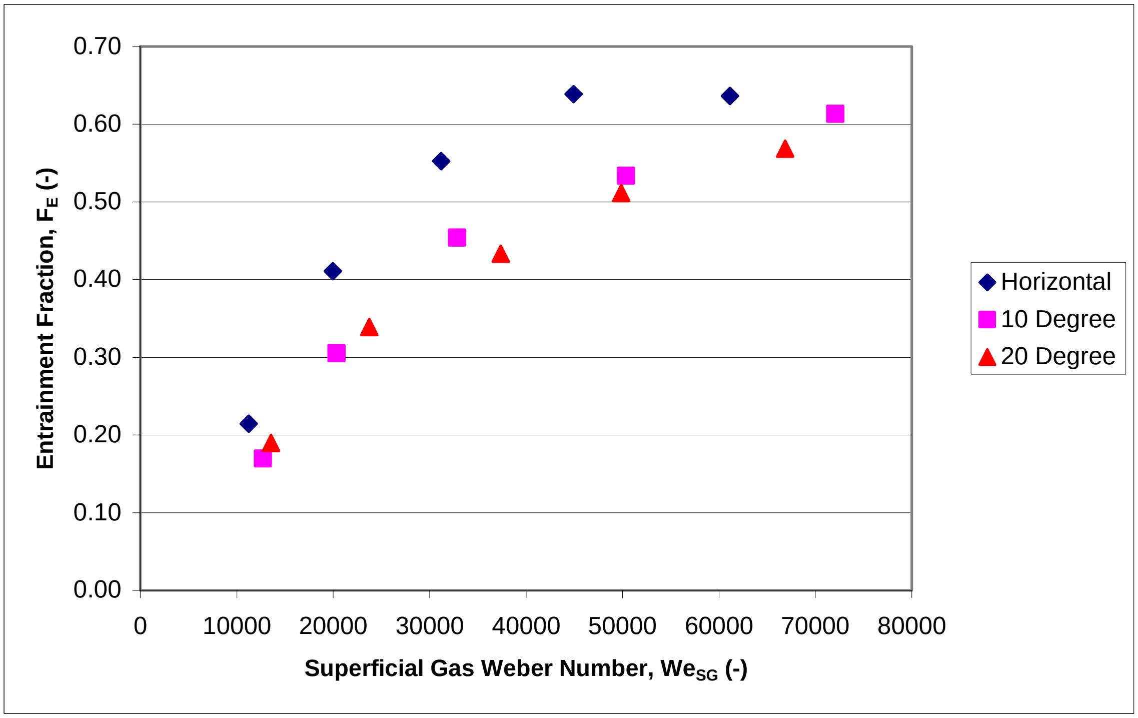


Figure 12. Entrainment Fraction Data at Re_{SL} = 380 for Various Inclination Angles.

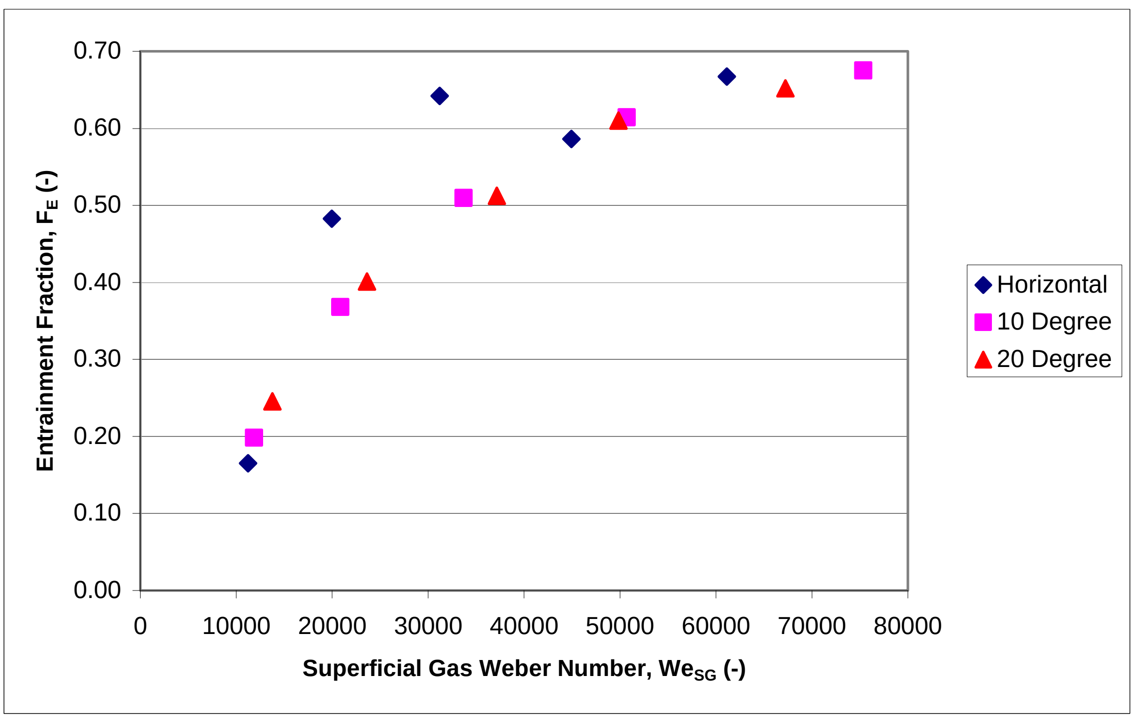


Figure 13. Entrainment Fraction Data at Re_{SL} = 760 for Various Inclination Angles.

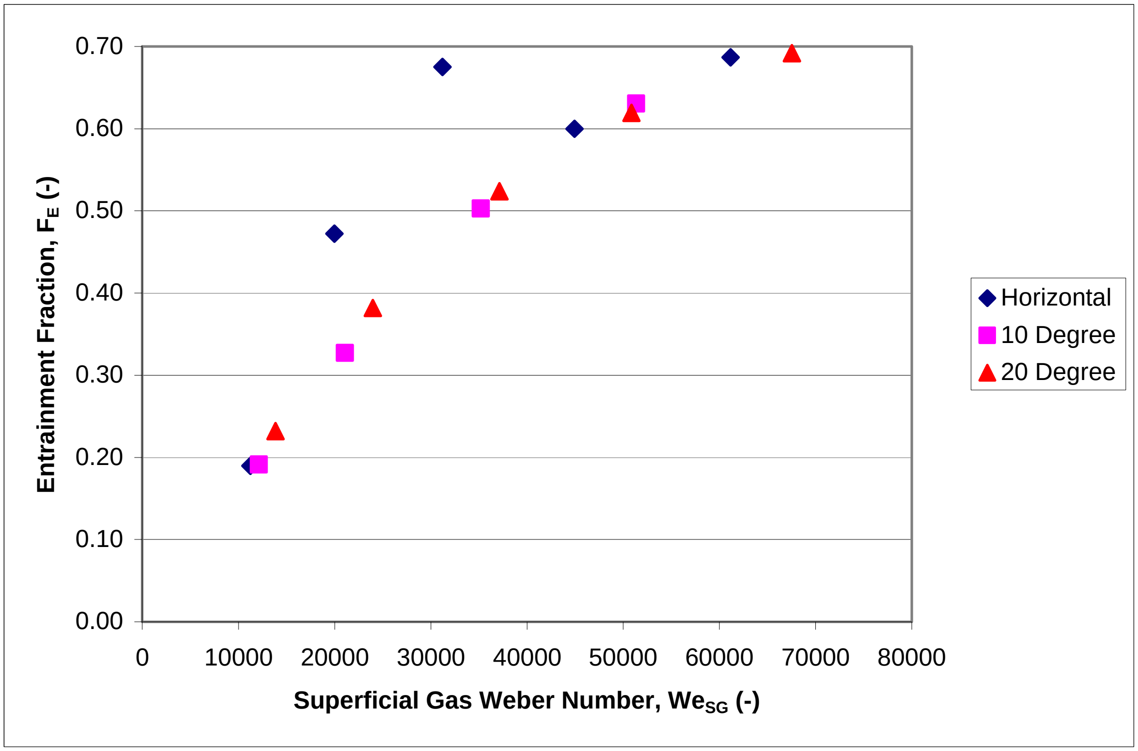


Figure 14. Entrainment Fraction Data at Re_{SL} = 1520 for Various Inclination Angles.

Upward Multiphase Flow in a Vertical Annulus

- Significance
  - Production Through Annulus
  - Liquid Loading Problem
- Objective
  - Significant Improvements in Multiphase Flow Modeling Since 1985
  - Development of an Improved Mechanistic Model for Vertical Annulus
- Past Studies
  - Caetano
    - Thorough Experimental and Modeling Study in 1985

Upward Multiphase Flow in a Vertical Annulus …

- Current Study
  - Developed a New Model Based on Unified Modeling Approach
  - New Model Outperforms the Original Unified Model

Fluid Flow Projects

Modeling of Gas-Liquid Flow in Upward Vertical Annuli

Tingting YU

Advisory Board Meeting, March 25, 2009

Outline

- Objectives
- Introduction
- Hydrodynamic Models for Individual Flow Patterns
- Flow Pattern Transition Models
- Performance Analysis
- Schedule

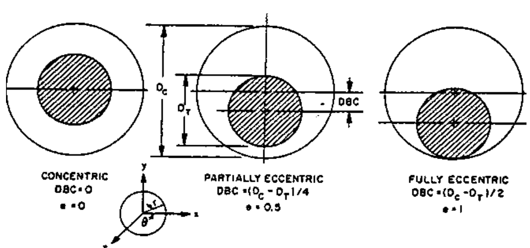
Objectives

- Theoretically Investigate Gas-Liquid Flow in Upward Vertical Concentric and Eccentric Annuli
- Develop a New Model
- Validate the Model with Experimental Data

Introduction

- Annulus Formed by Two Circular Pipes
- Two Geometrical Parameters

Annulus Configuration



$$K = \frac{d_T}{d_C}$$

$$e = \frac{2D_{BC}}{(d_C - d_T)}$$

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Reference Diameter

- Representative Diameter

$$d_r = \sqrt{d_C{}^2 - d_T^2}$$

- Hydraulic Diameter

$$d_H = d_C - d_T$$

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Background

- No Hydrodynamic Models for Vertical Annulus Flow Since Caetano (1986)
- Unified Model Predictions of Liquid Holdup and Pressure Gradient of Annulus Flow Not Satisfactory
- New Model Developed by Taking Annulus Configuration into Account

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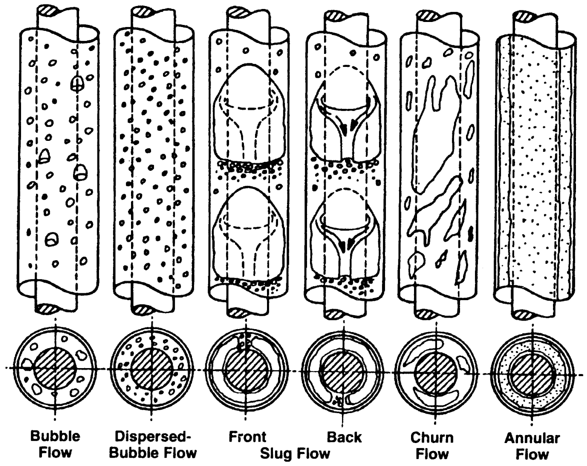
Flow Patterns

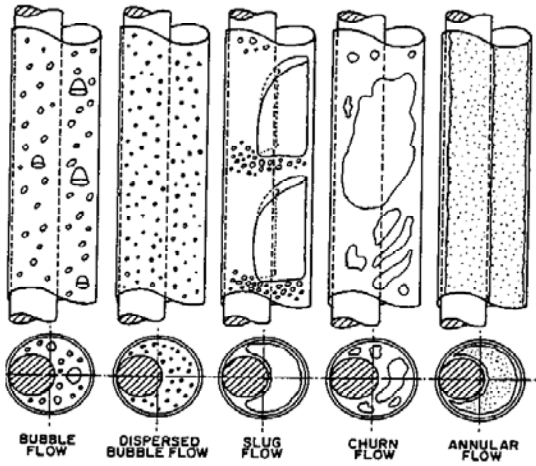
- Bubble Flow
  - Bubbly and Dispersed Bubble
- Intermittent Flow
  - Slug and Churn
- Separated Flow
  - Annular

Flow Patterns in Concentric Annulus



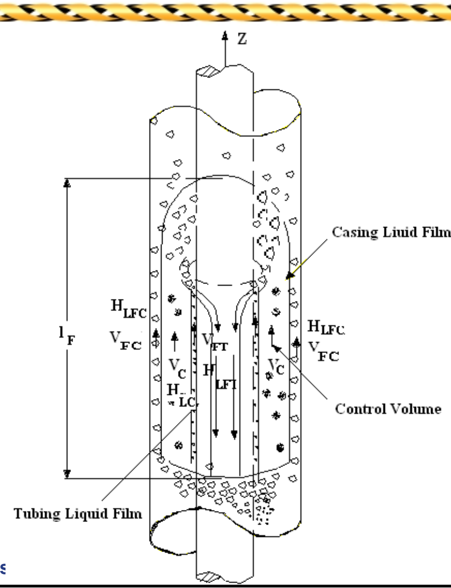
Caetano (1986)

Flow Patterns in Fully Eccentric Annulus



Caetano (1986)

Control Volume



Slug Flow Model

- Mass Conservation
  - Liquid in Liquid Film Zone

$$H_{LS}(v_T - v_S) = H_{LFC}(v_T - v_{FC}) + H_{LFT}(v_T - v_{FT})$$

  - Gas in Liquid Film Zone

$$(1 - H_{LS})(v_T - v_S) = (1 - H_{LFC} - H_{LFT})(v_T - v_C)$$

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Slug Flow Model…

- Continuity Equations
  - Liquid in Slug Unit

$$l_U v_{SL} = l_S H_{LS} v_S + l_F (H_{LFC} v_{FC} + H_{LFT} v_{FT})$$

  - Gas in Slug Unit

$$l_U v_{SG} = l_S (1 - H_{LS}) v_S + l_F (1 - H_{LFC} - H_{LFT}) v_C$$

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Slug Flow Model…

- **Momentum Equation for Liquid Film**
  - Casing Film

$$\frac{(P_2 - P_1)}{l_F} = \frac{\rho_L (v_{FC} - v_T)(v_{FC} - v_S)}{l_F} + \frac{\tau_{IC} S_{IC}}{H_{LFC} A} - \frac{\tau_{FC} S_{FC}}{H_{LFC} A} - \rho_L g \sin\theta$$

  - Tubing Film

$$\frac{(P_2 - P_1)}{l_F} = \frac{\rho_L (v_{FT} - v_T)(v_{FT} - v_S)}{l_F} + \frac{\tau_{IT} S_{IT}}{H_{LFT} A} - \frac{\tau_{FT} S_{FT}}{H_{LFT} A} - \rho_L g \sin\theta$$

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Slug Flow Model…

- **Momentum Equation for Gas Core**

$$\frac{(P_2 - P_1)}{l_F} = \frac{\rho_C (v_T - v_C)(v_S - v_C)}{l_F} + \frac{\tau_{IT} S_{IT} + \tau_{IC} S_{IC}}{(1 - H_{LFT} - H_{LFC}) A} - \rho_C g \sin\theta$$

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Slug Flow Model…

- **Combined Momentum Equation for Casing Liquid Film**

$$\frac{\rho_L(v_{FC}-v_T)(v_{FC}-v_S)-\rho_C(v_C-v_T)(v_C-v_S)}{l_F}-\frac{\tau_{FC}S_{FC}}{H_{LFC}A}$$
$$+\frac{\tau_{IT}S_{IT}}{(1-H_{LFC}-H_{LFT})A}+\tau_{IC}S_{IC}\left(\frac{1}{H_{LFC}A}+\frac{1}{(1-H_{LFC}-H_{LFT})A}\right)$$
$$+(\rho_C-\rho_L)g\sin\theta=0$$

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Slug Flow Model…

- **Combined Momentum Equation for Tubing Liquid Film**

$$\frac{\rho_L(v_{FT}-v_T)(v_{FT}-v_S)-\rho_C(v_C-v_T)(v_C-v_S)}{l_F}-\frac{\tau_{FT}S_{FT}}{H_{LFT}A}$$
$$+\frac{\tau_{IC}S_{IC}}{(1-H_{LFC}-H_{LFT})A}+\tau_{IT}S_{IT}\left(\frac{1}{H_{LFT}A}+\frac{1}{(1-H_{LFC}-H_{LFT})A}\right)$$
$$+(\rho_c-\rho_L)g\sin\theta=0.$$

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Annular Flow Model

- Momentum Equation for Casing Film

0

$$\frac{\tau_{FC}S_{FC}}{H_{LFC}A}$$

$$+\frac{\tau_{IT}S_{IT}}{(1-H_{LFC}-H_{LFT})A}+\tau_{IC}S_{IC}\left(\frac{1}{H_{LFC}A}+\frac{1}{(1-H_{LFC}-H_{LFT})A}\right)+(\rho_C-\rho_L)g\sin\theta=0$$

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Annular Flow Model …

- Momentum Equation for Tubing Film

0

$$\frac{\tau_{FT}S_{FT}}{H_{LFT}A}$$

$$+\frac{\tau_{IC}S_{IC}}{(1-H_{LFC}-H_{LFT})A}+\tau_{IT}S_{IT}\left(\frac{1}{H_{LFT}A}+\frac{1}{(1-H_{LFC}-H_{LFT})A}\right)+(\rho_c-\rho_L)g\sin\theta=0$$

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Churn Flow Model

- Combined Momentum Equation for Liquid Film and Gas Pocket

$$\frac{\rho_L(v_T-v_F)(v_S-v_F)-\rho_C(v_T-v_C)(v_S-v_C)}{l_F}+\frac{\tau_C S_C}{(1-H_{LF})A}$$

$$-\frac{\tau_F S_F}{H_{LF}A}+\tau_I S_I\left(\frac{1}{H_{LF}A}+\frac{1}{(1-H_{LF})A}\right)-(\rho_L-\rho_C)g\sin\theta=0$$

- Slug Length

$$l_s = 4.0d_r$$

Bubble Flow Model

- Bubbly Flow Model

$$v_O = 1.53\left[\frac{(\rho_L-\rho_G)g\sigma}{\rho_L{}^2}\right]^{1/4} H_L{}^n \qquad v_S = \frac{v_{SG}}{1-H_L}-\frac{v_{SL}}{H_L}$$

- Dispersed Bubble Model
  - Homogeneous Flow

Flow Pattern Transitions

- Transition to Annular Flow
  - $R_{FU} \rightarrow 1$ $$R_{FU} = \frac{l_F}{l_U}$$
- Transition to Dispersed Bubble Flow
  - Zhang et al. (2003) Model (v_{SG}>0.1m/s) and Barnea (1986) Model (v_{SG}<0.1m/s)

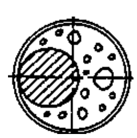
Flow Pattern Transitions…

- Transition to Bubble Flow
  - Concentric Annulus

$$v_{SG} = 0.25 v_{SL} + 0.306 \left[\frac{(\rho_L - \rho_G) g \sigma}{\rho_L^{\;2}} \right]^{1/4}$$

  - Fully Eccentric Annulus

$$v_{SG} = 0.18 v_{SL} + 0.306 \left[\frac{(\rho_L - \rho_G) g \sigma}{\rho_L^{\;2}} \right]^{1/4}$$



Flow Pattern Transitions…

- Transition from Slug Flow to Churn Flow
  - Kaya (1998) Model

$$v_{SG} = 2.76(1.2v_{SL} + v_O)$$

$$v_O = (0.35\sin\theta + 0.54\cos\theta)\left(\frac{g(\rho_L - \rho_g)d}{\rho_L}\right)^{1/2}$$

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Closure Relationships

- Film Liquid Holdup Ratio
- Slug Translational Velocity
- Wall Friction Factor
- Interfacial Friction Factor
- Liquid Entrainment Fraction in Gas Core
- Slug Liquid Holdup
- Slug Length

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Film Liquid Holdup Ratio

- Caetano's (1986) Liquid Film Holdup Equations

$$H_{LFT} = \frac{4\delta_T}{d_C} K \frac{\left(1+\frac{\delta_T}{d_T}\right)}{(1-K^2)} \qquad H_{LFC} = \frac{4\delta_C}{d_C} \frac{\left(1+\frac{\delta_C}{d_C}\right)}{(1-K^2)}$$

$$\frac{H_{LFT}}{H_{LFC}} = \frac{\delta_T}{\delta_C} K \qquad \frac{\delta_T}{\delta_C} = \frac{W_T{}'}{(2\pi - W_T{}')K}$$

$$W_T{}' = \frac{1}{(1-K^2)}\left(2 * \sin^{-1}(K) + 2K\sqrt{1-K^2} - K^2\pi\right)$$

Slug Translational Velocity

- Hasan and Kabir (1900)
  - Drift velocity

$$v_D = 0.345(1+0.1K)\sqrt{gd_c(\rho_L - \rho_G)/\rho_L}$$

- Nicklin (1962) $v_T = C_S v_S + v_D$
  - Laminar Flow $C_S = 2.0$
  - Turbulent Flow $C_S = 1.3$
  - Transition Region $C_S = 2.0 - 0.7(Re - 2000)/2000$

Wall Friction Factor

- **Friction Factor for Shear Stress at Wall**

$$f = C\,Re^{-n}$$

$C = 16,\ n = 1$ **Laminar**

$C = 0.046,\ n = 0.2$ **Turbulent**

- **Caetano's Friction Factor**

$$f_{CA} = \frac{16}{Re}\frac{(1-K)^2}{\left[\dfrac{1-K^4}{1-K^2} - \dfrac{1-K^2}{ln(1/K)}\right]} \quad \textbf{Laminar}$$

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Wall Friction Factor…

- **Caetano's Friction Factor**

$$\frac{1}{\left\{ f_{CA}\left(\dfrac{F_P}{F_{CA}}\right)^{0.45 exp\left[-(Re-3000)/10^6\right]}\right\}^{1/2}} = 4.0\,log\left\{ Re\left[f_{CA}\left(\dfrac{F_P}{F_{CA}}\right)^{0.45 exp\left[-(Re-3000)/10^6\right]}\right]^{1/2}\right\} - 0.40 \quad \textbf{Turbulent}$$

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Interfacial Friction Factor

- Ambrosini et al. (1991) Correlation

$$f_I = f_G\left(1+13.8\,\mathrm{We}_G^{0.2}\mathrm{Re}_G^{-0.6}(\mathrm{h}_F^{+}-200\sqrt{\rho_G/\rho_L})\right)$$

$$We_G = \frac{\rho_G v_C^2 d}{\sigma} \qquad \mathrm{Re}_G = \frac{\rho_G v_C d}{\mu_G}$$

$$h_F^{+} = \frac{\rho_G \delta_C v_C^{*}}{\mu_G} \qquad v_C^{*} = \sqrt{\tau_I / \rho_G}$$

$$f_G = 0.046\,\mathrm{Re}_G^{-0.2}$$

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Liquid Entrainment Fraction

- Wallis et al. (1969) Correlation

$$f_E = 1-exp\left[-0.125(\phi-1.5)\right]$$

$$\phi = 10^4\,\frac{v_{SG}\mu_G}{\sigma}\left(\frac{\rho_G}{\rho_L}\right)^{0.5}$$

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Slug Liquid Holdup

Zhang et al. (2003) Model

$$H_{LS} = \frac{1}{1 + \dfrac{T_{sm}}{3.16\left[(\rho_L - \rho_G) g\sigma\right]^{1/2}}}$$

$$T_{sm} = \frac{1}{C_e}\left(\begin{array}{l} \dfrac{f_S}{2}\rho_S v_S^2 + \\ \dfrac{d_C}{4}\dfrac{\rho_L (H_{LFC} + H_{LFT})(v_T - v_{FC})(v_S - v_{FC})}{l_S} \\ + \dfrac{d_C}{4}\dfrac{\rho_C (1 - H_{LFC} - H_{LFT})(v_T - v_C)(v_s - v_C)}{l_S} \end{array} \right)$$

$$C_e = \frac{2.5 - |\sin(\theta)|}{2}$$

Slug Length

- Taitel et al. (1980) and Barnea and Brauner (1985)

$$l_S = (32.0 \cos^2\theta + 16.0 \sin^2\theta) d_r$$

Experimental Data

- **Experimental Information**
  - Caetano (1986) Test Facility- 16-m (52.493-ft) Long with 76.2-mm (3-inch) I.D. Casing and 42.2 (1.66-inch) O.D. Tubing
  - Experimental Fluids-Air, Water and Kerosene
  - Annulus Configuration: Concentric and Fully Eccentric Annulus

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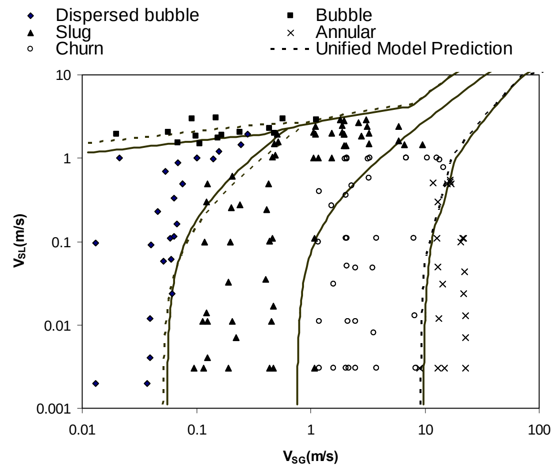
Flow Pattern Map



Air and Water in Concentric Annulus

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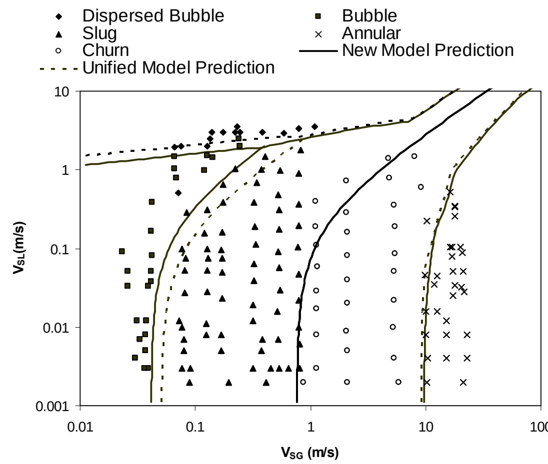
Flow Pattern Map…



Air and Water in Fully Eccentric Annulus

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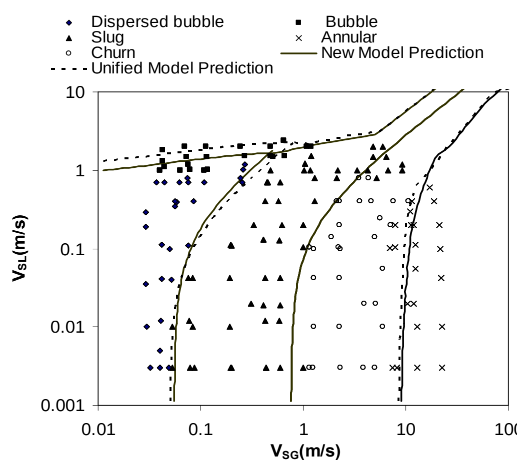
Flow Pattern Map…



Air and Kerosene in Concentric Annulus

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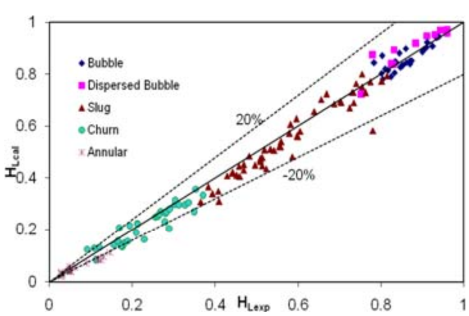
Flow Pattern Transition Model Performance

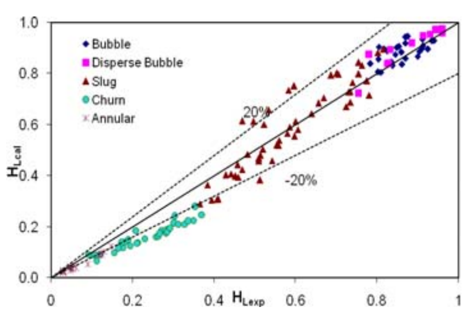
- Bubbly/Slug Flow Pattern Transition Work Well for All Flow Conditions
- Dispersed Bubble Flow Pattern Transition Perform Well Except for Air and Kerosene
- Slug/Churn Flow Pattern Transition Model Does Not Perform Well at Higher Liquid Flow Rate
- Annular Flow Pattern Transition Model Predict Higher Transition Velocity

Liquid Holdup Prediction Concentric Annulus

Air and Water



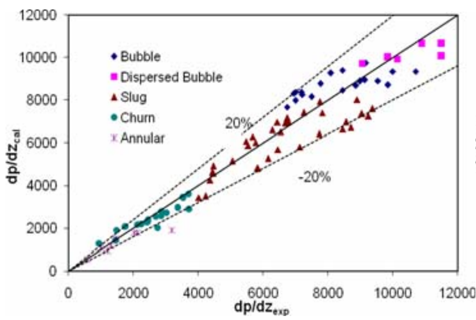
Present Model Performance

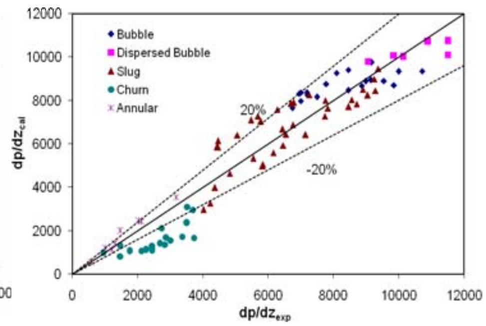


Unified Model Performance

Pressure Gradient Prediction
Concentric Annulus

Air and Water





Present Model Performance

Unified Model Performance

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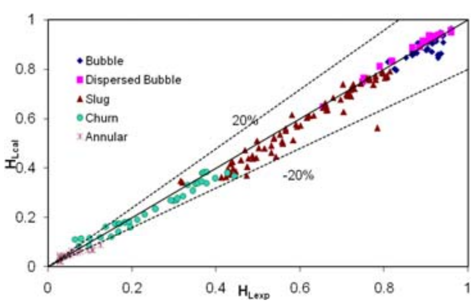
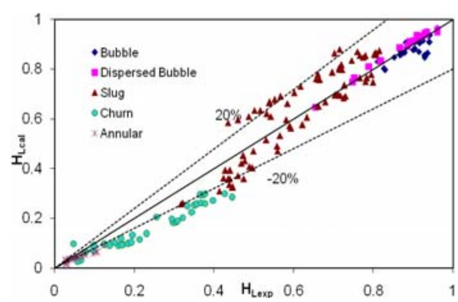
Liquid Holdup Prediction
Fully Eccentric Annulus

Air and Water





Present Model Performance

Unified Model Performance

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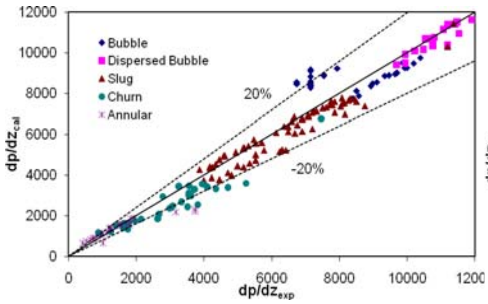
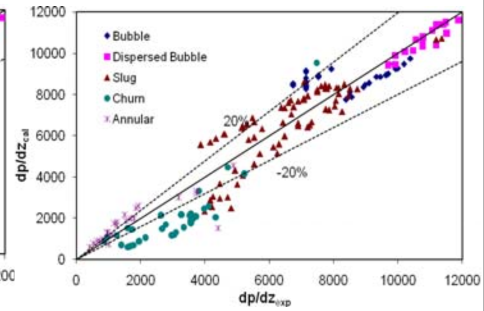
Pressure Gradient Prediction
Fully Eccentric Annulus

Air and Water

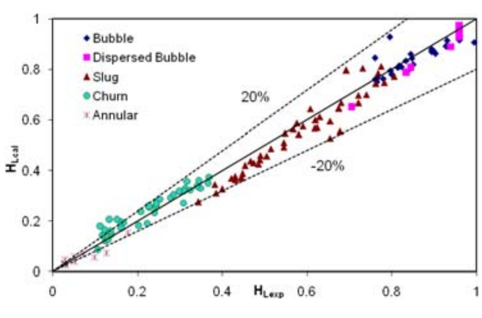


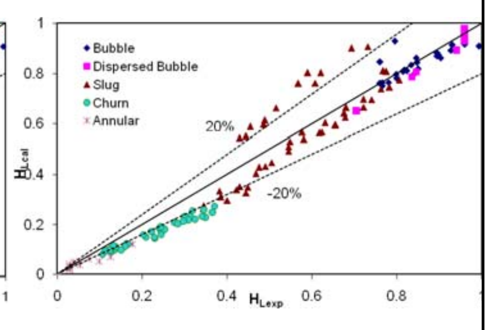


Present Model Performance **Unified Model Performance**

Liquid Holdup Prediction
Concentric Annulus

Air and Kerosene

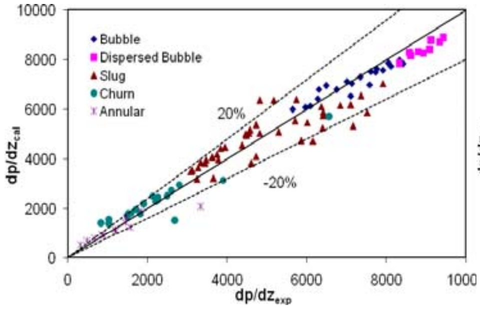




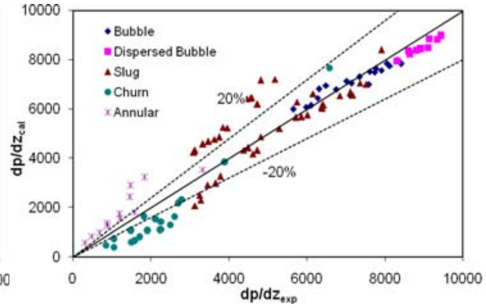
Present Model Performance **Unified Model Performance**

Pressure Gradient Prediction Concentric Annulus

Air and Kerosene



Present Model Performance



Unified Model Performance

Evaluation Criteria

- Absolute Average Percentage Error

$$E_2 = \left[\frac{1}{N} \sum_{j=1}^{N} \left| e_{rj} \right| \right] \times 100$$

- Standard Deviation

$$E_3 = \sum_{j=1}^{N} \sqrt{\frac{(e_{rj} - E_2)^2}{N-1}}$$

- wc, we, kc Refer to Air and Water in Concentric Annulus, Air and Water in Fully Eccentric Annulus and Air and Kerosene in Concentric Annulus

Slug Flow Model Performance

Slug Flow	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_{Lnew}	6.7%	4.0%	8.6%	6.7%	3.8%	5.8%
$H_{Lunified}$	12.6%	12.5%	19.0%	10.9%	8.8%	9.7%
$-dp/dz_{new}$	8.3%	5.8%	11.4%	7.8%	3.8%	5.3%
$-dp/dz_{unified}$	14.7%	15.4%	23.1%	12.0%	11.9%	17.0%

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Annular Flow Model Performance

Annular Flow	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_{Lnew}	19.7%	18.1%	24.1%	10.3%	10.9%	15.3%
$H_{Lunified}$	28.6%	24.2%	30.7%	12.2%	14.0%	16.3%
$-dp/dz_{new}$	14.5%	19.4%	14.0%	11.8%	15.2%	14.1%
$-dp/dz_{unified}$	14.8%	29.3%	48.5%	11.1%	16.4%	26.9%

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Churn Flow Model Performance

Churn Flow	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_{Lnew}	8.6%	11.2%	9.8%	7.9%	5.9%	7.2%
$H_{Lunified}$	26.9%	32.1%	26.6%	9.3%	10.3%	6.1%
$-dp/dz_{new}$	9.8%	16.6%	13.1%	7.8%	9.0%	11.5%
$-dp/dz_{unified}$	34.5%	39.8%	23.1%	34.2%	16.6%	19.6%

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Bubble Flow Model Performance

Bubble Flow	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_L	3.2%	3.5%	3.0%	2.9%	2.9%	3.7%
$-dp/dz$	9.9%	12.2%	2.9%	6.4%	6.0%	2.7%

Dispersed Bubble Flow	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_L	2.7%	1.6%	4.6%	2.0%	0.9%	2.9%
$-dp/dz$	4.9%	2.5%	2.9%	5.3%	1.8%	2.2%

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Concluding Remarks

- New Model Performs Well with Improvement Over Modified Model
- Big Error Points Existed Due to the Inaccuracy of Flow Pattern Prediction
- Annular Flow Model Still Needs to be Evaluated and Improved

Project Schedule

Model Development	Completed
Model Verification	Completed
Final Report	May 2009

Questions & Comments

Questions ?

Fluid Flow Projects

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Modeling of Gas-Liquid Flow in an Upward Vertical Annulus

Tingting YU

PROJECTED COMPLETION DATES:

Objectives

The objectives of this study are:

- Theoretically investigate gas-liquid two-phase flows in upward vertical concentric and eccentric annuli,
- Analyze data from a previous experimental study (Caetano, 1986) and develop a new model for gas-liquid two-phase flows in annuli.

Introduction

An annulus is formed by a pipe being located inside a larger pipe as shown in Fig. 1. Fluid flows through the area bounded by the outer pipe casing inner wall and the inner pipe tubing outer wall. There are two important parameters to identify this configuration: annulus pipe diameter ratio and the degree of eccentricity.

The pipe diameter ratio is given by:

$$K = \frac{d_T}{d_C} . \qquad (1)$$

Where, d_T is the outer diameter of the tubing and d_C is the inner diameter of the casing. The degree of eccentricity accounts for the displacement of the inner pipe center from the outer pipe center and is expressed by

$$e = \frac{2DBC}{(d_C - d_T)} . \qquad (2)$$

Where, DBC is the distance between the two pipe centers. In the petroleum industry, multiphase flow in wells normally occurs in a tubing string. However, many oil wells with high production rates produce through the casing-tubing annulus. This trend can be dictated by economics, multiple completions and regulated production rates. Although the number of these wells is small compared with all producing wells, these "casing flow" wells still account for a significant part of the world oil production.

Many applications of casing flow in the oil industry are also found in various types of artificial lift. In sucker rod pumping wells, a rod string is installed inside the tubing string to connect the prime mover unit on the surface to the pump at the bottom of the well. The fluids are pumped upward through the tubing-rod string annulus.

Another application of flow through an annulus is found in gas well production. In order to remove or "unload" undesirable liquids accumulated at the bottom of these wells, a siphon tube is often installed inside the tubing string. The normal permanency of the siphon tube in the tubing string requires the fluids to flow upward through the tubing string-siphon tube annulus.

Most researchers have treated the annulus based on the hydraulic diameter concept. The hydraulic diameter is four times the area for flow divided by the wetted perimeter. Then, for an annulus configuration, the hydraulic diameter is

$$d_H = d_C - d_T . \qquad (3)$$

However, the hydraulic diameter is not always the most representative characteristic dimension for flow in an annulus. Omurlu and Evren (2007) introduced a "representative diameter" d_r for a fully eccentric annulus. The representative diameter is defined as

$$d_r = \sqrt{d_C^{\,2} - d_T^2} \,. \tag{4}$$

Zhang *et al.* (2003) unified model was tested against Caetano's (1986) experimental data using hydraulic diameter and representative diameter, respectively. The results were not satisfactory due to errors in flow pattern prediction and large errors in slug and annular flows.

Considering the limitation of the previous models for annulus flow, new flow pattern transition models and hydrodynamic models are developed based on annulus configuration. The model predictions are compared with Caetano's (1986) experimental data.

Model Development

Hydrodynamic Model

The new hydrodynamic models are developed following the approach of Zhang *et al.* (2003) unified model.

Modeling of Slug Flow

Mass Balance Equations in Film Zone

The hydrodynamic model developed for slug flow in annuli considers two liquid films. The entire liquid film zone of a slug unit is taken as the control volume (Fig. 2).

Assuming incompressible flow and no liquid entrainment in the gas core, the mass balances can be expressed in terms of volumetric flow rates based on a Lagrangian coordinate system. The input liquid mass flow rate at the lower boundary in the film zone can be written as,

$$\rho_L A H_{LFC}(v_T - v_{FC}) + \rho_L A H_{LFT}(v_T - v_{FT}) \,,$$

and the output liquid mass flow rate at the upper boundary in the film zone is

$$\rho_L A H_{LS}(v_T - v_S) \,.$$

Where H_{LFC} and H_{LFT} are the liquid holdup in the casing film and tubing film, respectively, H_{LS} is the liquid holdup in the slug body, v_{FC} and v_{FT} are the velocity of casing liquid film and tubing liquid film, respectively. The fluid velocity in the slug body is expressed as,

$$v_S = v_{SL} + v_{SG} \,. \tag{5}$$

For fully developed slug flow, the input mass flow rate at the lower boundary of the film zone is equal the output mass flow rate at the upper boundary of the film zone. Thus, the mass balances for liquid and gas phases can, respectively, be written as,

$$H_{LS}(v_T - v_S) = H_{LFC}(v_T - v_{FC}) + H_{LFT}(v_T - v_{FT}) \,. \tag{6}$$

$$(1 - H_{LS})(v_T - v_S) = (1 - H_{LFC} - H_{LFT})(v_T - v_C) \,. \tag{7}$$

Where, v_C is the gas core velocity in the film region.

Overall Mass Balances

Considering the gas and liquid flow along the slug unit and the incompressibility of gas and liquid, the mass balance equations for liquid and gas can be written, respectively.

$$l_U v_{SL} = l_S H_{LS} v_S + l_F (H_{LFC} v_{FC} + H_{LFT} v_{FT}) \tag{8}$$

$$l_U v_{SG} = l_S (1 - H_{LS}) v_S + l_F (1 - H_{LFC} - H_{LFT}) v_C \,. \tag{9}$$

Where, l_U is the length of slug unit, l_S is the length of liquid slug, and l_F is the length of liquid film.

The slug unit length is the sum of slug length and film length,

$$l_U = l_S + l_F \,. \tag{10}$$

Momentum Equations

In Zhang *et al.* (2003) unified model, the forces acting on the left and right boundaries of the liquid film include momentum exchange between slug body and liquid film, frictional force acting at the wall, static pressure difference between left and right boundaries, frictional forces acting at the interface and gravitational force. In this study, the same forces are considered in deriving the momentum equations for annulus flow except that two liquid films are considered in the derivation process.

The momentum equations for casing liquid film and tubing liquid film are derived separately. The liquid flowing from slug body into the Taylor bubble region is considered to split into casing liquid film and tubing liquid film two parts. Similarly, the liquid films flowing from the Taylor bubble region to the slug body are also assumed to split into two parts in the slug body, which could not be visually seen in reality.

For casing liquid film, the momentum input at the lower boundary is

$\rho_L A H_{LFC}(v_{FC} - v_T) v_{FC}$.

The momentum output at the upper boundary is

$\rho_L A H_{LSC}(v_S - v_T) v_S$.

Where, H_{LSC} is the liquid holdup corresponding to the casing liquid film part.

For fully developed slug flow, the input mass flow rate equals to the output mass flow rate and both can be expressed as,

$\rho_L A H_{LFC}(v_{FC} - v_T)$.

Therefore, the momentum output can be expressed as

$\rho_L A H_{LFC}(v_{FC} - v_T) v_S$.

Thus the momentum exchange between casing liquid film and gas pocket can be obtained as,

$\rho_L A H_{LFC}(v_{FC} - v_T)(v_{FC} - v_S)$.

The frictional force acting on casing film at the wall is

$-\tau_{FC} S_{FC} l_F$.

The frictional force acting on the interface between casing film and gas pocket is

$-\tau_{IC} S_{IC} l_F$.

All forces acting on the casing liquid film should be in balance and hence the momentum equation for the casing film can be written as

$$\begin{aligned} &H_{LFC} A P_1 - H_{LFC} A P_2 + \\ &\rho_L A H_{LFC}(v_{FC} - v_T)(v_{FC} - v_S) \\ &+ \tau_{IC} S_{IC} l_F - \tau_{FC} S_{FC} l_F - \rho_L H_{LFC} l_F A g \sin\theta = 0 \end{aligned} \quad . \quad (11)$$

Similarly, the momentum equation for the tubing film can be written as

$$\begin{aligned} &H_{LFT} A P_1 - H_{LFT} A P_2 + \\ &\rho_L A H_{LFT}(v_{TF} - v_T)(v_{FT} - v_S) \\ &+ \tau_{TI} S_{TI} l_F - \tau_{TF} S_{TF} l_F - \\ &\rho_L H_{LFT} l_F A g \sin\theta = 0 \end{aligned} \quad . \quad (12)$$

Rearrange Eqs. (11) and (12) to get the momentum equations for the casing film and tubing film, respectively,

$$\begin{aligned} &\frac{(P_2 - P_1)}{l_F} = \frac{\rho_L (v_{FC} - v_T)(v_{FC} - v_S)}{l_F} \\ &+ \frac{\tau_{IC} S_{IC}}{H_{LFC} A} - \frac{\tau_{FC} S_{FC}}{H_{LFC} A} - \rho_L g \sin\theta \end{aligned} \quad . \quad (13)$$

$$\begin{aligned} &\frac{(P_2 - P_1)}{l_F} = \frac{\rho_L (v_{FT} - v_T)(v_{FT} - v_S)}{l_F} \\ &+ \frac{\tau_{IT} S_{IT}}{H_{LFT} A} - \frac{\tau_{FT} S_{FT}}{H_{LFT} A} - \rho_L g \sin\theta \end{aligned} \quad . \quad (14)$$

Similarly, the momentum equation for the gas pocket can be written as:

$$\begin{aligned} &\frac{(P_2 - P_1)}{l_F} = \frac{\rho_C (v_T - v_C)(v_s - v_C)}{l_F} \\ &+ \frac{\tau_{IT} S_{IT} + \tau_{IC} S_{IC}}{(1 - H_{LFT} - H_{LFC}) A} - \rho_C g \sin\theta = 0. \end{aligned} \quad (15)$$

Eqs. (13) and (15), and Eqs. (14) and (15) yield the combined equations for the casing and the tubing, respectively.

$$\begin{aligned} &\frac{\rho_L (v_{FC} - v_T)(v_{FC} - v_S) - \rho_C (v_C - v_T)(v_C - v_S)}{l_F} \\ &- \frac{\tau_{FC} S_{FC}}{H_{LFC} A} + \frac{\tau_{IT} S_{IT}}{(1 - H_{LFC} - H_{LFT}) A} \\ &+ \tau_{IC} S_{IC} \left(\frac{1}{H_{LFC} A} + \frac{1}{(1 - H_{LFC} - H_{LFT}) A}\right) \\ &+ (\rho_C - \rho_L) g \sin\theta = 0 \end{aligned} \quad . \quad (16)$$

$$\begin{aligned} &\frac{\rho_L (v_{FT} - v_T)(v_{FT} - v_S) - \rho_C (v_C - v_T)(v_C - v_S)}{l_F} \\ &- \frac{\tau_{FT} S_{FT}}{H_{LFT} A} + \frac{\tau_{IC} S_{IC}}{(1 - H_{LFC} - H_{LFT}) A} \\ &+ \tau_{IT} S_{IT} \left(\frac{1}{H_{LFT} A} + \frac{1}{(1 - H_{LFC} - H_{LFT}) A}\right) \\ &+ (\rho_c - \rho_L) g \sin\theta = 0 \end{aligned} \quad . \quad (17)$$

The unknown variables in the slug flow model include $v_{FC}, v_{FT}, H_{LFC}, H_{LFT}, H_{LF}, v_C, l_F$. H_{LF} and v_C can be solved from Eqs. (7) and (9) using an iterative process on l_F .

$$H_{LF} = \frac{l_F v_T H_{LS} + l_U (v_{SL} - v_S H_{LS})}{v_T l_F} . \quad (18)$$

$$v_C = \frac{v_T(1-H_{LF})-(1-H_{LS})(v_T-v_S)}{1-H_{LF}}. \quad (19)$$

Holdups for the two liquid films, H_{LFC} and H_{LFT}, are first calculated from the closure relationship of the casing-tubing liquid film holdup ratio. v_{FC} is iteratively calculated. Eq. (6) is used to calculate the value of v_{FT}, and the two liquid film velocities v_{FC} and v_{FT} are checked through Eq. (16). The new value of l_F is then calculated using Eq. (17) and checked with the old value of l_F. Iteration process is repeated until the accurate value of l_F is obtained.

Modeling of Annular flow

For annular flow in annulus, the momentum equations for casing and tubing flows can be obtained by removing momentum exchange terms from Eqs. (16) and (17),

$$\frac{\tau_{IC}S_{IC}}{(1-H_{LFC}-H_{LFT})A} + \tau_{IT}S_{IT}\left(\frac{1}{H_{LFT}A}+\frac{1}{(1-H_{LFC}-H_{LFT})A}\right). - \frac{\tau_{FT}S_{FT}}{H_{LFT}A}+(\rho_C-\rho_L)g\sin\theta=0 \quad (20)$$

$$\frac{\tau_{IT}S_{IT}}{(1-H_{LFC}-H_{LFT})A}+ \tau_{IC}S_{IC}\left(\frac{1}{H_{LFC}A}+\frac{1}{(1-H_{LFC}-H_{LFT})A}\right). - \frac{\tau_{FC}S_{FC}}{H_{LFC}A}+(\rho_C-\rho_L)g\sin\theta=0 \quad (21)$$

The relationships between superficial velocities and the local fluid velocities are

$$v_{SL}=H_{LFC}v_{FC}+H_{LFT}v_{FT}+H_{LC}v_C, \quad (22)$$

$$v_{SG}=(1-H_{LFC}-H_{LFT}-H_{LC})v_C. \quad (23)$$

Where H_{LC} is the liquid holdup in the gas core.

Liquid entrainment fraction in the gas core is defined as

$$F_E=\frac{H_{LC}v_C}{v_{SL}}. \quad (24)$$

The liquid entrainment is considered in the annular flow, and the gas core includes liquid and gas phases. Therefore, the continuity equation in the gas core can be expressed as,

$$v_C(1-H_{LF})=v_{SG}+F_Ev_{SL}. \quad (25)$$

The unknown variables in the annular flow model include $v_{FC}, v_{FT}, H_{LFC}, H_{LFT}, v_C, and\ H_{LC}$. First, $v_C\ and\ H_{LC}$ are calculated solving Eqs. (24) and (25) through an iterative process on H_{LF}. Two liquid films holdup H_{LFC} and H_{LFT} can be calculated from the closure relationships. v_{FC} is calculated through an iteration process. Eq. (22) is used to calculate the value of v_{FT} and the two liquid film velocities v_{FC} and v_{FT} are checked through Eq. (20). The new value of H_{LF} is then calculated using Eq. (21) and checked with the old value of H_{LF}. Iteration process is repeated until the accurate value of H_{LF} is obtained.

Modeling of Churn Flow

Annulus churn flow is similar to annulus slug flow, but it is more chaotic, frothy and disordered. The liquid bridging the pipe is also much shorter and frothy compared to slug flow. This is because of the higher gas void fraction in liquid slug which breaks the continuity of the liquid in the liquid slug between successive Taylor bubbles. As this happens, the slug collapses, falls back and merges with the following slug. The bullet-shaped Taylor bubble is then distorted and churn flow occurs. Churn flow is similar to the churn flow in pipes.

There is no available mechanistic model to predict hydrodynamic behaviors of churn flow in pipes. Most researchers apply slug flow model in churn flow without any modifications.

Caetano (1986) described the flow characteristics of churn flow, but no relevant churn flow models were given. Kelessidis (1988) tried to predict the slug/churn flow pattern transition in concentric and eccentric annuli based on Taitel's (1980) slug/churn flow pattern transition in pipes. However, no hydrodynamic model for annulus churn flow has been developed.

In this study, a hydrodynamic model for annulus churn flow is developed based on the slug flow model in Zhang *et al.* (2003) unified model with necessary modifications. Only one liquid film is considered by using representative diameter. Different slug liquid holdup equations and slug length are used in the churn flow model.

The combined momentum equation for liquid film and gas pocket is given by

$$\frac{\rho_L(v_F - v_T)(v_F - v_S) - \rho_C(v_C - v_T)(v_C - v_S)}{l_F}$$
$$-\frac{\tau_F S_F}{H_{LF}A} + \frac{\tau_C S_C}{(1-H_{LF})A} + \tag{26}$$
$$\tau_I S_I\left(\frac{1}{H_{LF}A} + \frac{1}{(1-H_{LF})A}\right) - (\rho_L - \rho_C)g\sin\theta = 0$$

The Gregory *et al.* (1978) slug liquid holdup correlation is used in the churn flow model,

$$H_{LS} = \frac{1}{1+\left(\frac{v_S}{8.66}\right)^{1.39}} . \tag{27}$$

Taitel *et al.* (1980) treated churn flow as the entry region of slug flow. Churn flow was considered to have short liquid slugs and Taylor bubbles. As the flow proceeds upward, larger liquid slugs develop as a result of the fallback of unstable slugs overtaken by the Taylor bubble and merge with the trailing slugs. At the same time, Taylor bubble also doubles. Accordingly, the slug length for churn flow in upward annulus is set as

$$l_s = 4.0d . \tag{28}$$

Modeling of Bubble Flow

Bubble flow in this model includes dispersed bubble flow and bubbly flow. For dispersed bubble flow, the gas and liquid phases are assumed to be homogeneously mixed. Liquid holdup and pressure gradient are calculated based on this assumption. For bubbly flow, slippage between liquid and gas phase is considered, and Taitel *et al.* (1980) model (i.e., transition from bubble to slug flow) constitutes a rudimentary model for bubble flow.

$$v_{SL} = v_{SG}\frac{1-\alpha}{\alpha} - (1-\alpha)v_{o\infty} . \tag{29}$$

$$v_{o\infty} = 1.53\left[\frac{g(\rho_L - \rho_G)\sigma}{\rho_L^{\ 2}}\right]^{0.25} . \tag{30}$$

where a is the gas void fraction.

Flow Pattern Transition Models

A new flow pattern transition model is developed. The model is based on Zhang *et al.* (2003) approach with addition of a slug/churn flow pattern transition model.

Transition to Annular flow

The transition to annular flow occurs when the film length becomes infinitely long, which makes the momentum exchange term in the momentum equation zero. Given the superficial liquid velocity, v_{SL} , superficial gas velocity, v_{SG} can be calculated through an iteration process. Guessing v_{SG} and casing film velocity, v_{FC}, and based on the known variables, film liquid holdup H_{LF} , and gas core velocity, v_C , can be calculated. These parameters can be used to solve combined momentum equations for annular flow to calculate new casing and tubing liquid film velocity. The two liquid film velocities are used to calculate the new gas core velocity and the new gas superficial velocity.

$$H_{LF} = \frac{(H_{LS}(v_T - v_S) + v_{SL})(v_{SG} + v_{SL}F_E) - v_T v_{SL} F_E}{v_T v_{SG}} . \tag{31}$$

$$v_C = \frac{v_{SL} - v_{FC}H_{LFC} - v_{FT}H_{LFT}}{H_{LC}} . \tag{32}$$

$$v_{SG} = v_C(1.0 - H_{LF}) - v_{SL}F_E . \tag{33}$$

Transition to Dispersed Bubble Flow

At high gas flow rate, Zhang *et al.* (2003) mechanistic model for slug liquid holdup and transition between slug flow and dispersed bubble flow is used as the transition criterion. In this model, dispersed bubble flow was assumed to accommodate the maximum gas holdup at the transition boundary, which is the same as the slug body. The balance between the turbulent kinetic energy of liquid phase and the surface free energy of dispersed gas bubbles was used to calculate the liquid velocity corresponding to the transition.

The balance equation is shown as follows,

$$E_T = C_e E_S . \tag{34}$$

$$1 + \frac{T_{sm}}{3.16\left[(\rho_L - \rho_g)g\sigma\right]^{1/2}} = \frac{1}{H_{LS}} . \tag{35}$$

where,
$$T_{SM} = \frac{1}{C_e}\left(\frac{f_S}{2}\rho_S v_M^2 + \frac{d}{4}\frac{\rho_L H_{LF}(v_T - v_F)(v_M - v_F)}{l_S}\right) . \tag{36}$$

At lower gas flow rate, Barnea (1987) model was used as the transition boundary and this model is based on the concept of turbulent forces overcoming surface tension forces and dispersing gas-phase into small bubbles in the continuous liquid phase.

$$d_c \geq \left[0.725 + 4.15\left(\frac{v_{SG}}{v_m}\right)^{1/2}\right]. \left(\frac{\sigma}{\rho_L}\right)^{3/5}\left(\frac{2f_m}{d}v_m^3\right)^{-\frac{2}{5}} \quad (37)$$

Where d_c is the critical diameter above which the bubbles will be deformed by the turbulent forces.

$$d_c = 2\left(\frac{0.4\sigma}{(\rho_L - \rho_g)g}\right)^{1/2}. \quad (38)$$

Slug Flow to Bubbly Flow

The bubbly-slug flow pattern transition model is developed based on Taitel *et al.* (1980) model. At low gas flow rates, the discrete bubbles move upward in a linear path without colliding and coalescing because no turbulence exists. When the gas flow rate increases, gas bubbles become larger. When the gas bubbles are above a critical size, they tend to deform and move in a zig-zag path with considerable randomness. The bubbles collide and coalesce and form bubbles with a spherical cap, similar to the Taylor bubbles in slug flow. This results in a transition to slug flow.

Taitel *et al.* (1980) gave gas void fraction 0.25 as the bubbly-slug transition criteria. This is based on the maximum packing idea. The coalescence of bubbles begins to increase sharply when the spacing between bubbles is half their radius. However, according to Caetano's experiments, the average value of void fraction is 0.20 at the bubbly to slug flow pattern transition for flow through concentric annulus. In fully eccentric annulus, the value becomes 0.15. The reason for the lower value of void fraction in fully eccentric annulus is due to the migration of gas bubbles into the wide region of the annulus cross section area which creates a higher local void fraction, so when the local void fraction reached 0.2, the average void fraction for the whole cross-section area is actually lower than 0.2. The bubble shape aspect might be another reason that causes the different void fraction because the bubbles in concentric annulus and fully eccentric annulus are not the same and it affects the bubble rise velocity.

According to the void fraction, the flow pattern transition criterion is given as a relation between superficial gas and liquid velocities.

$$v_{SG} = \frac{a}{1-a}v_{SL} + av_{o,\infty}, \quad (39)$$

Where

$$v_{o,\infty} = 1.53\left[\frac{(\rho_L - \rho_G)g\sigma}{\rho_L^2}\right]^{0.25}. \quad (40)$$

Slug Flow to Churn Flow

In the present study, Kaya (1998) model is used for the slug-churn flow pattern transition. This model is based on the drift flux approach. It was first proposed by Tengesdal *et al.* (1998) and then modified by Kaya (1990). The global void fraction in a slug unit can be expressed by the following equation:

$$a = \frac{v_{SG}}{1.2v_M + v_D}, \quad (41)$$

where

$$v_D = (0.35\sin\theta + 0.54\cos\theta) \left(\frac{g(\rho_L - \rho_g)d}{\rho_L}\right)^{1/2}. \quad (42)$$

Different void fractions for the transition have been used. Chokshi (1994) found that the transition from slug to churn flow occurred when the void fraction is 0.8, while Garber and Varanasi (1997) introduced another value of 0.73. Owen (1986) conducted experiments and found that the transition occurred at a void fraction of 0.78 and that was also used by Kaya in his model. In the present study, 0.64 is used as the void fraction value at the transition from slug flow to churn flow. Substituting 0.64 into Eq. (32) to solve the superficial gas velocity,

$$v_{SG} = 2.7586(1.2v_{SL} + v_D). \quad (43)$$

Closure Relationships

Shear Stress

The shear stresses in the combined momentum equations are calculated as

$$\tau_{FC} = f_{FC}\frac{\rho_L v_{FC}^2}{2}, \quad (44)$$

$$\tau_{FT} = f_{FT}\frac{\rho_L v_{FT}^2}{2}, \quad (45)$$

$$\tau_{IC} = f_{IC}\frac{\rho_C(v_C - v_{FC})|v_C - v_{FC}|}{2}, \quad (46)$$

$$\tau_{IT} = f_{IT}\frac{\rho_C(v_C - v_{FT})|v_C - v_{FT}|}{2}. \quad (47)$$

The friction factors f_{FC} and f_{FT} are estimated in two different ways.

The common way to predict friction factor is the application of hydraulic diameter in the Fanning friction factor calculation,

$$f = C\,\mathrm{Re}^{-n} . \quad (48)$$

Where C=16, n=1 for laminar flow, if the Reynolds number is less than 2000 and C=0.046, n=0.2 for turbulent flow when Reynolds number is larger than 3000. The discontinuity of friction factor in the transition region between laminar flow and turbulent flow was addressed in Zhang *et al.* (2003) Unified Model by interpolation between laminar and turbulent flows.

The Reynolds number for the casing and tubing liquid films and gas core are defined as

$$\mathrm{Re}_{FC} = \frac{4A_{FC}v_{FC}\rho_L}{S_{FC}\mu_L}, \quad (49)$$

$$\mathrm{Re}_{FT} = \frac{4A_{FT}v_{FT}\rho_L}{S_{FT}\mu_L}, \quad (50)$$

$$\mathrm{Re}_{C} = \frac{4A_{C}v_{C}\rho_G}{(S_C + S_I)\mu_L}. \quad (51)$$

The other way to calculate the friction factor in annulus was proposed by Caetano (1986) by taking annulus configuration into account. The friction factor in concentric annulus is determined from solution of the continuity equation, equation of motion and Fanning equation.

For Laminar flow:

$$f_{CA} = \frac{16}{\mathrm{Re}} \frac{(1-K)^2}{\left[\dfrac{1-K^4}{1-K^2} - \dfrac{1-K^2}{\ln(1/K)}\right]}. \quad (52)$$

For turbulent flow,

$$\frac{1}{\left\{ f_{CA}\left(\dfrac{F_P}{F_{CA}}\right)^{0.45\exp\left[-(\mathrm{Re}-3000)/10^6\right]}\right\}^{1/2}} = 4.0\log\left\{\mathrm{Re}\left[f_{CA}\left(\frac{F_P}{F_{CA}}\right)^{0.45\exp\left[-(\mathrm{Re}-3000)/10^6\right]}\right]^{1/2}\right\} - 0.40 \quad (53)$$

A_{FC}, A_{FT} and A_C are cross section areas occupied by the casing liquid film, tubing liquid film and the gas pocket:

$$A_{FC} = H_{LFC}A, \quad (54)$$

$$A_{FT} = H_{LFT}A, \quad (55)$$

$$A_C = (1 - H_{LFC} - H_{LFT})A. \quad (56)$$

Interfacial Friction Factor for Slug Flow

According to Andritsos *et al.* (1987) correlation modified by Zhang *et al.* (2003), the interfacial friction factor can be written as

$$f_{IC} = f_C\left(1.0 + 15.0\left(\frac{2\delta_C}{d_C}\right)^{0.5}\left(\frac{v_{SG}}{v_{SG,t}} - 1.0\right)\right), \quad (57)$$

$$f_{IT} = f_C\left(1.0 + 15.0\left(\frac{2.0\delta_T}{d_T}\right)^{0.5}\left(\frac{v_{SG}}{v_{SG,t}} - 1.0\right)\right), \quad (58)$$

$$v_{SG,t} = 5.0\sqrt{\frac{\rho_a}{\rho_g}}. \quad (59)$$

Where, δ_C and δ_T are casing and tubing liquid film thickness.

Interfacial Friction Factor for Annular Flow

Ambrosini *et al.* (1991) improved Asali (1984) equation for interfacial friction factor in annular flow.

Casing Interfacial Friction Factor

$$f_{IC} = f_{GC}\left(1 + 13.8\mathrm{We}_{GC}^{0.2}\mathrm{Re}_{GC}^{-0.6}\left(h_{FC}^{+} - 200\sqrt{\rho_G/\rho_L}\right)\right), \quad (60)$$

$$We_{GC} = \frac{\rho_G v_C^2 D_C}{\sigma}, \quad (61)$$

$$\mathrm{Re}_{GC} = \frac{\rho_G v_C d_C}{\mu_G}, \quad (62)$$

$$h_{CF}^{+} = \frac{\rho_G \delta_C v_C^{*}}{\mu_G}, \quad (63)$$

$$v_C^{*} = \sqrt{\tau_{IC}/\rho_G}, \quad (64)$$

$$f_{GC} = 0.046\,\mathrm{Re}_{GC}^{-0.2}. \quad (65)$$

Tubing Interfacial Friction Factor

$$f_{IT} = f_{GT} \left(1+13.8We_{GT}^{0.2}\,\mathrm{Re}_{GT}^{-0.6}(h_{FT}^{+}-200\sqrt{\rho_G/\rho_L})\right), \quad (66)$$

$$We_{GT} = \frac{\rho_G v_C^2 d_T}{\sigma}, \quad (67)$$

$$\mathrm{Re}_{GT} = \frac{\rho_G v_C d_T}{\mu_G}, \quad (68)$$

$$h_{FT}^{+} = \frac{\rho_G \delta_T v_C^*}{\mu_G}, \quad (69)$$

$$v_C^* = \sqrt{\tau_{IT}/\rho_G}, \quad (70)$$

$$f_{GT} = 0.046\,\mathrm{Re}_{GT}^{-0.2}. \quad (71)$$

Casing and Tubing Liquid Film Holdup Ratio

According to Caetano (1986), liquid film holdup equation and liquid film thickness are related as follows:

$$H_{LFC} = \frac{4\delta_C}{d_C}\frac{\left(1+\frac{\delta_C}{d_T}\right)}{(1-K^2)}, \quad (72)$$

$$H_{LFT} = \frac{4\delta_T}{d_C}K\frac{\left(1+\frac{\delta_T}{d_T}\right)}{(1-K^2)}, \quad (73)$$

$$\frac{H_{LFT}}{H_{LFC}} = \frac{\delta_T}{\delta_C}\frac{K\left(1+\frac{\delta_T}{d_T}\right)}{\left(1-\frac{\delta_C}{d_C}\right)}. \quad (74)$$

Since δ_C and δ_T are very small compared to casing and tubing diameter, δ_T/d_T and δ_C/d_C can be neglected and the Eq. (71) can be written as:

$$\frac{H_{LFT}}{H_{LFC}} = \frac{\delta_T}{\delta_C}K. \quad (75)$$

Caetano proposed an expression for the tubing and casing liquid film thickness ratio,

$$\frac{\delta_T}{\delta_C} = \frac{W_T'}{(2\pi - W_T')K}, \quad (76)$$

$$W_T' = \frac{1}{(1-K^2)}(2*\sin^{-1}(K)+2K\sqrt{1-K^2}-K^2\pi). \quad (77)$$

Wetted Wall Fraction and Interfacial Perimeter

Based on the annulus geometry and the assumption of uniform film thickness, the perimeters of casing liquid film, tubing liquid film and gas core are respectively given by

$$S_{IC} = \pi(d_C - 2\delta_C), \quad (78)$$

$$S_{IT} = \pi(d_T - 2\delta_T), \quad (79)$$

$$S_{CC} = \pi d_C, \quad (80)$$

$$S_{CT} = \pi d_T. \quad (81)$$

The wetted wall perimeter for casing and tubing are

$$S_{FC} = \pi d_C, \quad (82)$$

$$S_{FT} = \pi d_T. \quad (83)$$

The hydraulic diameters for casing and tubing wetted wall and gas core are respectively given by,

$$d_{FC} = 4\left(\frac{A_{FC}}{S_{FC}+S_{IC}}\right) \quad (84)$$

$$d_{FT} = 4\left(\frac{A_{FT}}{S_{FT}+S_{IT}}\right) \quad (85)$$

$$d_C = 4\left(\frac{A_C}{S_{CC}+S_{CT}}\right) \quad (86)$$

Liquid Entrainment in Gas Core

Wallis *et al.* (1969) correlation is used to calculate the liquid entrainment in the gas core.

$$f_E = 1-\exp[-0.125(\phi-1.5)]. \quad (87)$$

Where,

$$\phi = 10^4\frac{v_{SG}\mu_G}{\sigma}\left(\frac{\rho_G}{\rho_L}\right)^{0.5}. \quad (88)$$

Slug Liquid Holdup

Zhang *et al.* (2003) developed a mechanistic model to predict slug liquid holdup based on the balance between turbulent kinetic energy of the liquid phase and the surface free energy of dispersed gas bubbles in the slug body. This model can be used in annulus slug flow using the representative diameter concept.

$$H_{LS} = \frac{1}{1+\dfrac{T_{sm}}{3.16((\rho_L-\rho_G)g\sigma)^{1/2}}}, \quad (89)$$

$$T_{sm}=\frac{1}{C_e}\left(\begin{array}{l}\frac{f_S}{2}\rho_S v_S^2+\\ \frac{d_C}{4}\frac{\rho_L(H_{LFC}+H_{LFT})(v_T-v_{FC})(v_S-v_{FC})}{l_S}\\ +\frac{d_C}{4}\frac{\rho_C(1-H_{LFC}-H_{LFT})(v_T-v_C)(v_s-v_C)}{l_S}\end{array}\right), \quad (90)$$

$$C_e=\frac{2.5-|\sin\theta|}{2}, \quad (91)$$

$$\rho_S=\rho_L H_{LS}+\rho_G(1-H_{LS}), \quad (92)$$

$$\mathrm{Re}=\frac{\rho_S v_S d_C}{\mu_L}. \quad (93)$$

Before solving the combined momentum equation, the slug liquid holdup is estimated using the Gregory *et al.* (1978) correlation,

$$H_{LS}=\frac{1}{1+\left(\frac{v_S}{8.66}\right)^{1.39}}. \quad (94)$$

Translational Velocity and Slug Length
Hasan and Kabir (1990) proposed a new equation for drift velocity in upward annuli, and it proves to perform better than the Sadatomi *et al.* (1982) Taylor bubble rise velocity. The equation is expressed as

$$v_D=0.345(1+0.1K)\sqrt{gd_c(\rho_L-\rho_G)/\rho_L}. \quad (95)$$

The translational velocity can be expressed as

$$v_T=C_S v_S+v_D. \quad (96)$$

The coefficient C_S is considered to be the ratio of the maximum to the mean velocity of a fully developed velocity profile and it varies with different conditions. According to Nicklin (1962), Bendiksen (1984) and Zhang *et al.* (2003), C_S equals to 2.0 in laminar flow, and 1.3 in turbulent flow. In the transition area ($2000<\mathrm{Re}<4000$) between laminar and turbulent flow, C_S is given by

$$C_S=2.0-0.7*(\mathrm{Re}-2000)/2000. \quad (97)$$

It has been proposed that the slug length is related to pipe diameter, but the closure relationship for slug flow varies with different models or correlations. According to Taitel *et al.* (1980) and Barnea and Brauner (1985), the slug length for vertical pipes can be estimated by applying the representative diameter concept in annulus slug flow.

$$l_S=16.0d. \quad (98)$$

In present study, representative diameter is used in the flow pattern transition models and hydrodynamic models.

Model Performance

The mechanistic model developed in this study has been coded in FORTRAN to calculate liquid holdup and pressure gradient.

Caetano's (1986) experimental flow pattern data are the only data available to be used in the present study to evaluate the flow pattern transition models. The two-phase fluids used in these experiments were air-water and air-kerosene. According to the characteristic configuration of annuli and the different fluids used in experiments, the experimental data include three sets: air and water in concentric annulus, air and water in fully eccentric annulus, air and kerosene in concentric annulus. Each set of experimental data were run with the new model and modified TUFFP unified model separately. The simulation results of pressure gradient and liquid holdup were then compared with the experimental results to evaluate the performance of new model and TUFFP unified model.

Comparison Criteria

Two different statistical parameters are used to evaluate the new model and unified model predictions using Caetano's experimental data. Following are the definitions of the statistical parameters used for this purpose.

Absolute Average Percentage Error
The absolute average percentage error is expressed as follows:

$$E_3=\left[\frac{1}{N}\sum_{j=1}^{N}\left|e_{rj}\right|\right]\times 100. \quad (99)$$

Smaller values of this parameter indicate a better agreement with the experimental results.

Standard Deviation
The standard deviation is expressed as:

$$E_3=\sum_{j=1}^{N}\sqrt{\frac{\left(e_{rj}-E_2\right)^2}{N-1}}. \quad (100)$$

The standard deviation is a simple measure of the variability or dispersion of a data set and it indicates the scatter of the error about its average error. The smaller value of E_3 indicates that the data points are close to the average value.

Flow Pattern Transitions

The flow pattern data are usually presented using gas and liquid superficial velocities as mapping coordinates. The experimental flow pattern maps obtained for air and water in concentric annulus, and air and kerosene in concentric annulus are given in Figs. 4 and 5, respectively. Figure 6 shows the flow pattern map for air and water in fully eccentric annulus.

Figure 4 shows that the predictions from flow pattern transition models are consistent with the experimental results. However, there is still discrepancy for slug-churn flow pattern transition. This might be due to the uncertainty in classification of slug and churn flows since it is hard to visually identify the slug or churn flow near the transition.

Comparing Fig. 4 with Fig. 5, it can be seen that the flow pattern transition model for air and kerosene flow in concentric annulus does not work as well as for air and water flows in concentric annulus. The air and kerosene flow experiences earlier transition to annular flow than air and water flow.

Figures 4 and 6 show no major difference between flow pattern map for air and water flows in concentric and fully eccentric annuli. The flow pattern transition model performs well for air and water flow in fully eccentric annulus.

Overall Performance of Hydrodynamic Models

The hydrodynamic models are compared with Caetano's (1986) experimental data.

Figures 7 to 10 show the comparisons for the liquid holdup and pressure gradient predicted by the present hydrodynamic model and the unified model against Caetano's (1986) slug flow experimental data of air-water flow in concentric annulus. The absolute average error of the new model in predicting liquid holdup and pressure gradient are 8.7% and 12.7%. In comparison, the unified model average errors are 16.2% and 19.5%.

Figures 11 to 14 are the overall comparison results between the present model predictions and the unified model against Caetano's (1986) experimental data of air-water flow in fully eccentric annulus. The absolute average errors of the new model for liquid holdup and pressure gradient prediction are 8.4% and 13.4%. In comparison, the unified model average errors are 15.7% and 21.1%.

Figures 15 to 18 show the comparisons for liquid holdup and pressure gradient predictions against Caetano's (1986) experimental data for air and kerosene in concentric annulus. The absolute average errors of liquid holdup and pressure gradient of present model are 12.7% and 16.1%. The corresponding average errors for unified model are 20.6%, and 23.5%.

The above comparisons show good agreement between experimental results and the new hydrodynamic model. The performance of the new model is significantly better when compared with that of the unified model. Comparing the three different flow conditions, the new model works better for air and water flows in both concentric and fully eccentric annulus than air and kerosene flow in concentric annulus.

Large errors exist in the comparisons for different flow conditions. One of the reasons is due to the inaccuracy of the flow pattern prediction, e.g. annular flow predicted as churn flow. Another reason might be related to the uncertainties of experimental measurements. More experimental data or field data are required to evaluate the new model.

Individual Hydrodynamic Models Performance

The hydrodynamic models for slug flow, annular flow and churn flow are evaluated against Caetano's (1986) experimental data separately, in order to better understand the individual hydrodynamic model performance.

The results of the performance are given in Tables 1-5 for individual hydrodynamic model: slug flow, churn flow, annular flow, dispersed bubble flow, and bubble flow, respectively.

Slug Flow Model Performance

Table 1 shows the comparisons of the liquid holdup and pressure gradient predicted by the present hydrodynamic model and the unified model for slug flow against Caetano's slug flow experimental data. The comparison results are given for air-water flow in concentric annulus, air and water in fully eccentric annulus and air and kerosene in concentric annulus, respectively. The comparison results show that the unified model simulation results are more scattered, and the new model predicts liquid holdup and pressure gradient significantly better.

Comparing the performance for different flow conditions, the model performs best for fully eccentric annulus flow, suggesting that the model adequately accounts for the effect of the eccentricity.

Annular Flow Model Performance
Table 2 shows the comparisons for liquid holdup and pressure gradient predicted by the present model and the unified model against Caetano's (1986) experimental data of annular flow. The comparison results are given for air-water flow in concentric annulus, air and water in fully eccentric annulus and air and kerosene in concentric annulus, respectively.

The comparison results indicate that the new model improves the degree of prediction accuracy for liquid holdup and pressure gradient over the unified model. However, the error of the liquid holdup and pressure gradient prediction is still high when compared to other flow patterns. Comparing the performance for different flow conditions, the model performs best for air and water in concentric annulus and it does not perform well for air and water in fully eccentric annulus.

The large error of the model performance might be due to the uncertainties of the experimental measurements for annular flow as mentioned by Caetano (1986). The quick-closing ball valves used to measure liquid holdup were less accurate than for other flow patterns due to the low liquid holdup value for annular flow, and the reported holdup values are sometimes below the 3% minimum value possible to measure. Another reason might be due to interfacial friction factors. The interfacial friction factors acting on the interface between casing and tubing liquid films and gas core have significant effect on the pressure gradient and liquid holdup calculations. Careful selection of the interfacial friction factors is critical for the model prediction accuracy. Therefore, more data of annular flow in concentric and eccentric annuli and modifications of some closure relationships are required to improve the model performance.

Churn Flow Model Performance
Table 3 shows the comparison results for liquid holdup and pressure gradient predictions by the present model and the unified model against Caetano's (1986) experimental results of churn flow.

The comparison results show that the new model performs better than the unified model in predicting liquid holdup and pressure gradient prediction. From the performance curves, it is seen that the unified model under predicts problem both liquid holdup and pressure gradient prediction in three flow conditions.

Bubble Flow Model Performance
Tables 4 and 5 present the new model simulation results for liquid holdup and pressure gradient for bubbly flow and dispersed bubble flow.

The results show that the dispersed bubble flow model performs very well for all flow conditions. For bubbly flow, the agreement between the model and experimental results is good, showing acceptable absolute average error and degree of scatter. However, the agreement is weaker for air and water in fully eccentric annulus. This might be caused by the data acquisition mentioned by Ceatano in his thesis. Part of the data was obtained at very low superficial phase velocities and the flow was very unstable under this condition due to a heading phenomenon.

Conclusions

In this study, a mechanistic model is developed to predict flow patterns, pressure gradient and liquid holdup for gas-liquid flow in upward vertical concentric and fully eccentric annuli.

The flow pattern transition model consists of modified Zhang *et al.* (2003) unified model for predicting dispersed bubble flow transition and annular flow pattern transition, Caetano's (1986) bubbly flow pattern transition and modified Kaya's (1990) slug to churn flow pattern transition. The bubbly flow pattern transition model takes annulus configuration into account by applying different values for transition liquid holdup in concentric and fully eccentric annulus.

The hydrodynamic mechanistic models for each flow pattern are developed based on the dynamics of slug flow and the film zone is used as the control volume. For annulus flow configuration, two liquid films are considered in the control volume for hydrodynamic models of slug flow and annular flow. A new churn flow model is developed based on Zhang *et al.* (2003) unified model for pipe flow by modifying the slug length and slug liquid holdup. Zhang *et al.* unified model idea is first introduced to the hydrodynamic model development in annuli and the two liquid films are taken into account in the annulus slug flow model development.

The flow pattern transition models and hydrodynamic models for each flow patterns are compared with Caetano's (1986) experimental data and predictions of Zhang *et al.* (2003) unified model. The comparison results show that the performance of the new hydrodynamic models for each flow pattern is significantly improved over the unified model. The slug flow, churn flow, dispersed bubble flow and bubbly flow models perform very well and the average absolute errors for liquid holdup and pressure gradient are within 10%. The models work better for fully eccentric annulus. The annular flow model still needs to be further evaluated and improved due to the current limited data points.

Recommendation for Future Studies

The new model is evaluated only with Caetano's (1986) experimental data. More experimental data are required to give a better understand and improve the annulus flow modeling.

Nomenclature

A	= cross section area
d	= pipe diameter
e	= eccentricity
f	= friction factor
F_E	= liquid entrainment
F_r	= Froude number
H	= liquid holdup
K	= pipe diameter ratio
l	= length of the slug unit
P	= pressure
R_e	= Reynolds number
S	= perimeter
v	= velocity
W_e	= Weber number

Greek Letters

α	= gas void fraction
δ	= liquid film thickness
Θ	= pipe circumferential wetted fraction
μ	= viscosity
θ	= pipe inclination angle
ρ	= density
σ	= surface tension
τ	= shear stress

Subscripts

C	= casing or gas core
CA	= concentric annulus
D	= drift
FC	= casing film
FT	= tubing film
GC	= gas casing
GT	= gas tubing
H	= hydraulic diameter
IC	= casing interfacial
IT	= tubing interfacial
LS	= slug liquid holdup
LFC	= casing liquid film holdup
LFT	= tubing liquid film holdup
r	= representative diameter
S	= slug
SL	= liquid superficial
SG	= gas superficial
T	= tubing or translational velocity

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Table 1: Comparison Results for Slug Flow

	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_{Lnew}	6.7%	4.0%	8.6%	6.7%	3.8%	5.8%
$H_{Lunified}$	12.4%	12.5%	19.0%	10.9%	8.8%	9.7%
$-dp/dz_{new}$	8.3%	5.8%	11.4%	7.8%	3.8%	5.3%
$-dp/dz_{unified}$	14.7%	15.4%	23.1%	12.0%	11.9%	17.0%

Table 2: Comparison Results for Annular Flow

	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_{Lnew}	19.7%	18.1%	24.1%	10.3%	10.9%	15.3%
$H_{Lunified}$	28.6%	24.2%	30.7%	12.2%	14.0%	16.3%
$-dp/dz_{new}$	14.5%	19.4%	14.0%	11.8%	15.2%	14.1%
$-dp/dz_{unified}$	14.8%	15.4%	48.5%	11.1%	16.4%	26.9%

Table 3: Comparison Results for Churn Flow

	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_{Lnew}	8.6%	11.2%	9.7%	7.9%	5.9%	7.2%
$H_{Lunified}$	26.9%	32.1%	26.5%	9.3%	10.3%	6.1%
$-dp/dz_{new}$	9.8%	16.6%	13.1%	7.8%	9.0%	11.5%
$-dp/dz_{unified}$	34.5%	39.8%	23.1%	34.2%	16.6%	19.6%

Table 4: Comparison Results for Bubble Flow

	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_L	3.2%	3.5%	3.0%	2.9%	2.9%	3.7%
$-dp/dz$	9.9%	12.2%	2.9%	6.4%	6.0%	2.7%

Table 5: Comparison Results for Dispersed Bubble Flow

	E_{2wc}	E_{2we}	E_{2kc}	E_{3wc}	E_{3we}	E_{3kc}
H_L	2.7%	1.6%	4.6%	2.0%	0.9%	2.9%
$-dp/dz$	4.9%	2.5%	2.9%	5.3%	1.8%	2.2%

where $E_{2wc}, E_{2we}, E_{2kc}, E_{3wc}, E_{3we}, E_{3kc}$ refer to the average absolute error and standard deviation for air and water in concentric annulus, air and water in fully eccentric annulus, air and kerosene in concentric annulus, $H_{Lnew}, H_{Lunified}$, dp/dl_{new}, $dp/dl_{unified}$ refer to the liquid holdup and pressure gradient for new model and unified model simulation results.

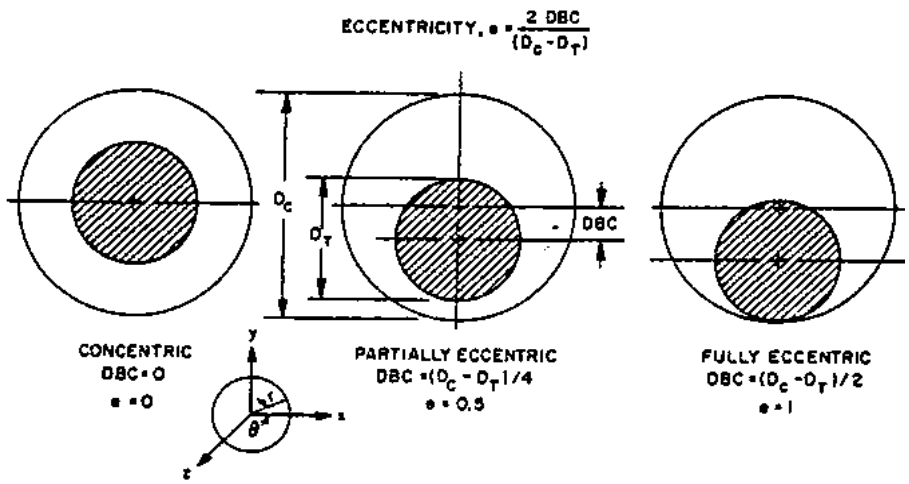

Figure 1: Annular Flow Configuration (Caetano, 1986)

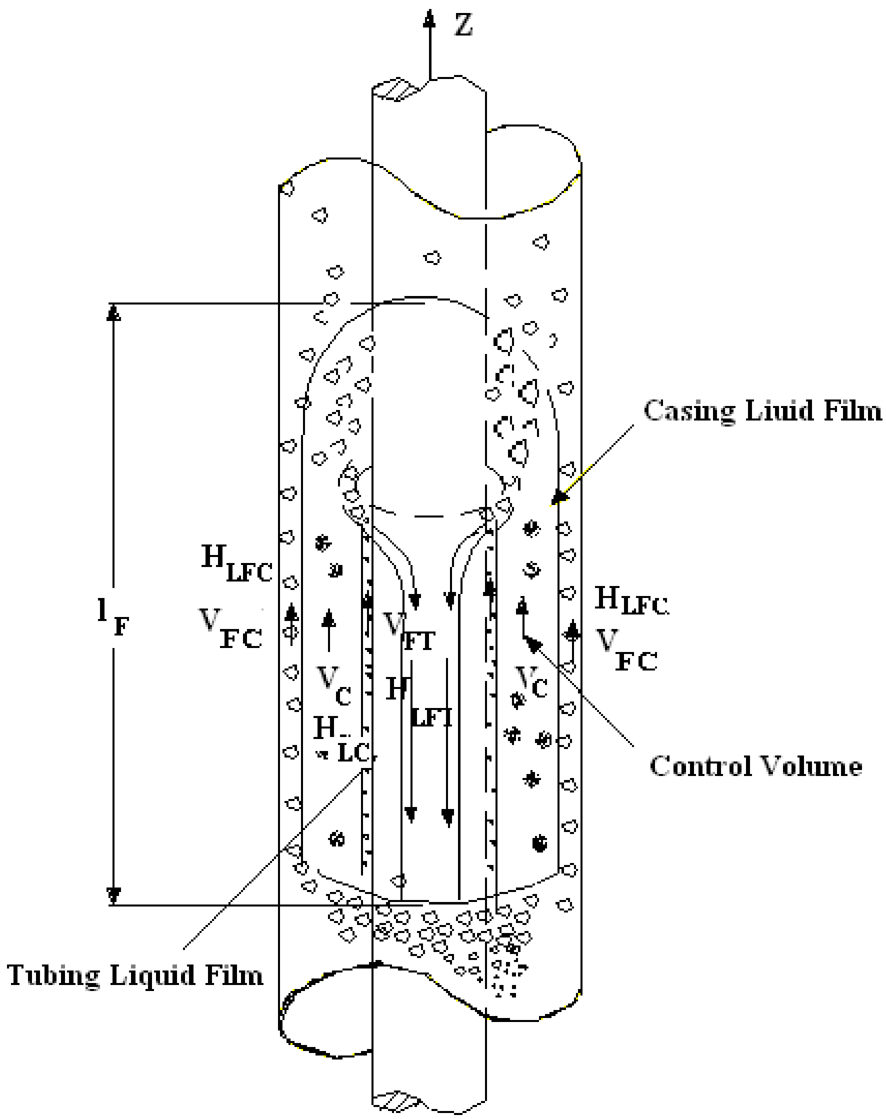

Figure 2: Control Volume Used in Slug Flow Modeling

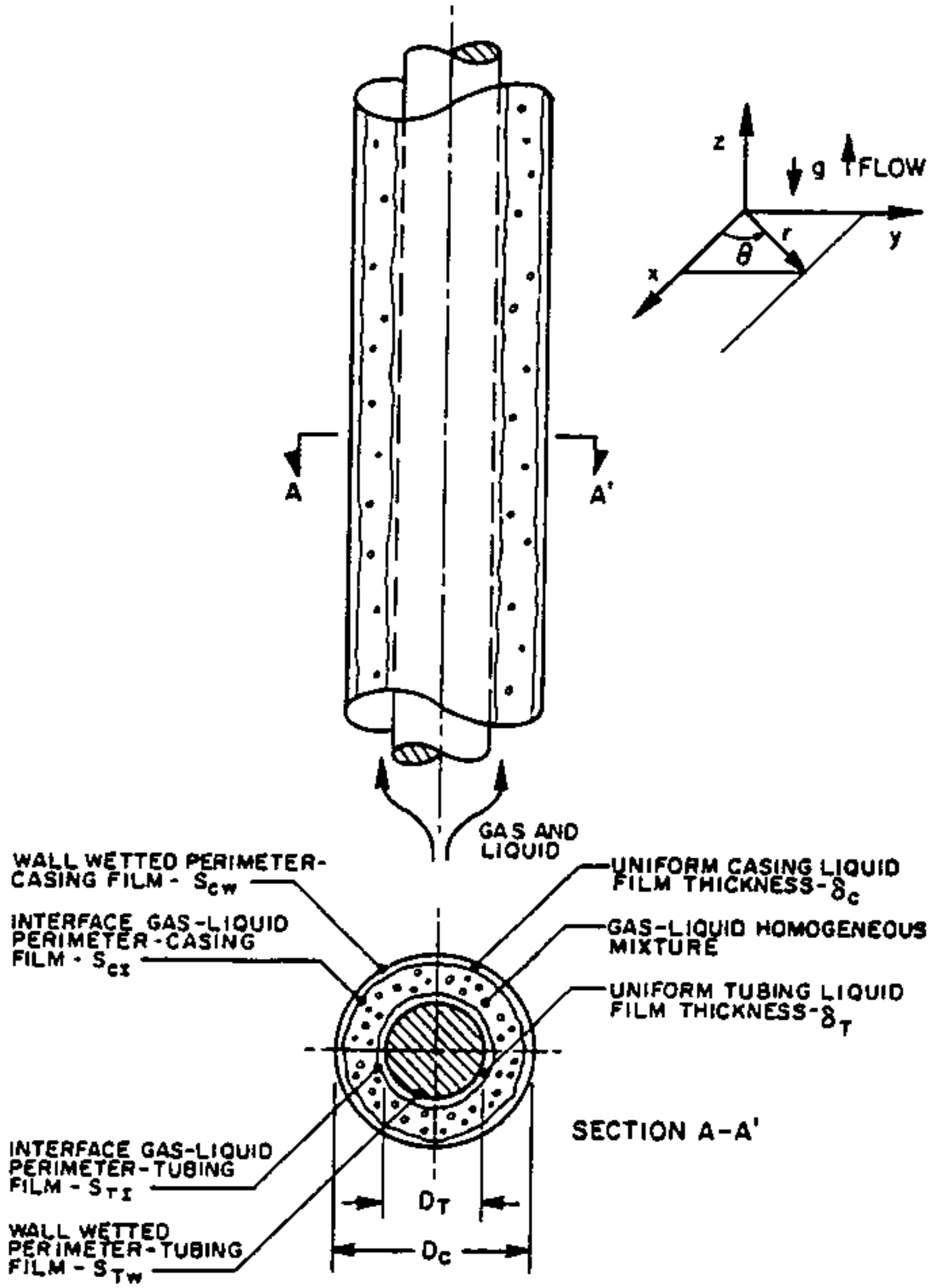

Figure 3: Idealized Annular Flow in Concentric Annulus-Geometry and Parameters

(Caetano, 1986)

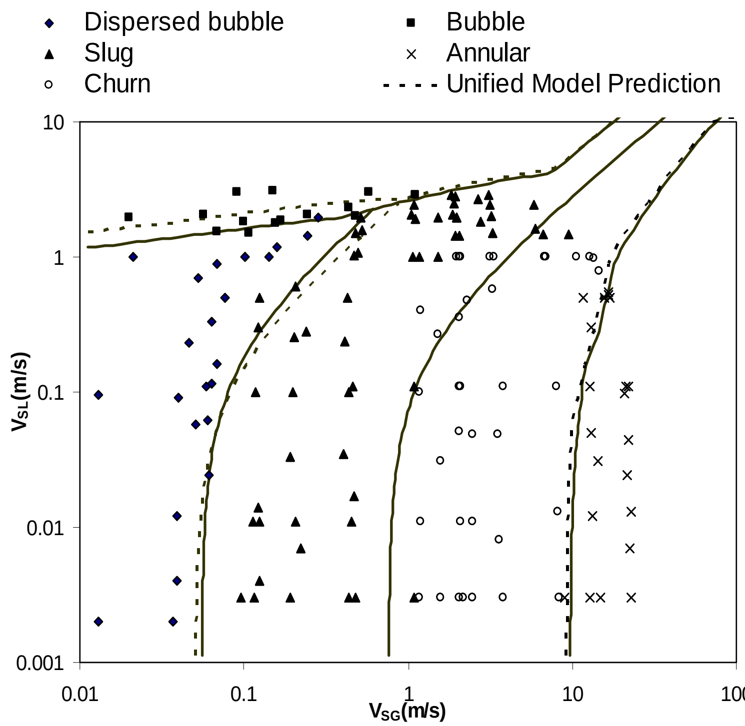

Figure 4: Flow Pattern Map for Air and Water Flow in Upward Vertical Concentric Annulus

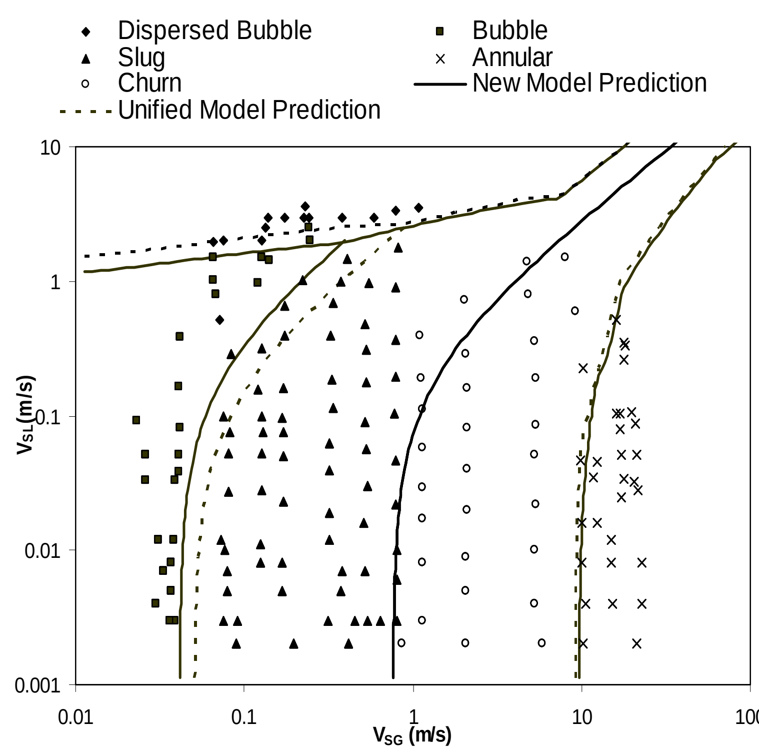

Figure 5: Flow Pattern Map for Air and Water Flow in Upward Vertical Fully Eccentric Annulus

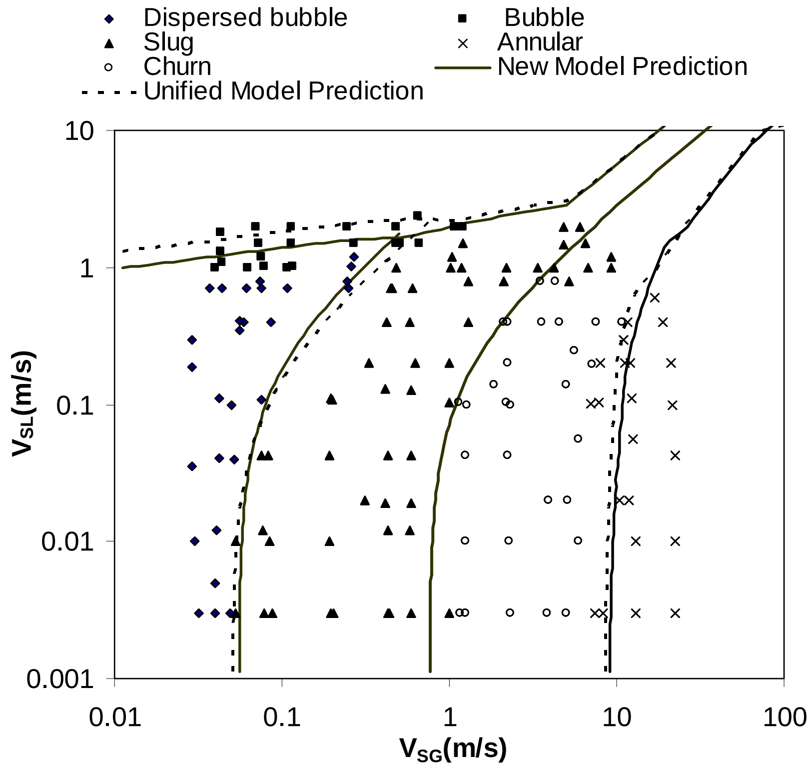

Figure 6: Flow Pattern Map for Air and Kerosene Flow in Upward Vertical Concentric Annulus

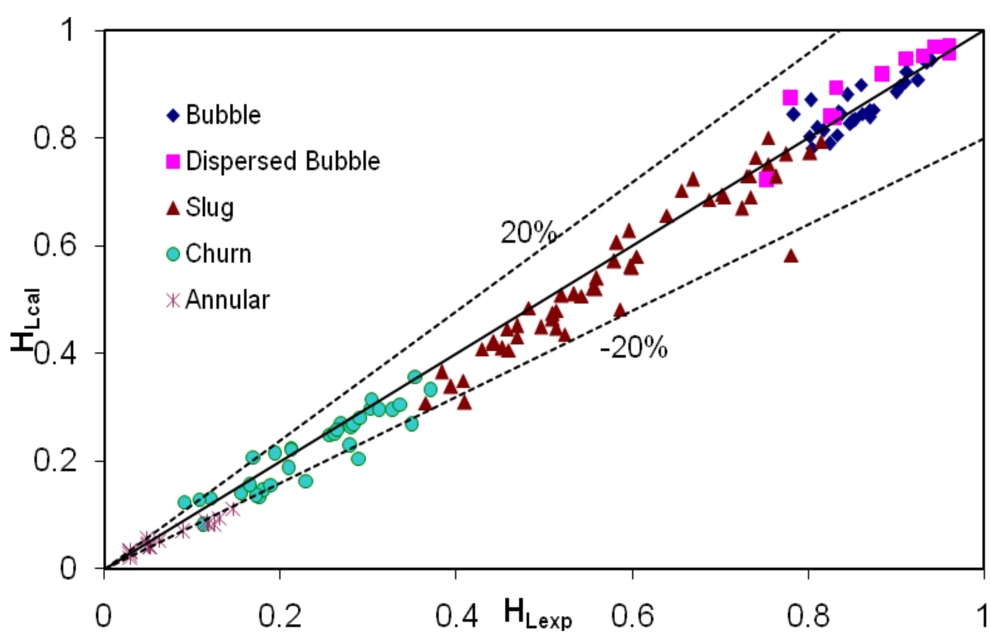

Figure 7: Present Model Performance for Liquid Holdup Prediction
(Air and Water Flow in Concentric Annulus)

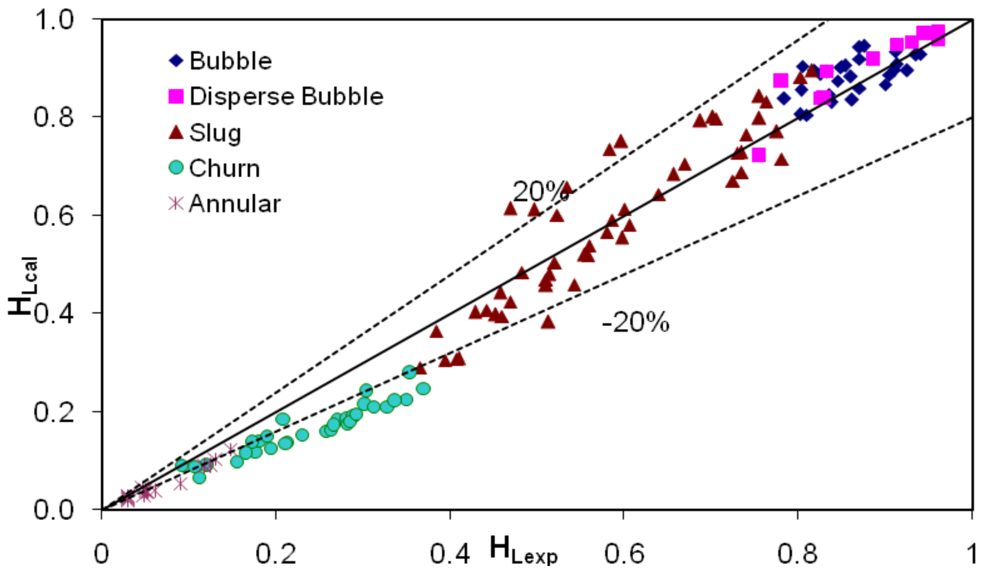

Figure 8: Zhang *et al.* Unified Model Performance for Liquid Holdup Prediction (Air and Water Flow in Concentric Annulus)

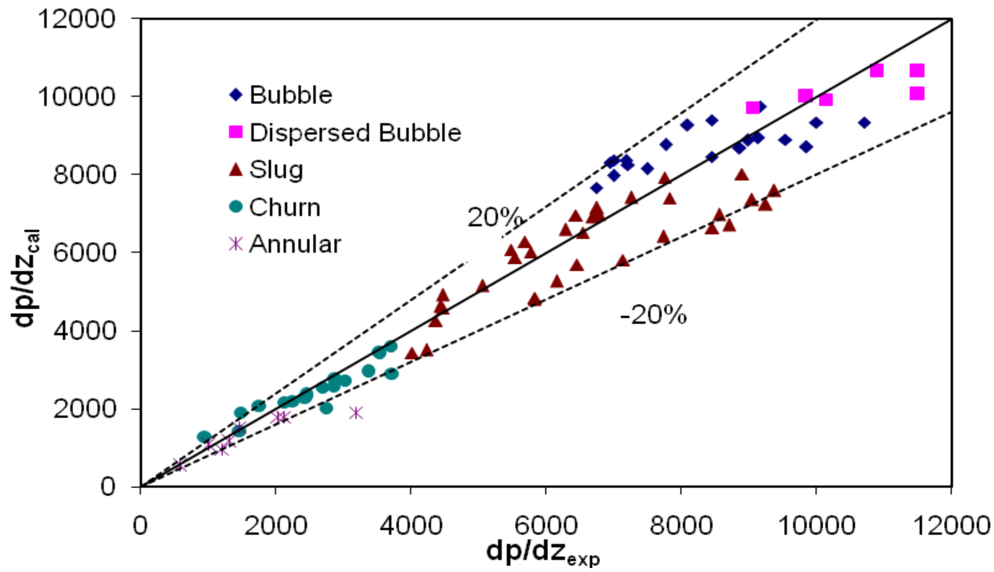

Figure 9: Present Model Performance for Pressure Gradient Prediction (Air and Water Flow in Concentric Annulus)

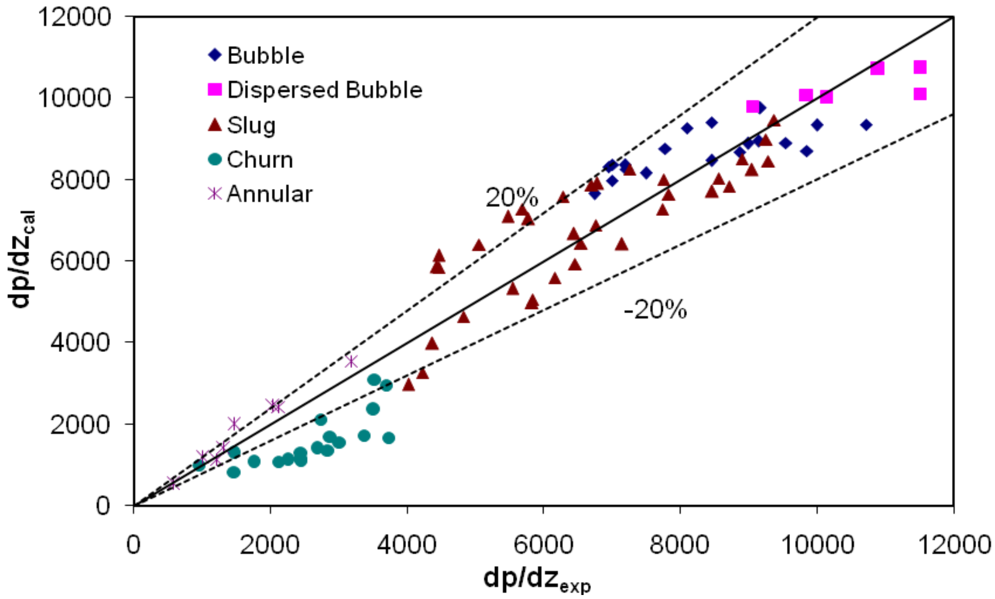

Figure 10: Zhang *et al.* Unified Model Performance for Pressure Gradient Prediction (Air and Water Flow in Concentric Annulus)

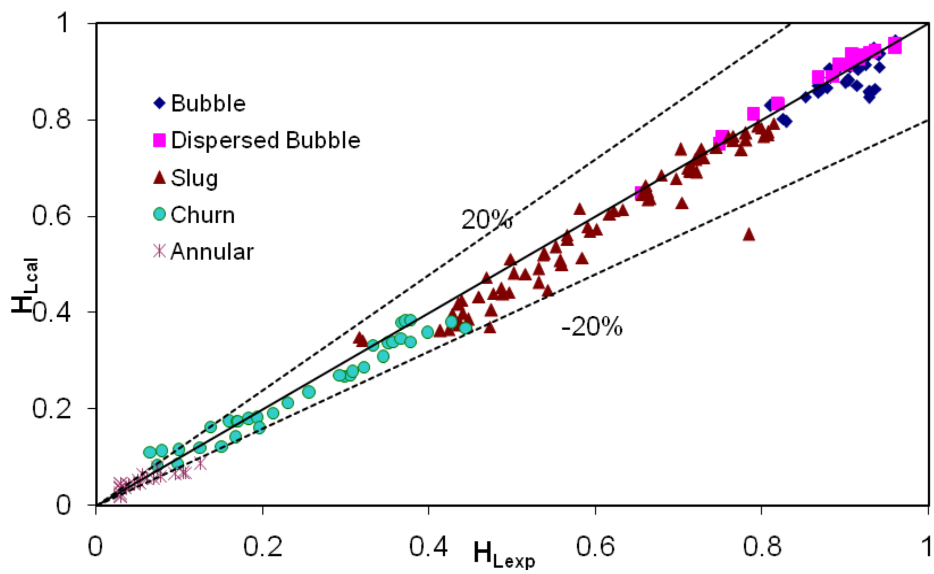

Figure 11: Present Model Performance for Liquid Holdup Prediction (Air and Water Flow in Fully Eccentric Annulus)

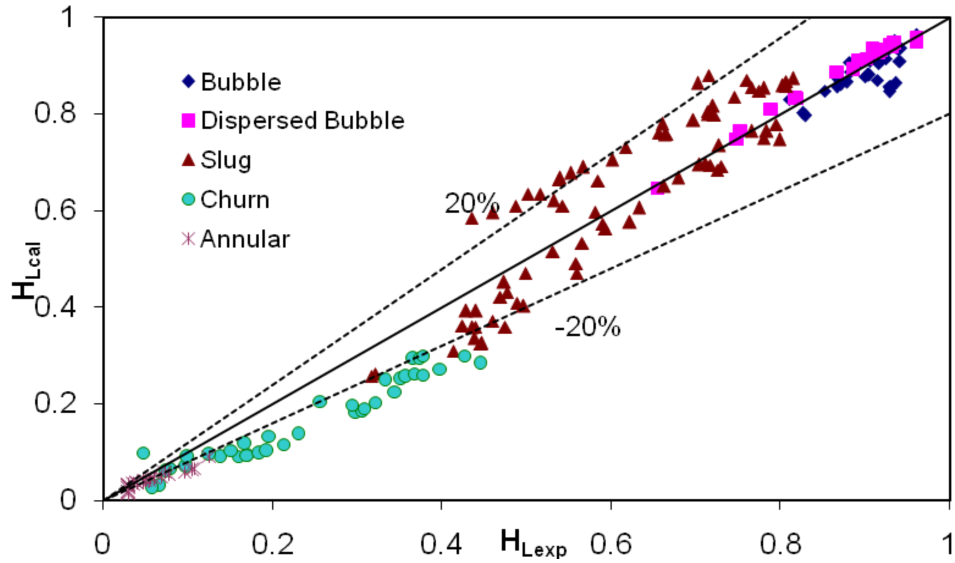

Figure 12: Zhang *et al.* Unified Model Performance for Liquid Holdup Prediction (Air and Water Flow in Fully Eccentric Annulus)

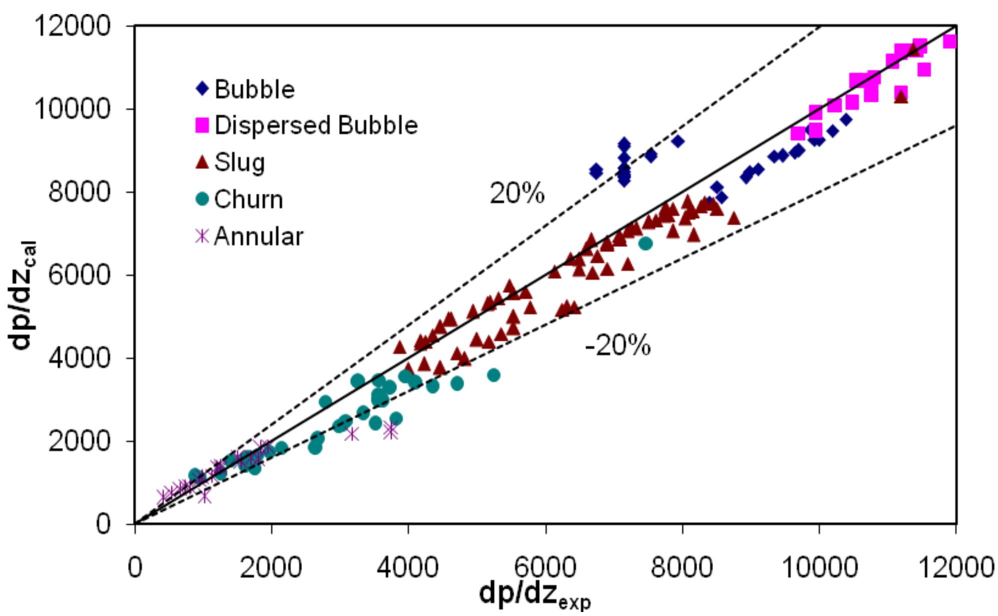

Figure 13: Present Model Performance for Pressure Gradient Prediction (Air and Water Flow in Fully Eccentric Annulus)

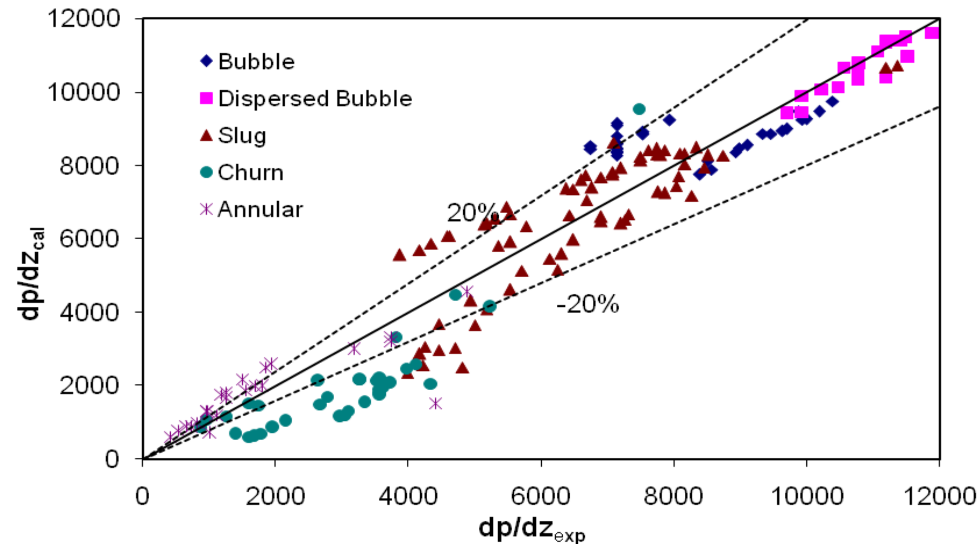

Figure 14: Zhang *et al.* Unified Model Performance for Pressure Gradient Prediction (Air and Water Flow in Fully Eccentric Annulus)

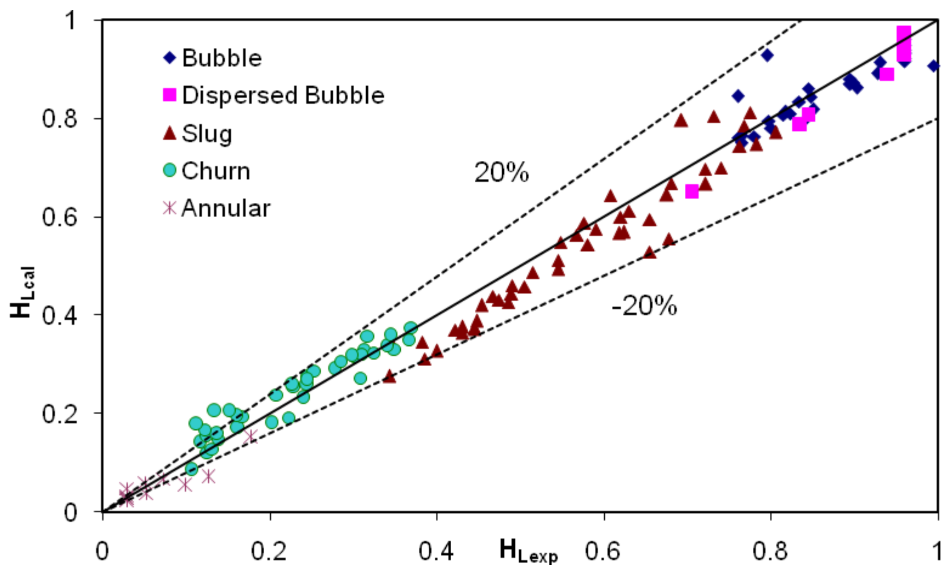

Figure 15: Present Model Performance for Liquid Holdup Prediction (Air and Kerosene Flow in Concentric Annulus)

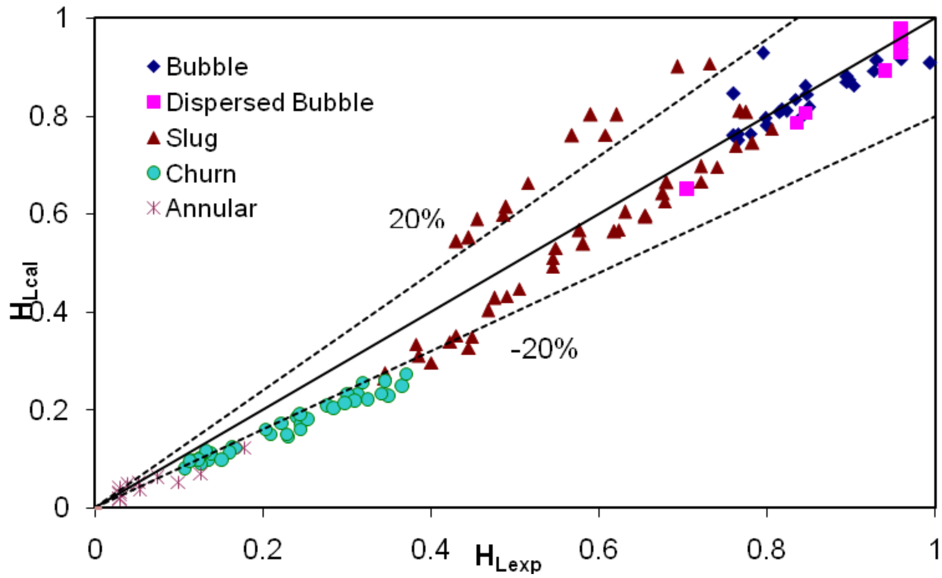

Figure 16: Zhang *et al.* Unified Model Performance for Liquid Holdup Prediction (Air and Kerosene Flow in Concentric Annulus)

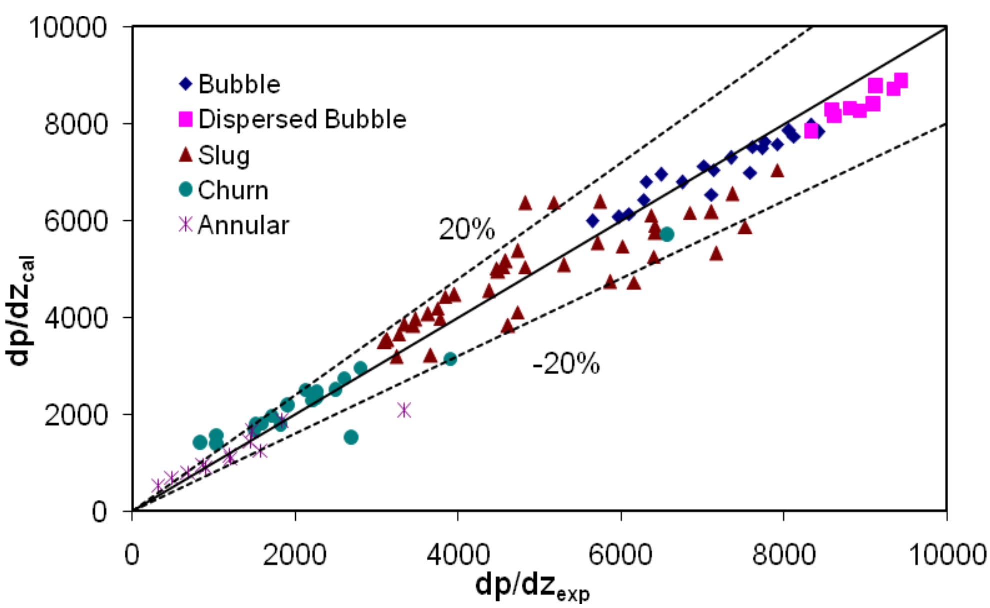

Figure 17: Present Model Performance for Pressure Gradient Prediction (Air and Kerosene Flow in Concentric Annulus)

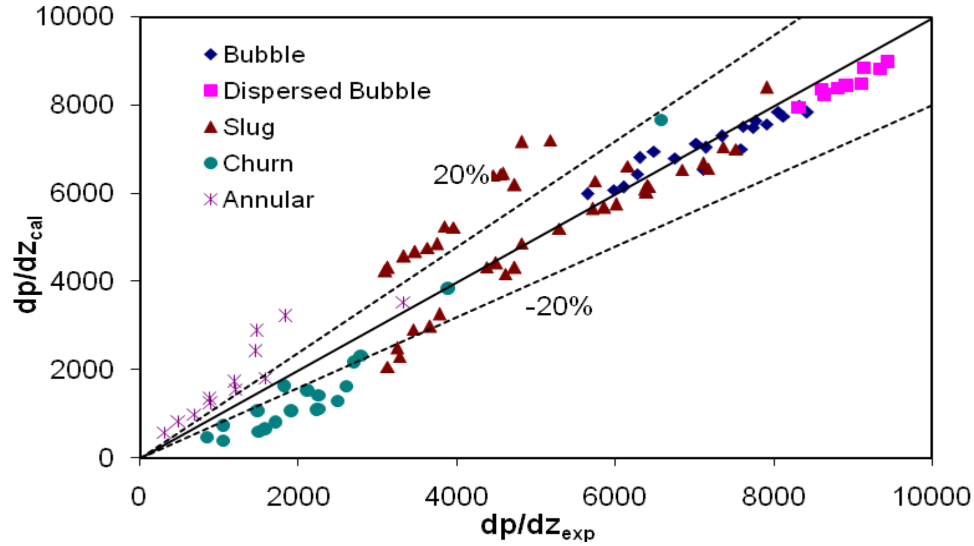

Figure 18: Zhang *et al.* Unified Model Performance for Pressure Gradient Prediction (Air and Kerosene Flow in Concentric Annulus)

Three-phase Flow Studies

- Significance
  - Good Understanding of Gas-Oil Flow
  - Poor Understanding of Gas-Oil-Water Flow
- Objective
  - Development of Improved Prediction Models
- Past Studies
  - Oil-Water
    - Trallero (1994), Horizontal
    - Flores (1996), Vertical and Deviated
    - Alkaya (1999), Inclined

Three-phase Flow Studies …

- Past Studies …
  - Three-phase
    - Keskin (2007), Experimental Horizontal Three-phase Study
    - Zhang and Sarica (2005), Three-phase Mechanistic Model Development
    - Need to More Research on Oil-Water Flow
  - Recent Oil-Water Studies with Emphasis on Droplets
    - Vielma (2006), Horizontal Flow
    - Atmaca (2007), Inclined Flow

Three-phase Flow Studies …

- Current Study (Oil-Water Flow Modeling in Pipelines)
  - Progress
    - New Modeling Approach Based on Energy Minimization
    - Performs Better than Other Models When Compared with Data

Fluid Flow Projects

Modeling of Hydrodynamics and Dispersions in Oil-Water Pipe Flow

Anoop Kumar Sharma

Advisory Board Meeting, March 25, 2009

Outline

- Objectives
- Significance
- Preliminary Model
- Modeling
- Results
- Conclusions

Objectives

- Development of a Better Model for Oil-water Flow
- Validation of Model Using Available Present Experimental Data

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Significance

- Oil-water Flow is Encountered in Various Processes in Petroleum Industry
- Existing Predictive Models
  - Do not Properly Represent Physics
  - Can not Capture Gradual Transition Between Flow Patterns

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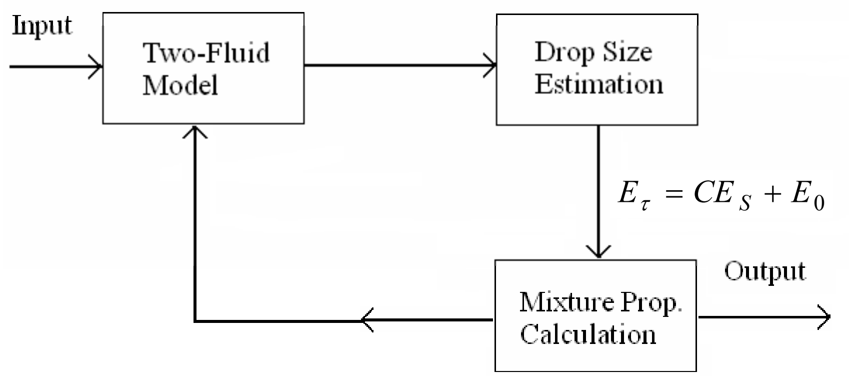
Preliminary Model (Last ABM)



Preliminary Model (Last ABM) ...

- Closure Relationship

$$E_\tau = CE_S + E_0$$

- Challenge With Closure Relationship
  - Estimation of E_0
  - Lack of Experimental Study and Data

Modeling

- Switched to New Modeling Approach
- Working Principle
  - A System Stabilizes to It's Minimized Total Energy
  - Charabarti et al. (2005) Used Similar Approach for Horizontal Segregated Flow Pattern

Modeling …

- Segregated Layers
  - *Combined Momentum Balance Equation → 0*
  - *Total Energy → Minimum*
- Full Dispersion
  - *Total Energy → Minimum*

Modeling …

- Transition to Dispersion
  - Can D_{MAX} be Accommodated in Continuous Phase?
- Angeli and Hewitt (2000)

$$d_{SM} v_C^{1.8} = 2 \times 10^{-8} \times f_C^{-3.12}$$

$$d_{MAX} v_C^{1.8} = 4.2 \times 10^{-8} \times f_C^{-3.13}$$

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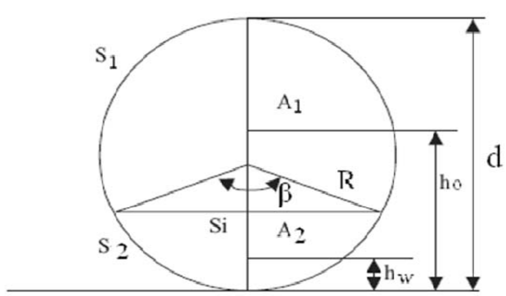
Modeling …

- Pipe Geometry



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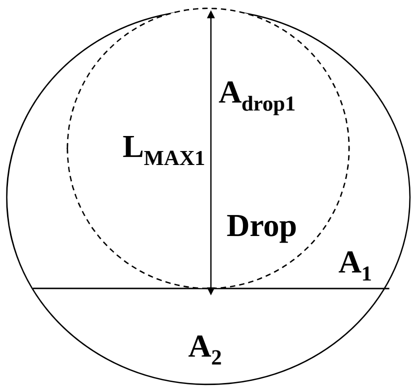
Modeling …

- Transition to Dispersion



$$\left(\frac{A_{drop1}}{A_1}\right) \le H_{INV1}$$

$$d_{MAX1} \le L_{MAX1}$$

Modeling …

- Mixture Property Calculation
  - Density

$$\rho_1 = \rho_o\left(1 - H_{D1}\right) + \rho_w H_{D1}$$

$$\rho_2 = \rho_w\left(1 - H_{D2}\right) + \rho_o H_{D2}$$

  - Viscosity (Brinkman Correlation)

$$\mu_1 = \mu_O\left(1 - H_{D1}\right)^{-2.5}$$

$$\mu_2 = \mu_w\left(1 - H_{D2}\right)^{-2.5}$$

Modeling …

- Inversion Point (Zhang et al., 2003)

$$H_{O,INV} = \frac{\left(\frac{\mu_O}{\mu_W}\right)^{0.4}}{\left(1+\frac{\mu_O}{\mu_W}\right)^{0.4}}$$

Fluid Flow Projects

Modeling …

- Continuity Equation

$$q_{O,T} = A_1(1-H_{D1})v_1 + A_2H_{D2}v_2$$

$$q_{W,T} = A_2(1-H_{D2})v_2 + A_1H_{D1}v_1$$

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Modeling …

- **Combined Momentum Equation**

$$F = -\frac{\tau_{W2} S_2}{A_2} + \frac{\tau_{W1} S_1}{A_1} + \tau_I S_I \left(\frac{1}{A_1} + \frac{1}{A_2} \right) +$$

$$(\rho_1 - \rho_2) g \sin(\theta) = 0$$

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Modeling …

- **Interfacial Shear Stress**

$$\tau_I = \left[\rho_{MIX} C_f (\tau_1 H_1 + \tau_2 H_2) \right]^{1/2} (v_1 - v_2)$$

$$\rho_{MIX} = (\rho_1 H_1 + \rho_2 H_2)$$

$$C_f = \left(\frac{f_1 H_1 + f_2 H_2}{2} \right)$$

$$H_1 = \left(\frac{A_1}{A_1 + A_2} \right) \qquad H_2 = \left(\frac{A_2}{A_1 + A_2} \right)$$

Fluid Flow Projects

Modeling …

- **Total Potential Energy**

$$PE = A_1\rho_1 g h_1 + A_2\rho_2 g h_2$$

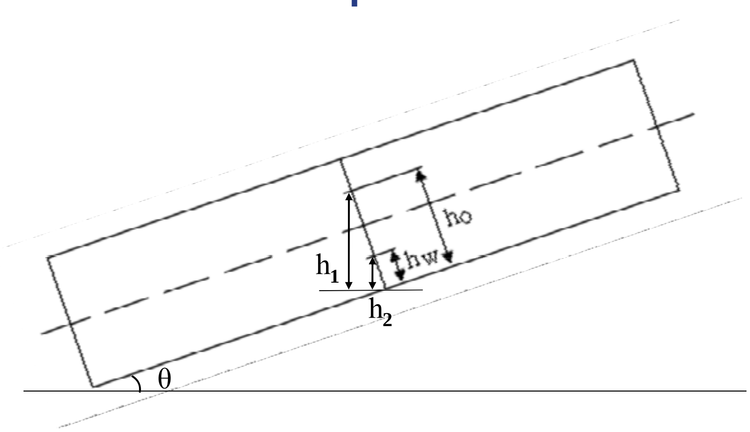
$$h_1 = h_o\cos(\theta)$$

$$h_2 = h_w\cos(\theta)$$

$$h_o = \frac{d}{2}\left[1+\frac{4}{3}\frac{\sin^3\left(\frac{\beta}{2}\right)}{(2\pi-\beta-\sin(\beta))}\right] \qquad h_w = \frac{d}{2}\left[1-\frac{4}{3}\frac{\sin^3\left(\frac{\beta}{2}\right)}{(\beta-\sin(\beta))}\right]$$

Modeling …

- **Side View of Pipe**



Modeling …

- Total Kinetic Energy

$$KE = \frac{1}{2} A_1 \rho_1 v_1^2 + \frac{1}{2} A_2 \rho_2 v_2^2$$

- Total Surface Energy

$$SE = \sigma \left[d \ sin\left(\frac{\beta}{2}\right) + \frac{6 \ A_1 H_{D1}}{d_{SM1}} + \frac{6 \ A_2 H_{D2}}{d_{SM2}} \right]$$

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Modeling …

- Total Energy

$$TE = SE + KE + PE$$

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Modeling …

- **Pressure Gradient Calculation**
  - **Segregated Flow Patterns**
    - **Two Fluid Model (Trallero Oil-Water, 1995)**
  - **Fully Dispersed Flow Patterns**
    - **Homogeneous Model**

Modeling …

- **Homogeneous Model**

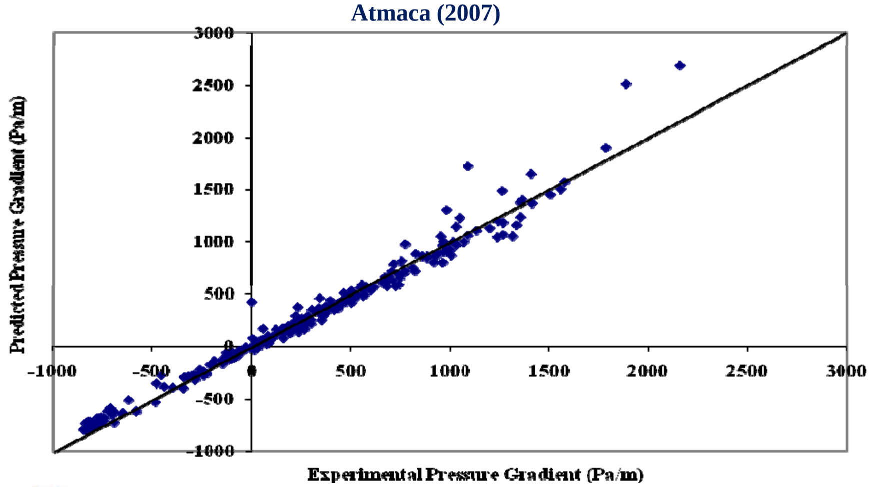
$$\left(\frac{dP}{dL}\right) = \left(\frac{dP}{dL}\right)_{Friction} + \left(\frac{dP}{dL}\right)_{Gravitational}$$

$$\left(\frac{dP}{dL}\right)_{Friction} = \frac{2 f_M \rho_M v_M^2}{d}$$

$$\left(\frac{dP}{dL}\right)_{Gravitational} = \rho_M g \sin(\theta)$$

Results

Model Predictions of Pressure Gradient Compared with Experimental Data of Atmaca (2007)



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Results …

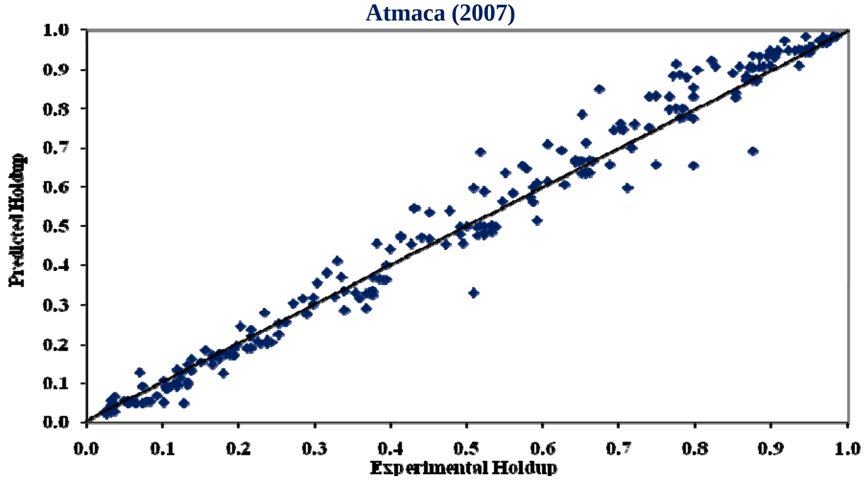
- Pressure Gradient Comparison with Unified Model for Atmaca (2007) Data

	ε_1 (%)	ε_2 (%)	ε_3 (%)	ε_4 (Pa/m)	ε_5 (Pa/m)	ε_6 (Pa/m)
Unified Model	2.5	20.6	40.7	41.5	63.6	99.4
Current Model	-7.0	14.8	30.4	-0.4	49.4	86.9

Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Results …

Model Predictions of Holdup (1º, 2º,-1º, -2º, -5º) compared with Experimental Data of Atmaca (2007)



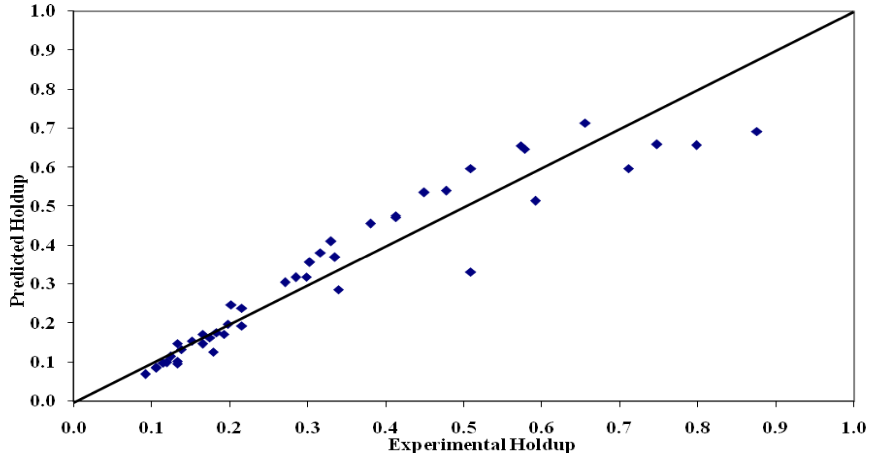
Results …

- Holdup Comparison With Unified Model for Atmaca (2007) Data

	ε_1 (%)	ε_2 (%)	ε_3 (%)	ε_4	ε_5	ε_6
Unified Model	-2.94	8.93	13.21	-0.02	0.03	0.04
Current Model	0.47	9.73	15.58	0.01	0.03	0.05

Results …

Model Predictions of Holdup for Stratified Flow Pattern (1º, 2º,-1º, -2º, -5º) compared with Experimental Data of Atmaca (2007)

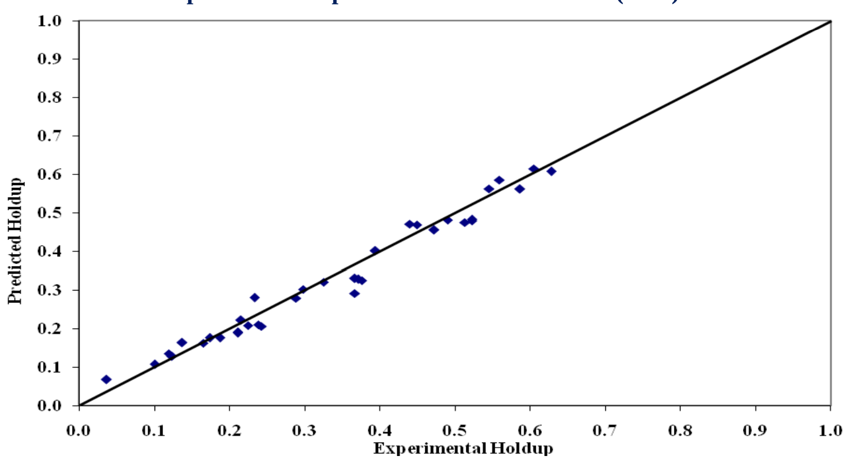


Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Results …

Model Predictions of Holdup for Dual Dispersion Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)



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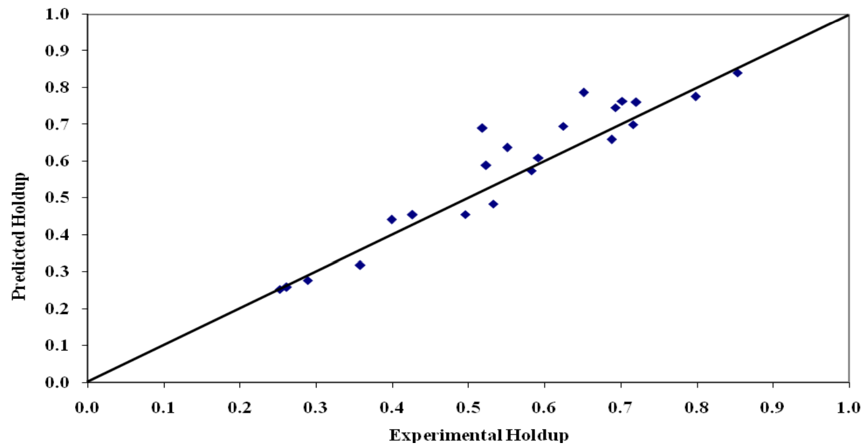
Results …

Model Predictions of Holdup for Oil in Water Dispersion and Oil Layer Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)

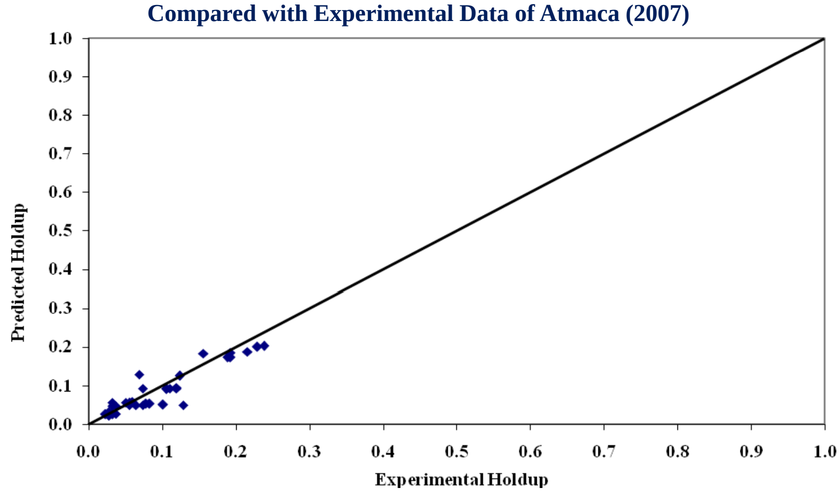


Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Results …

Model Predictions of Holdup for Water in Oil Dispersion Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)

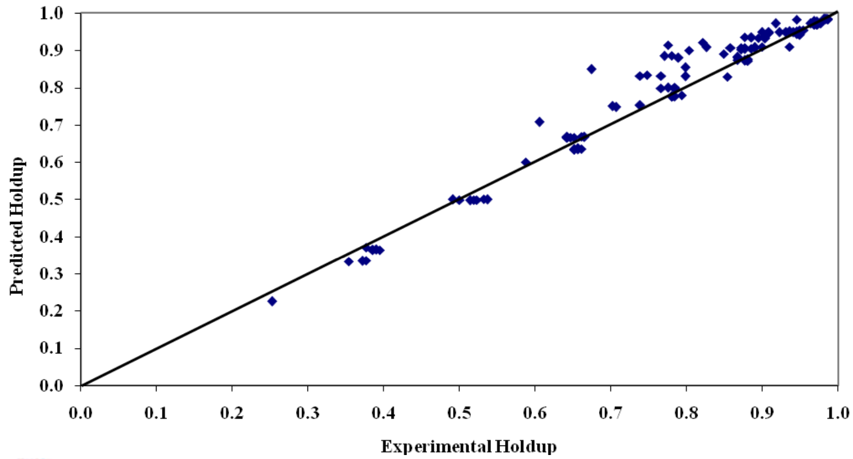


Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Results …

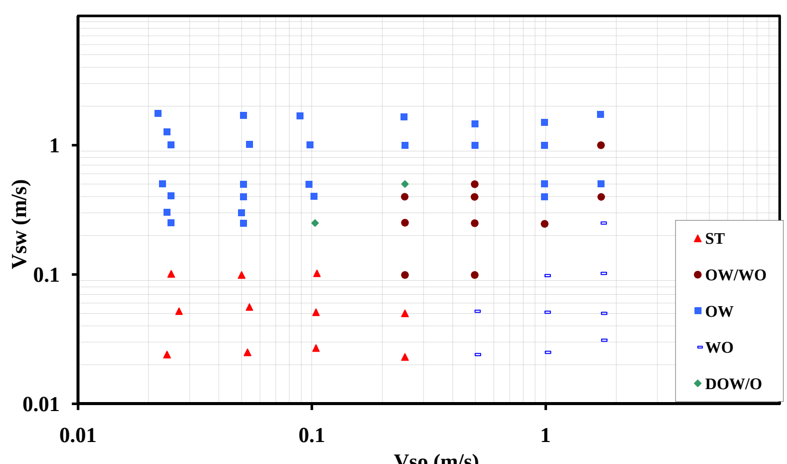
Model Predictions of Holdup for Oil in Water Dispersion Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)



Fluid Flow Projects   Advisory Board Meeting, March 25, 2009

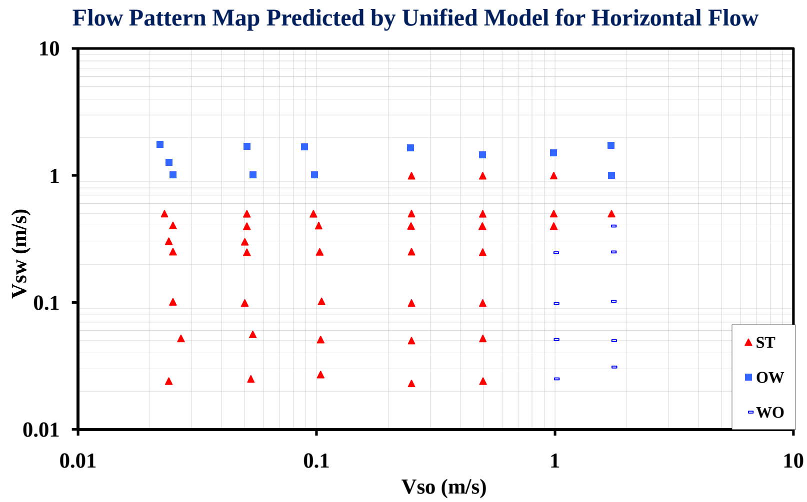
Results …

Flow Pattern Map Predicted by Present Model for Horizontal Flow



Fluid Flow Projects   Advisory Board Meeting, March 25, 2009

Results …



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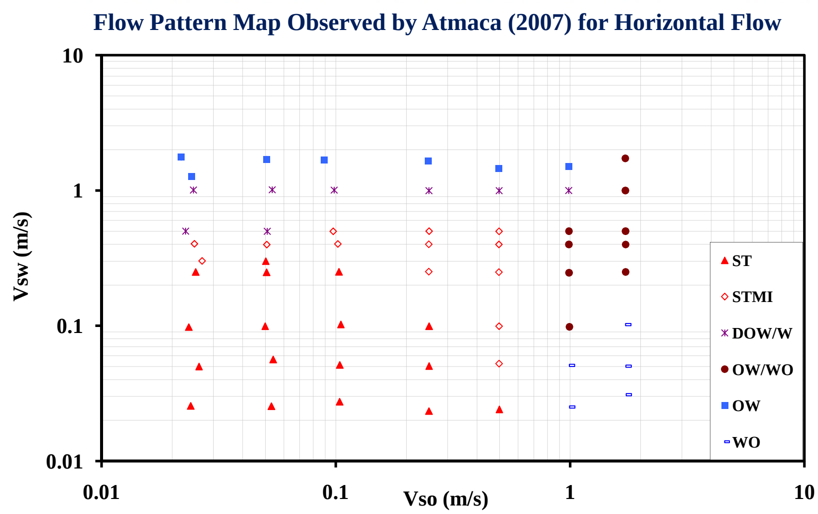
Results …



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Conclusions

- Results are Encouraging
- Useful for Modeling Near Horizontal Flow
- Captures Gradual Changes in Flow Configuration
- Model Gives Extensive Information
- Sensitive to Closure Relationships
  - Droplet Size Correlations
  - Viscosity Correlation

Schedule

Literature Review	Completed
Model Development	Completed
Model Validation	Ongoing
Final Report and Thesis	May 2009

Questions & Comments

Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Modeling of Hydrodynamics and Dispersions in Oil-Water Pipe Flow

Anoop Kumar Sharma

PROJECT COMPLETETION DATES:

Literature Review..Completed
Model Development…...Completed
Model Validation…………...…………………….…………..Ongoing
Final Report and Thesis……………………...…………………………....…………..…..May 2009

Objectives

The objectives of this study are:

- to develop a better model for oil-water flow which captures the physics behind the process, especially for transitions between flow patterns, and
- to validate the model using available experimental data.

Introduction

The flow of two immiscible liquids is encountered in a diverse range of processes and equipment, particularly in the petroleum industry, where mixtures of oil and water are often transported in pipes over long distances. Accurate prediction of oil-water flow characteristics, such as flow pattern, water holdup and pressure gradient, is important in many engineering applications. However, despite their importance, liquid-liquid flow has not been explored to the same extent as gas-liquid flow. The density difference between the phases in a liquid-liquid system is relatively small. However, the viscosity ratio encountered can extend over several orders of magnitude. Moreover, oils and oil-water emulsions can show either a Newtonian or non-Newtonian rheological behavior. Therefore, concepts of gas-liquid two-phase flow cannot be readily applied to liquid-liquid systems.

Existing models are based on first predicting flow patterns and then calculation of design parameters such as pressure gradient and holdup. This approach results in artificially abrupt changes and forces the flow to conform to a particular flow pattern. This study focuses on facilitating gradual changes in flow configuration, therefore resulting in better predictions of the design parameters.

Model Development

A preliminary model was presented at the last ABM. Challenges with the turbulent energy and surface energy relationships were also mentioned. This closure relationship was crucial to the model. Therefore, an in depth analysis of turbulent energy and surface energy relationships was conducted. A rigorous attempt was made to estimate the threshold turbulent energy at the onset of entrainment, which was the key for the closure relationship. Because of the lack of experimental data, it was difficult to estimate this effectively. After realizing the hurdle, a different approach is used for modeling. The new approach is being based on the minimization of energy. It is simple and produced encouraging results.

The new model is based on the principle that a system stabilizes to its minimized total energy, including the fluid flowing in the pipe. Chakrabarti et al. (2005) used a similar approach for pressure prediction for a horizontal, segregated flow pattern. In this model, for each inlet condition, the total energy is minimized. Moreover, for the segregated flow pattern, the continuity equation and combined momentum balance equation are also solved. For full dispersion conditions, only total energy is minimized and the continuity equation is solved.

For all flow patterns under consideration, including dispersion, the mixture properties for each phase can be calculated as following:

$$\rho_1 = \rho_o\left(1 - H_{D1}\right) + \rho_w H_{D1}, \tag{1}$$

$$\rho_2 = \rho_w\left(1 - H_{D2}\right) + \rho_o H_{D2}, \tag{2}$$

$$\mu_1 = \mu_w\left(1 - H_{D1}\right)^{-2.5}, \tag{3}$$

$$\mu_2 = \mu_w\left(1 - H_{D2}\right)^{-2.5}. \tag{4}$$

Where, subscripts 1 and 2 represent layer 1 (oil continuous phase) and layer 2 (water continuous phase), respectively. Subscripts D, o, w represents dispersed phase, pure oil phase and pure water phase, respectively. H, μ and ρ represents the holdup, viscosity and density, respectively. Equations 3 and 4 use the Brinkman viscosity correlation to calculate the viscosity of dispersions. Although the model uses the Brinkman viscosity correlation, other mixture viscosity correlations can also be used. H_{D1} and H_{D2} will vary from 0 to the phase inversion point. The inversion point in this model is calculated by using Eq. 5, as used by Zhang et al. (2003).

$$H_{O,INV} = \frac{\left(\dfrac{\mu_O}{\mu_W}\right)^{0.4}}{\left(1 + \dfrac{\mu_O}{\mu_W}\right)^{0.4}}. \tag{5}$$

Where $H_{O,INV}$ is the holdup of the dispersed phase (oil) in an oil-water dispersion at which inversion takes place. The velocity of each layer can be calculated by simultaneously solving continuity equations for both oil and water.

$$q_{O,T} = A_1\left(1 - H_{D1}\right)v_1 + A_2 H_{D2} v_2, \tag{6}$$

$$q_{W,T} = A_2\left(1 - H_{D2}\right)v_2 + A_1 H_{D1} v_1. \tag{7}$$

Where, $q_{O,T}$, $q_{W,T}$, v_1 and v_2 are the total inlet flow rate of oil and water and velocities of the oil continuous and the water continuous phases, respectively. A_1 and A_2 are areas corresponding to oil continuous and water continuous phases as shown in Fig. 1, respectively.

Trallero (1995) presented a two-fluid model to predict pressure drop in two-phase segregated flow. The two-fluid model solves the combined momentum equation to predict the pressure drop. Assuming smooth, equilibrium, horizontal stratified flow, the following momentum balance equations can be derived for each phase (phases 1 and 2):

$$-A_1\left(\frac{dP}{dL}\right)_1 - \tau_{W1}S_1 \pm \tau_I S_I - \rho_1 A_1 g \sin(\theta) = 0. \tag{8}$$

$$-A_2\left(\frac{dP}{dL}\right)_2 - \tau_{W2}S_2 \pm \tau_I S_I - \rho_2 A_2 g \sin(\theta) = 0. \tag{9}$$

$$F = -\frac{\tau_{W2}S_2}{A_2} + \frac{\tau_{W1}S_1}{A_1} + \tau_I S_I\left(\frac{1}{A_1} + \frac{1}{A_2}\right) +$$

$$(\rho_1 - \rho_2) g \sin(\theta) = 0. \tag{10}$$

Where, A, S and $\left(\dfrac{dP}{dL}\right)$ denote area, perimeter and pressure gradient, respectively. F represents the combined momentum. Subscripts 1, 2 and I represent the oil continuous phase, water continuous phase and the interface, respectively. θ is the inclination angle from horizontal. $\tau_{W1}, \tau_{W2}, \tau_I$ are oil, water and interfacial shear stresses, respectively. These can be expressed in terms of the corresponding fluid friction factors f_1, f_2 and f_I.

$$\tau_{W1} = \frac{f_1 \rho_1 v_1^2}{2}. \tag{11}$$

$$\tau_{W2} = \frac{f_2 \rho_2 v_2^2}{2}. \tag{12}$$

The fanning friction factor can be expressed for any phase j, assuming a smooth pipe wall.

$$f_j = C\left(\frac{d_j v_j \rho_j}{\mu_j}\right)^{-n}. \quad (13)$$

Where, coefficient C and exponent n are equal to 16 and 1 for laminar flow and to 0.046 and 0.2 for turbulent flow. Equivalent hydraulic diameters are determined on the basis of which phase is faster.

For $v_1 > v_2$,

$$d_1 = \frac{4A_1}{S_1 + S_I}, \quad (14)$$

$$d_2 = \frac{4A_2}{S_2}. \quad (15)$$

For $v_1 < v_2$,

$$d_1 = \frac{4A_1}{S_1}, \quad (16)$$

$$d_2 = \frac{4A_2}{S_2 + S_I}. \quad (17)$$

For $v_1 \approx v_2$,

$$d_1 = \frac{4A_1}{S_1}, \quad (18)$$

$$d_2 = \frac{4A_2}{S_2}. \quad (19)$$

In the model, the interfacial shear stress (τ_I) is calculated using Eq. 20, as proposed by Zhang et al. (2005).

$$\tau_I = \left[\rho_{MIX} C_f \left(\tau_1 H_1 + \tau_2 H_2\right)\right]^{1/2}\left(v_1 - v_2\right). \quad (20)$$

$$\rho_{MIX} = \left(\rho_1 H_1 + \rho_2 H_2\right). \quad (21)$$

$$C_f = \left(\frac{f_1 H_1 + f_2 H_2}{2}\right). \quad (22)$$

$$H_1 = \left(\frac{A_1}{A_1 + A_2}\right). \quad (23)$$

$$H_2 = \left(\frac{A_2}{A_1 + A_2}\right). \quad (24)$$

h_o and h_w shown in Fig. 1 are the centroid of the oil continuous phase and water continuous phase from the pipe base, respectively:

$$\mathrm{h_o} = \frac{\mathrm{d}}{2}\left[1 + \frac{4}{3}\frac{\sin^3\left(\frac{\beta}{2}\right)}{(2\pi - \beta - \sin(\beta))}\right]. \quad (25)$$

$$\mathrm{h_w} = \frac{\mathrm{d}}{2}\left[1 - \frac{4}{3}\frac{\sin^3\left(\frac{\beta}{2}\right)}{(\beta - \sin(\beta))}\right]. \quad (26)$$

For inclined pipe, the height of the centroids of oil and water continuous phases from the bottom, as shown in Fig. 2, will be

$$\mathrm{h_1} = \mathrm{h_o}\cos(\theta). \quad (27)$$

$$\mathrm{h_2} = \mathrm{h_w}\cos(\theta). \quad (28)$$

The total pressure energy per unit length of pipe can be calculated as:

Total Potential Energy (PE) = PE of water continuous phase + PE of oil continuous phase (29)

$$PE = A_1 \rho_1 g h_1 + A_2 \rho_2 g h_2. \quad (30)$$

The total kinetic energy per unit length of pipe can be calculated as:

Total Kinetic Energy (KE) = KE of water continuous phase + KE of oil continuous phase (31)

$$KE = \frac{1}{2} A_1 \rho_1 v_1^2 + \frac{1}{2} A_2 \rho_2 v_2^2. \quad (32)$$

The total surface energy per unit length can be calculated as:

Total Surface Energy (SE) = SE of interface + SE of water droplets in oil phase + SE of oil droplets in water phase (33)

$$SE = \sigma \left[\mathrm{d} \sin\left(\frac{\beta}{2}\right) + \frac{6\,\mathrm{A}_1 \mathrm{H}_{\mathrm{D1}}}{\mathrm{d}_{\mathrm{SM1}}} + \frac{6\,\mathrm{A}_2 \mathrm{H}_{\mathrm{D2}}}{\mathrm{d}_{\mathrm{SM2}}} \right]. \quad (34)$$

Where, σ, d_{SM1} and d_{SM2} are the interfacial tension between oil and water, Sauter Mean Diameter (SMD) of the water droplets and SMD of the oil droplets. SMD can be estimated by several different models and correlations. In this model, the correlations of Angeli and Hewitt (2000) are used to estimate Maximum Diameter (d_{MAX}) and SMD (d_{SM}).

$$d_{SM} v_C^{1.8} = 2 \times 10^{-8} \times (4 f_C)^{-3.12}. \quad (35)$$

$$d_{MAX} v_C^{1.8} = 4.2 \times 10^{-8} \times (4 f_C)^{-3.13}. \quad (36)$$

Where, v_C and f_C are continuous phase velocity and friction factor (Fanning), respectively.

Finally, the total energy of the system is:

$$TE = SE + KE + PE. \quad (37)$$

The pressure gradient is calculated by the two-fluid model for all flow patterns in which two distinct phases are present in the pipe. For fully dispersed flow, the pressure gradient is calculated from a homogeneous model as follows,

$$\left(\frac{dP}{dL}\right) = \left(\frac{dP}{dL}\right)_{Friction} + \left(\frac{dP}{dL}\right)_{Gravitational}. \quad (38)$$

$$\left(\frac{dP}{dL}\right)_{Friction} = \frac{2 f_M \rho_M v_M^2}{d}. \quad (39)$$

$$\left(\frac{dP}{dL}\right)_{Gravitational} = \rho_M g \sin(\theta). \quad (40)$$

where, f_M, ρ_M and v_M are the friction factor of the mixture, mixture density and mixture velocity, respectively.

The model first takes all the inputs, including fluid properties, superficial velocities of the oil and water, etc. The model then varies β from 0º to 360º and calculates TE and F for the system. The solution is the condition where the total energy is minimum (*TE → Minimum*) and combined momentum equation is near zero (*F→0*) simultaneously. For the two cases of full dispersions, i.e. β = 0º (dispersion of water in oil) and β = 360º (dispersion of oil in water), the model does not consider the combined momentum equation (*F=0*) and only considers the minimization of total energy. For Reynolds number smaller than 1000, it is assumed that the flow will be laminar and the continuous phase will have a negligible amount of turbulent energy, if any, to cause any dispersion. Hence, the continuous phase remains in segregated form. To determine whether there is any dispersion for Reynolds number higher then 1000, d_{MAX} is calculated and compared with the maximum distance (L_{MAX}) between the interface and pipe wall across the continuous phase, as shown in Fig. 3.

$$d_{MAX} \le L_{MAX} \quad (41)$$

Moreover, to avoid inversion, the maximum cross section area swept by the biggest droplet should not cross the inversion point of dispersed phase holdup in the continuous phase.

$$\left(\frac{A_{drop}}{A}\right) \le H_{INV}. \quad (42)$$

where, A_{drop}, A and H_{INV} represents the cross section al area of the drop, cross sectional area of the layer where the drop exists and the holdup of the dispersed phase at the inversion point below which the particular droplet dispersion is stable, respectively. These assumptions are very reasonable, conservative and biased towards the dispersions. If either of the conditions is not satisfied then it is not possible to have dispersions and the continuous phase still remains in segregated form. If the continuous phase can accommodate d_{MAX} , then the hold-up of the dispersed phase (H_D) is varied from 0 to the phase inversion point and for each case both *TE* and *F* are minimized. This is done simultaneously in both layers.

Results and Discussions

A VBA code was written for the model. Results obtained are compared with experimental results obtained by Atmaca (2007) and are very encouraging. In Fig. 4, the predicted pressure gradients are compared with the experimental pressure gradients, for near-horizontal inclinations (0º, 1º, 2º, -1º, -2º, -5º). In Fig. 5, water holdup predictions of the current model are compared with the Atmaca (2007) experimental data for inclinations 1º, 2º, -1º, -2º and -5º. Figures 6-10 show water holdup predictions by the present model compared with the Atmaca (2007) experimental data (1º, 2º, -1º, -2º and -5º) for different flow patterns. Statistical evaluation parameters are defined in Table 1 and error analyses of both the TUFFP unified model and the current model is given in Tables 2 and 3 for pressure gradient and water holdup prediction, respectively. Figures 11 and 12 show the flow pattern maps predicted by the current model and the TUFFP unified model for horizontal flow, respectively. In Fig. 13, the flow pattern map observed by Atmaca (2007) for horizontal flow is shown. In Figs. 11-13, ST, STMI, DOW/W, DOW/O, O/W and W/O denote stratified, stratified mixing, dispersion of oil in water with water layer, dispersion of oil in water with oil layer, dispersion of oil in water and dispersion of water in oil, respectively.

The current model performs very well in pressure gradient predictions and holdup predictions. Its performance is more accurate for the pressure gradient predictions, than the TUFFP unified model. For the holdup predictions it is on par with the TUFFP unified model. The flow pattern maps are showing general flow patterns trend very well. The flow pattern map heavily depends on the droplet size correlations and how we are defining the criteria for onset of entrainment.

Conclusions

It is evident from the results and comparisons that the model estimates the pressure gradient and flow pattern very well on the basis of the minimization of total energy. The concept of minimization of total energy is very simple, basic and can be applied to any flowing condition in the pipe. It can be a very useful tool for modeling behavior at near-horizontal angles. It gives extensive information about the flow, e.g. pressure gradient, flow pattern (including dual dispersions, single layer dispersions), droplet size information, holdup and dispersed phase holdup for individual layers. Unlike existing oil-water flow models, the current model does not have strict different criterions for different flow pattern transitions. The model doesn't force the flow to conform to a particular flow pattern and captures the gradual changes in flow configuration, therefore resulting in better predictions of the design parameters.

There is ample room to improve model predictions. The model is very sensitive to the closure relationships, especially the droplet size correlations. The mixture viscosity closure relationship is also very crucial for the model. Any improvements in these will be directly reflected in the model's performance.

Near Future Tasks

The main tasks for the future are:

- Validation of the model with more data and experimental results.
- Writing final report and M.S. Thesis.

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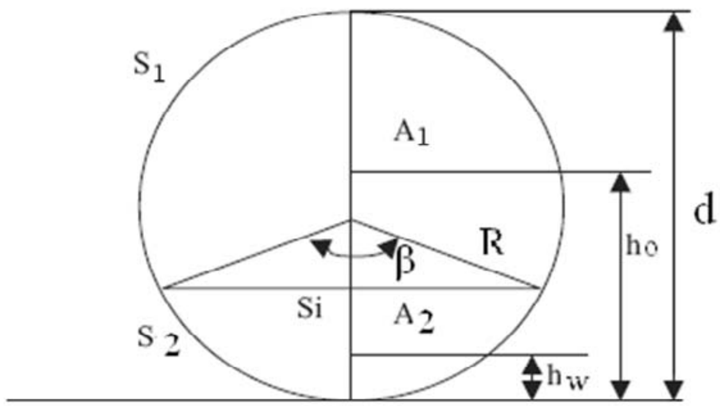


Figure 1: Cross-Sectional View of Pipe

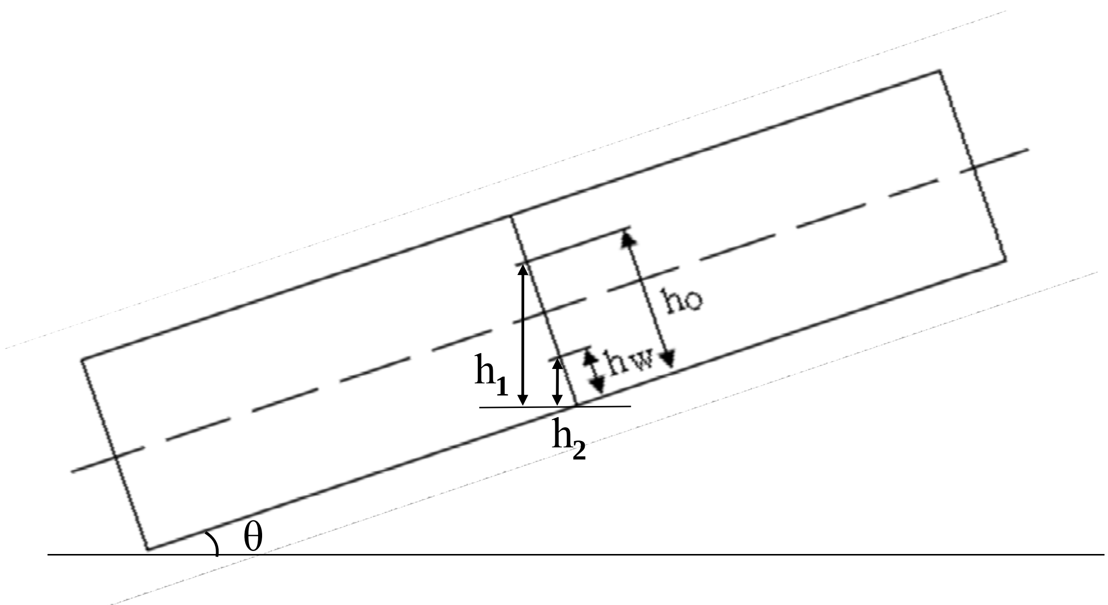
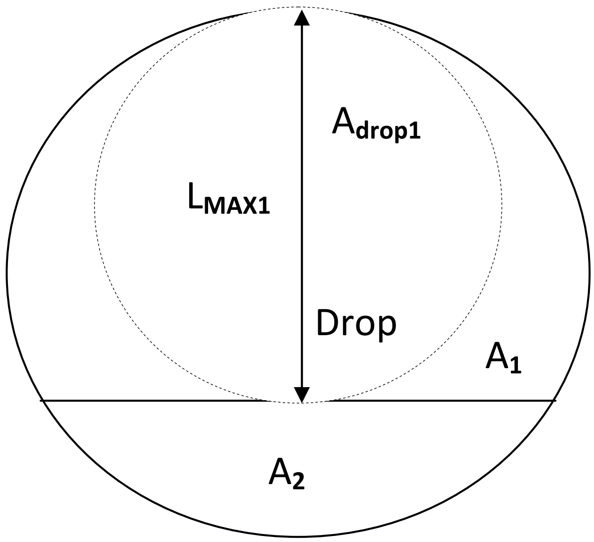


Figure 2: Side View of Pipe



$$\left(\frac{A_{drop1}}{A_1}\right) \le H_{INV1}$$

$$d_{MAX1} \le L_{MAX1}$$

Figure 3: Transition Criteria for Onset of Entrainment

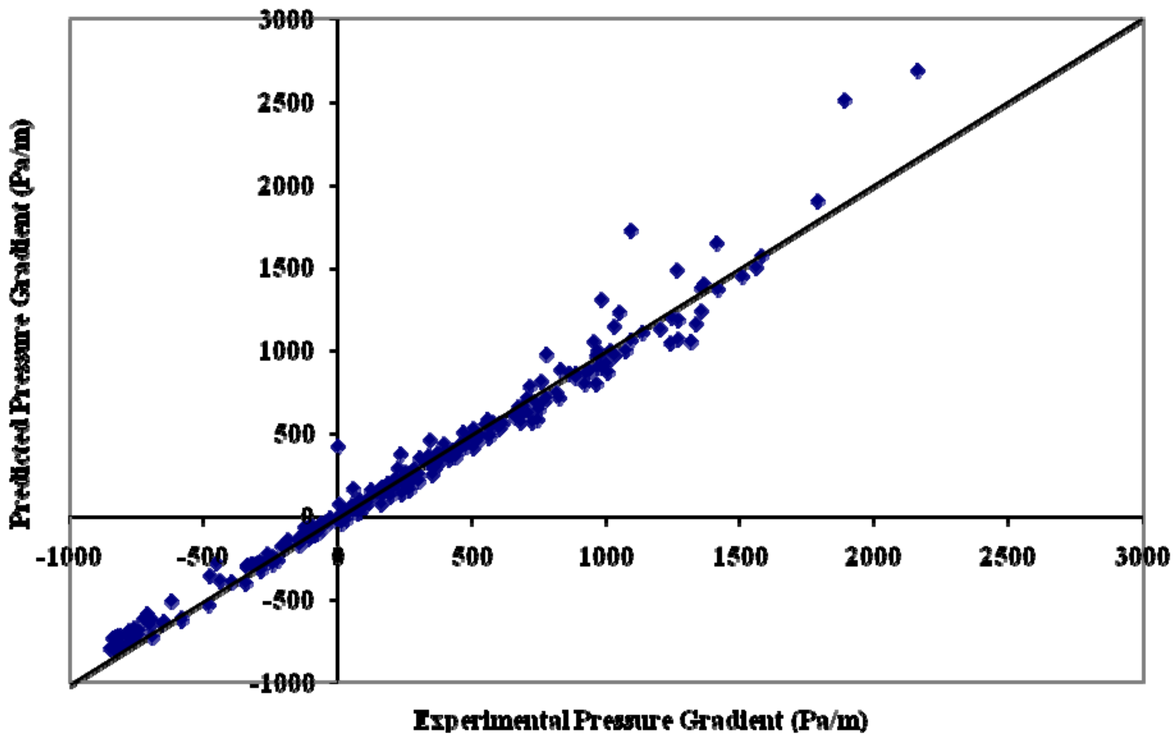


Figure 4: Model Predictions of Pressure Gradient Compared with Experimental Data of Atmaca (2007)

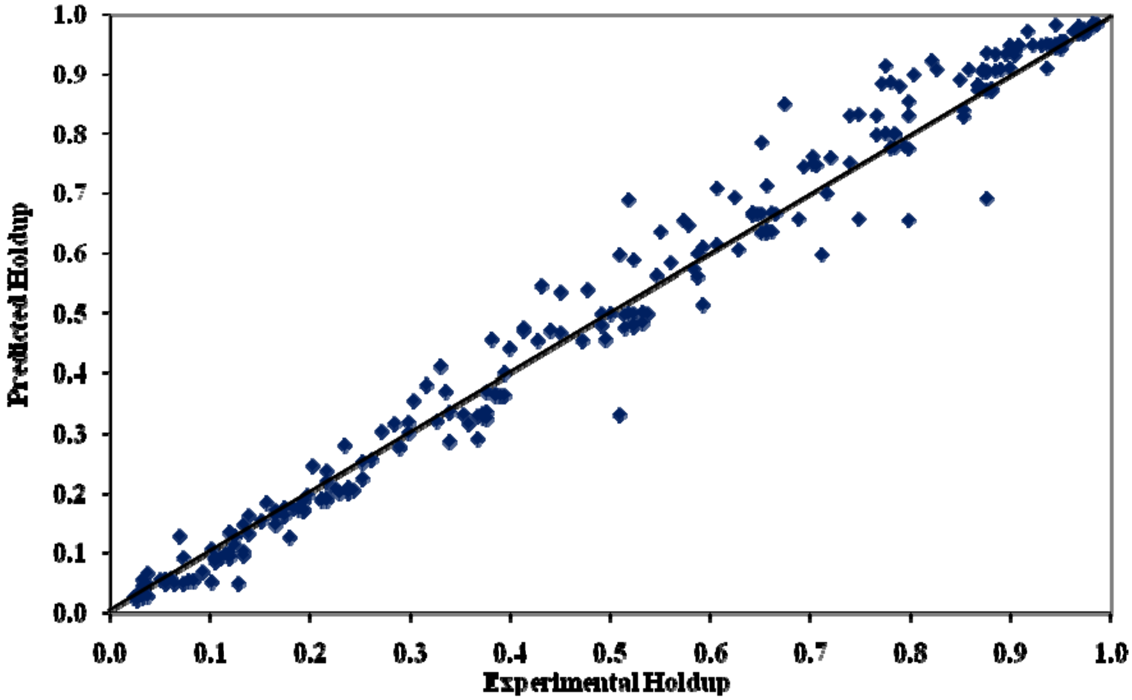


Figure 5: Model Predictions of Holdup (1º, 2º,-1º, -2º, -5º) compared with Experimental Data of Atmaca (2007)

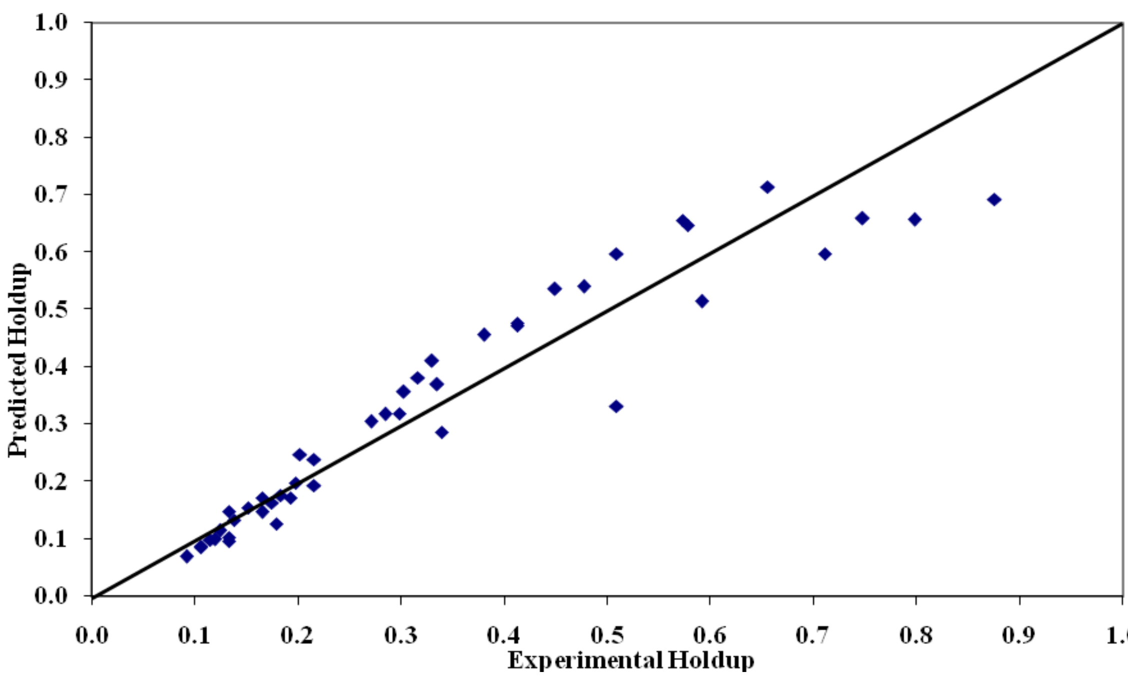


Figure 6: Model Predictions of Holdup for Stratified Flow Pattern (1º, 2º,-1º, -2º, -5º) compared with Experimental Data of Atmaca (2007)

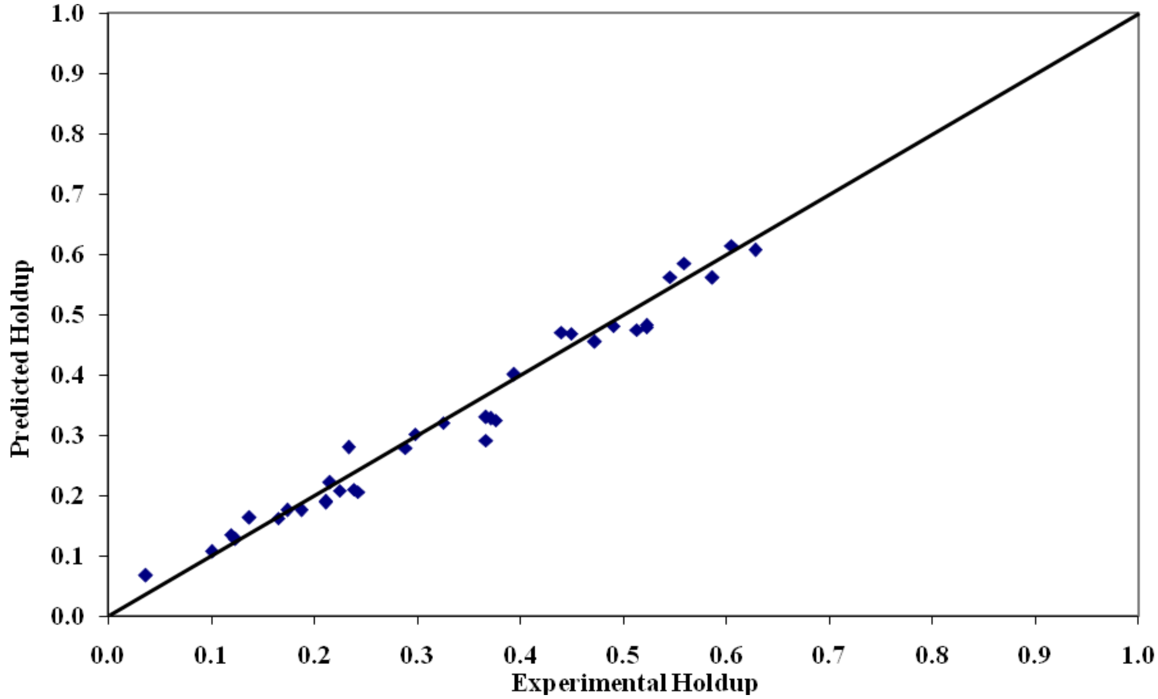


Figure 7: Model Predictions of Holdup for Dual Dispersion Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)

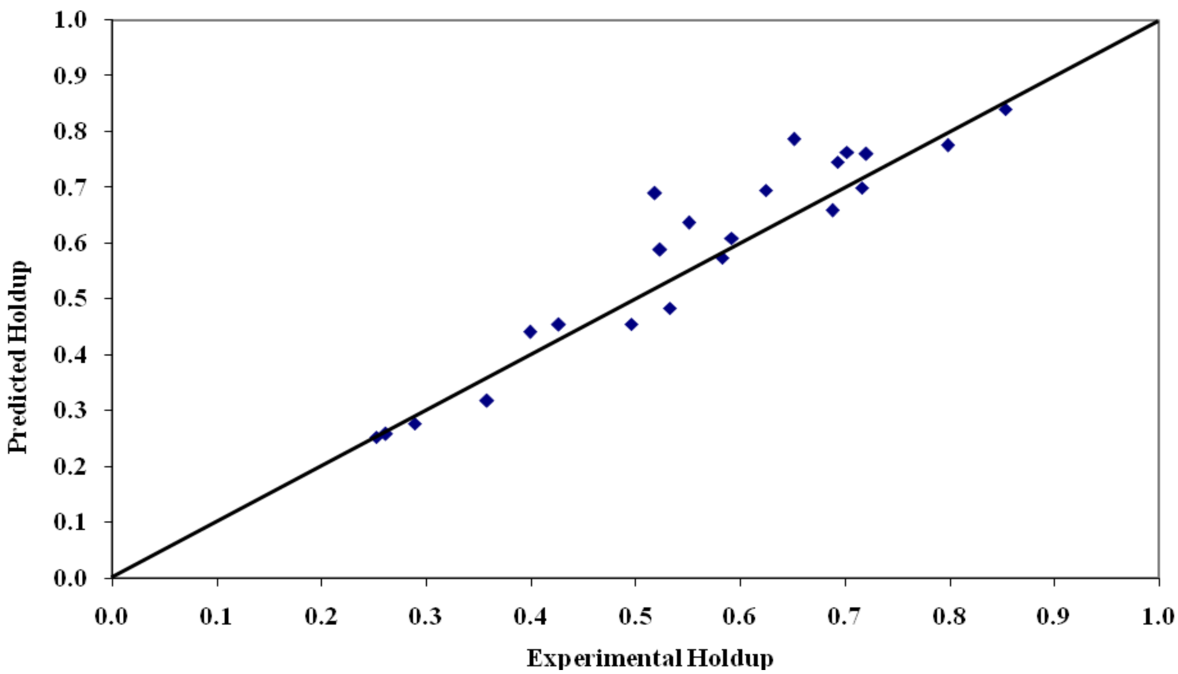


Figure 8: Model Predictions of Holdup for Oil in Water Dispersion and Oil Layer Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)

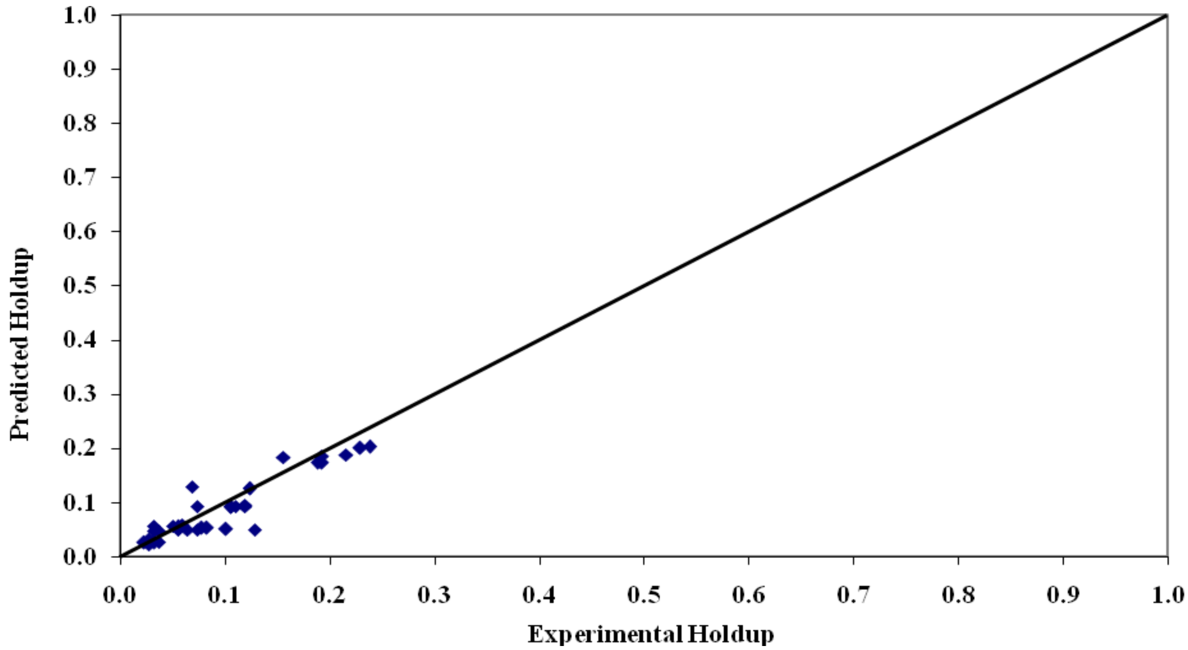


Figure 9: Model Predictions of Holdup for Water in Oil Dispersion Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)

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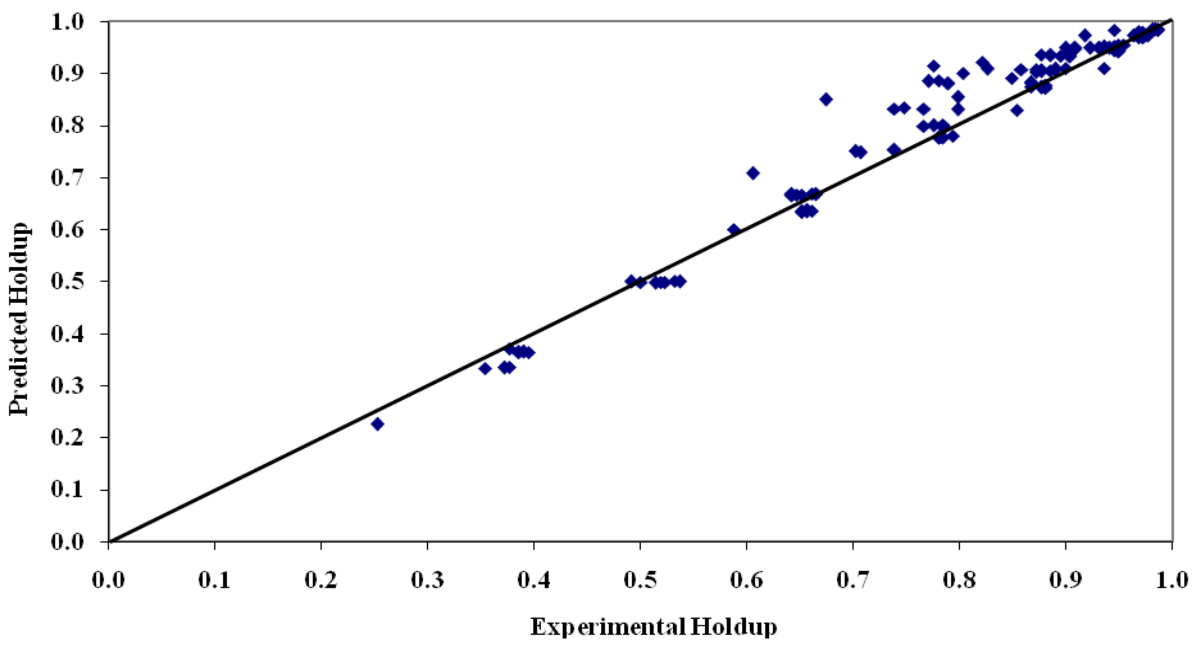

Figure 10: Model Predictions of Holdup for Oil in Water Dispersion Flow Pattern (1º, 2º,-1º, -2º, -5º) Compared with Experimental Data of Atmaca (2007)

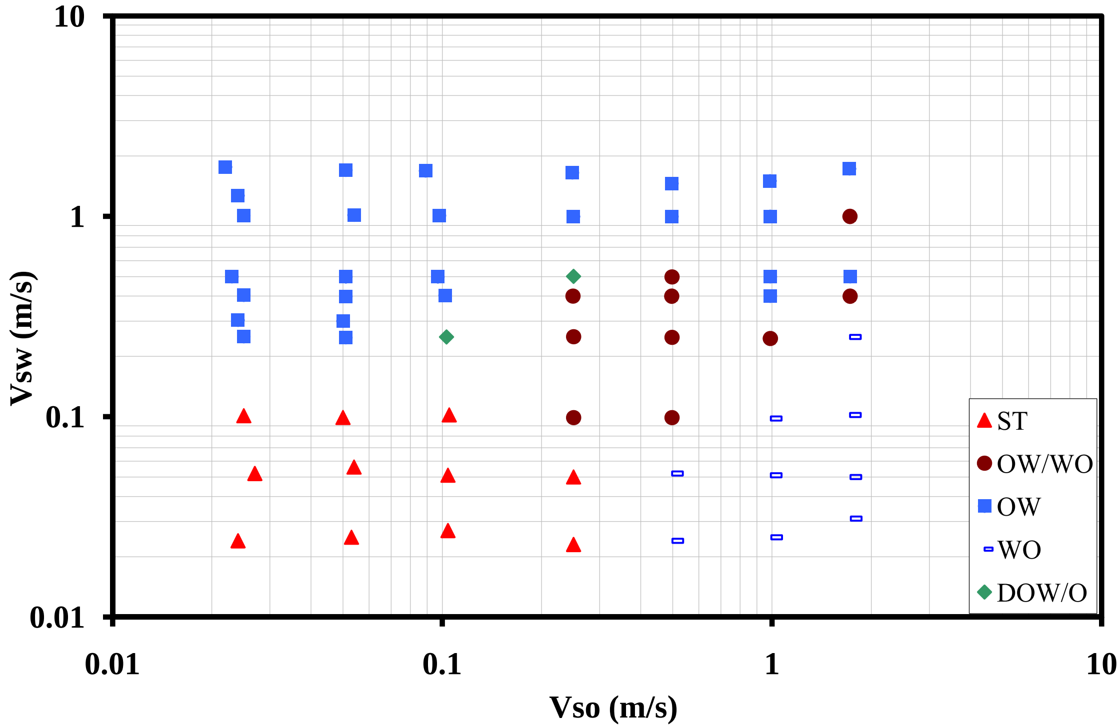

Figure 11: Flow Pattern Map Predicted by Present Model for Horizontal Flow

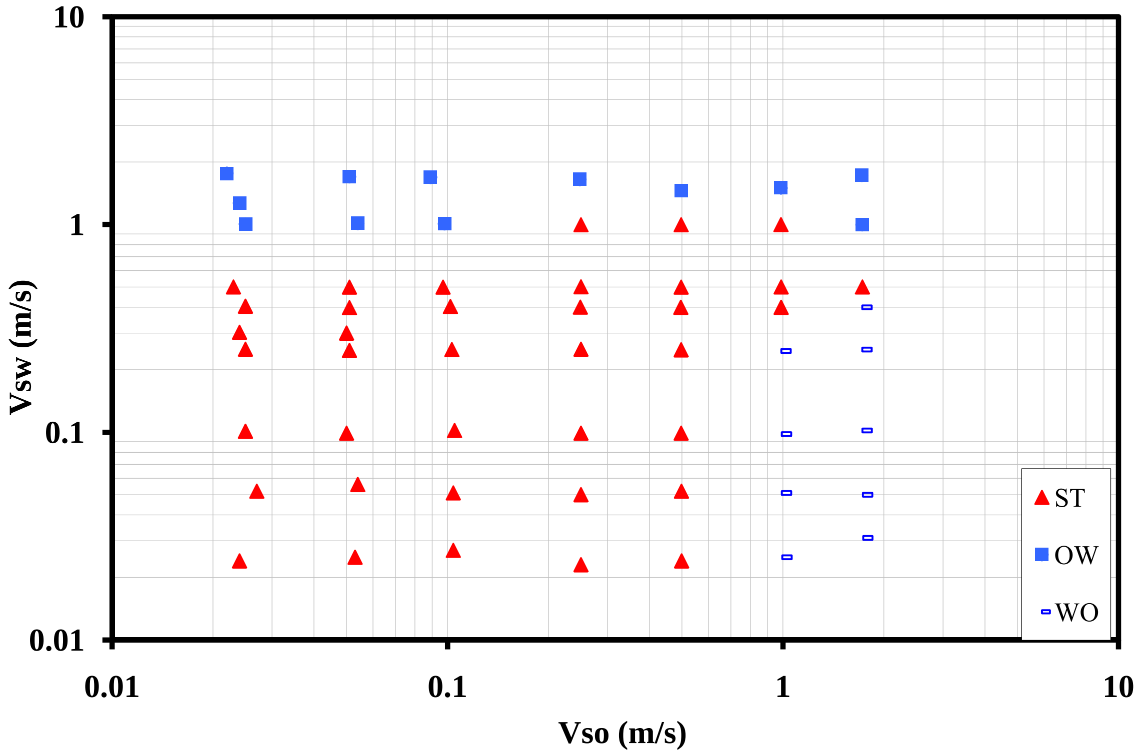

Figure 12: Flow Pattern Map Predicted by TUFFP Unified Model for Horizontal Flow

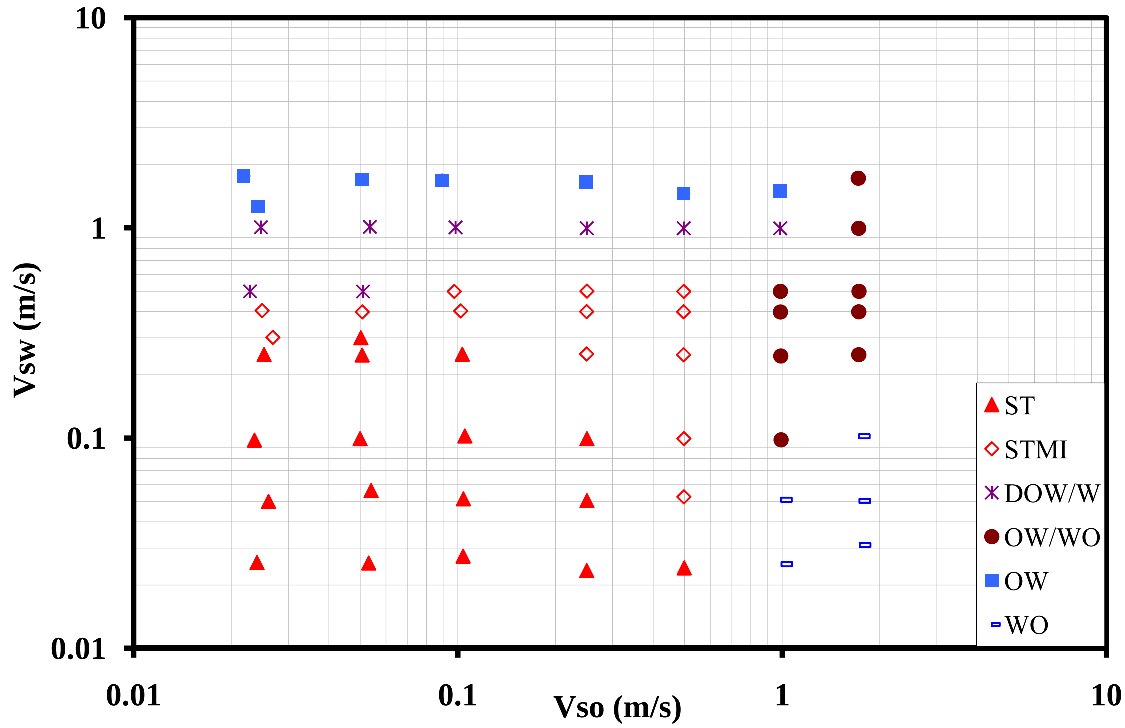

Figure 13: Flow Pattern Map Observed by Atmaca (2007) for Horizontal Flow

Table 1: Statistical Parameters

Definitions
$\varepsilon_1 = \left[\frac{1}{N} \sum_{i=1}^{N} e_{ri} \right]$
$\varepsilon_2 = \left[\frac{1}{N} \sum_{i=1}^{N} \lvert e_{ri} \rvert \right]$
$\varepsilon_3 = \sqrt{\frac{\sum_{i=1}^{N} (e_{ri} - \varepsilon_1)^2}{N-1}}$
$\varepsilon_4 = \frac{1}{N} \sum_{i=1}^{N} e_i$
$\varepsilon_5 = \frac{1}{N} \sum_{i=1}^{N} \lvert e_i \rvert$
$\varepsilon_6 = \sqrt{\frac{\sum_{i=1}^{N} (e_i - \varepsilon_4)^2}{N-1}}$

Table 2: Pressure Gradient Prediction by TUFFP Unified Model and Current Model Compared with Atmaca (2007) Data

	ε_1 (%)	ε_2 (%)	ε_3 (%)	ε_4 (Pa/m)	ε_5 (Pa/m)	ε_6 (Pa/m)
Unified Model	2.5	20.6	40.7	41.5	63.6	99.4
Current Model	-7.0	14.8	30.4	-0.4	49.4	86.9

Table 3: Water Holdup for 1º, 2º,-1º, -2º, -5º Inclinations by TUFFP Unified Model and Current Model Compared with Atmaca (2007) Data

	ε_1 (%)	ε_2 (%)	ε_3 (%)	ε_4	ε_5	ε_6
Unified Model	-2.94	8.93	13.21	-0.02	0.03	0.04
Current Model	0.47	9.73	15.58	0.01	0.03	0.05

Three-phase Hilly Terrain Flow

- Significance
  - Valleys and Hills may Act as Local Separation Devices for Fluids
  - Location, Amount and Residence Time of Water in a Pipe can have Significant Impact on Flow Assurance Issues such as Hydrate Formation and Corrosion

Three-phase Hilly Terrain Flow …

- Past Studies
  - Hilly Terrain Flow of Two Phases has been Studied Extensively
    - Al-Safran, 1999 and 2003
    - Others Outside of TUFFP
  - No Available Research is Found on Three-phase Flow

Three-phase Hilly Terrain Flow …

- Current Project
  - Objectives
    - Observe Flow Behavior and Identify Flow Characteristics
    - Develop Predictive Tools (Closure Relationships or Models)

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Three-phase Hilly Terrain Flow …

- Status
  - Testing is Complete
  - Data Analysis and Model Development are Underway

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Slug Flow Evolution of Gas-Oil-Water Flow in Hilly-Terrain Pipelines

Gizem Ersoy Gokcal

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Outline

- Introduction
- Objectives
- Experimental Study
- Preliminary Modeling
- Project Schedule

Introduction …

Hilly-Terrain Pipelines Cause

- Operational Problems
  - Flooding of Downstream Facilities
  - Severe Pipe Corrosion
  - Structural Instability of Pipelines
- Flow Assurance Problems
  - Hydrates
  - Emulsions
  - Paraffin Deposition
  - Corrosion

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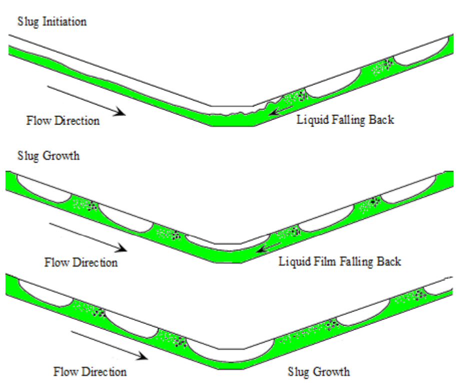
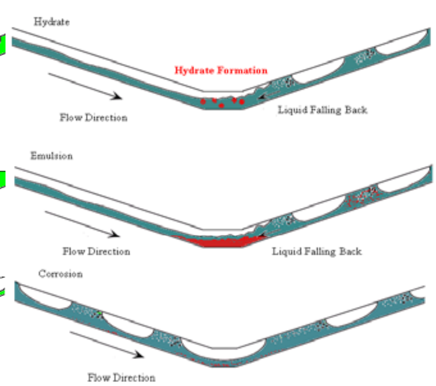
Introduction …

Hydrodynamics



Flow Assurance



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Objectives

- Investigate Gas-Oil-Water Flow in Hilly-Terrain Pipelines
- Develop Closure Models for Flow in Hilly-Terrain Pipelines on
  - Three-Phase Slug Initiation and Dissipation
  - Mixing Status of Phases

Experimental Study

- Experimental Facility
- Instrumentation
- Data Acquisition System
- Test Fluids
- Testing Ranges
- Testing Procedure
- Experimental Results

Experimental Facility

- Extended to 69-m (226-ft) Long
- 50.8-mm (2-in.) ID Pipes
- Single Hilly-Terrain Unit
  - 9.7-m (32-ft) Long Downhill
  - 1.5-m (5-ft) Long Horizontal
  - 9.7-m (32-ft) Long Uphill Sections (L/D=413)
- ±1°, ±2°, ±5° of Inclination Angles

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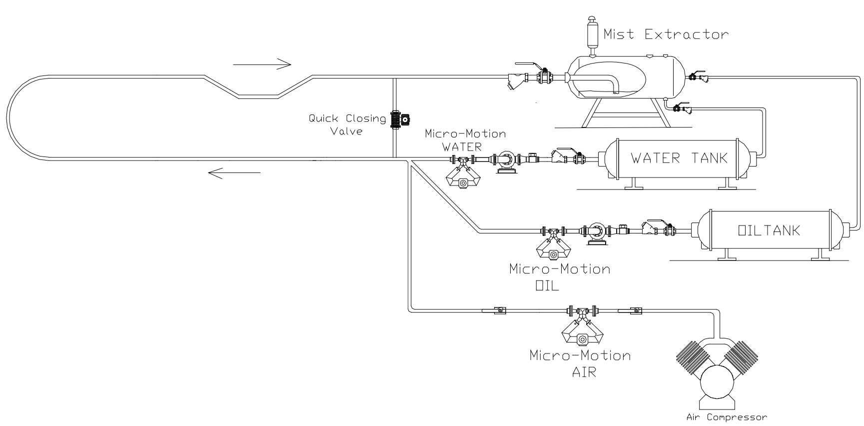
Experimental Facility



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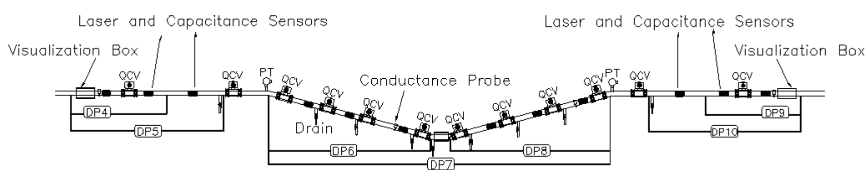
Experimental Facility …

Test Section



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Experimental Facility …

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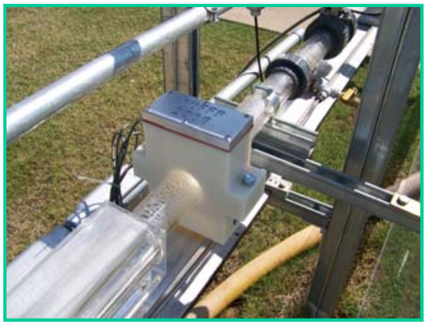
Instrumentation

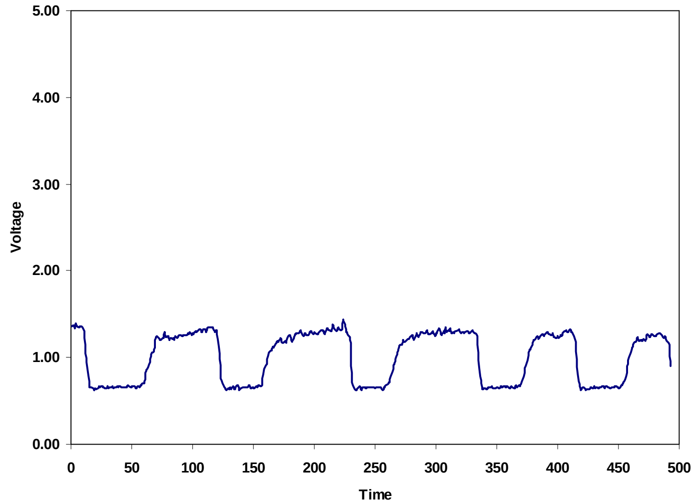
- Pressure & Differential Pressure Transducers
  - Pressure Drop
  - Identification of Flow Patterns
  - Connected to High-Speed DAQ
- Quick-Closing Valves
  - Average Gas, Oil, Water Holdups

Instrumentation …

- Laser Sensors
  - Slug Flow Characteristics
  - Connected to High-Speed DAQ
  - Tested for Three-Phase Slug Flow



Instrumentation …



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Instrumentation …

- Capacitance Sensors
  - Slug Flow Characteristics
  - Connected to High-Speed DAQ
  - Tested for Oil-Water and Three-Phase Slug Flow

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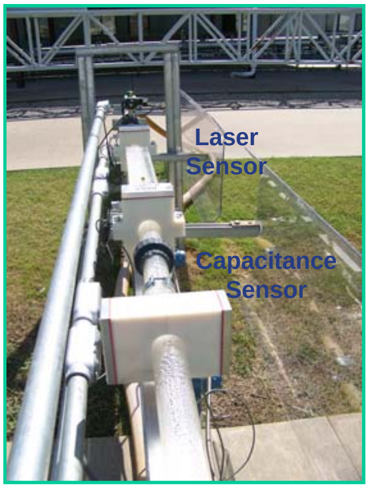
Instrumentation …

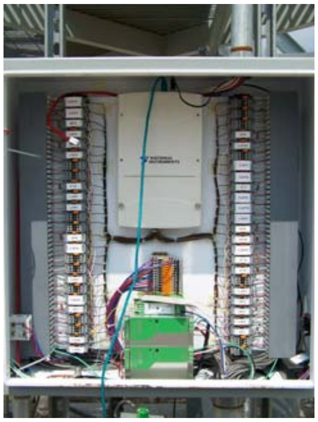


Instrumentation …

- Cameras
  - Identification of Flow Patterns
  - Slug Characteristics
  - Oil-Water Mixing Status
  - Validation of Laser and Capacitance Sensors

Data Acquisition System

- Lab VIEW™ 7.1 Software
- High-Speed Data Acquisition



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Test Fluids

- Air - Mineral Oil - Water
- Tulco Tech-80 Mineral Oil
  - API: 33.2°
  - Density: 858.75 kg/m^3 @ 15.6 °C (60°F)
  - Viscosity: 13.5 cP @ 40 °C (104 °F)
  - Surface Tension: 29.14 dynes/cm @ 25.1 °C (77.2 °F)

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Preliminary Testing

- Defined Instruments for Measuring Three-Phase Slug Flow Characteristics
  - Laser Sensors
  - Capacitance Sensors
- Modified Capacitance Sensors for Three-Phase Flow
- Tested Laser and Capacitance Sensors for Three-Phase Slug Flow
  - Calibration
  - Determination of Limits
  - Sensitivity Analysis
  - Optimized Locations Based on Preliminary Tests

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Testing Ranges

- Superficial Oil Velocity
  - 0.04 – 1 m/s
- Superficial Water Velocity
  - 0.025 – 1 m/s
- Superficial Gas Velocity
  - 0.1 – 5 m/s
- Water Fraction
  - 20%, 40%, 60%, 80%
  - 0% and 100% for Preliminary Tests
- Hilly-Terrain Unit
  - 5° for Valley Configuration

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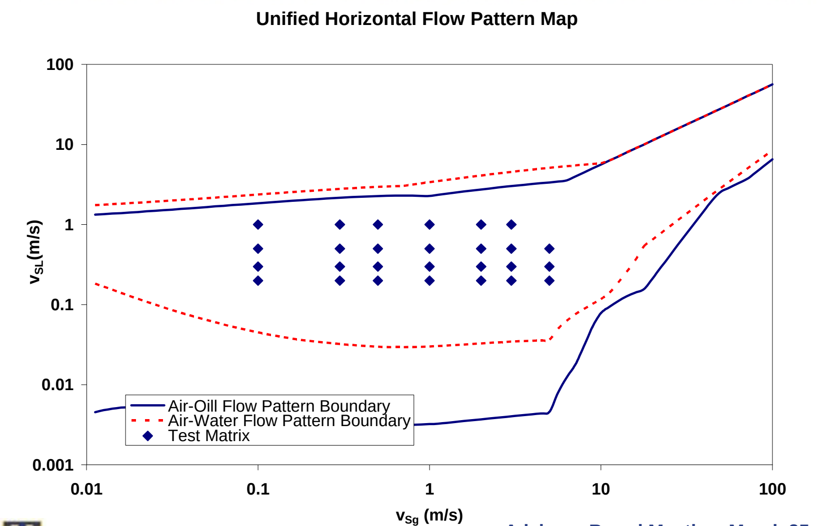
Testing Ranges …



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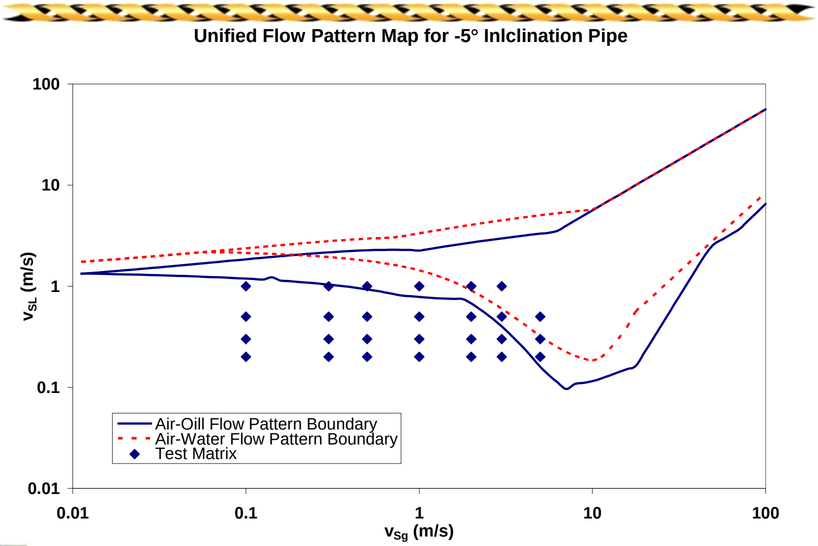
Testing Ranges …



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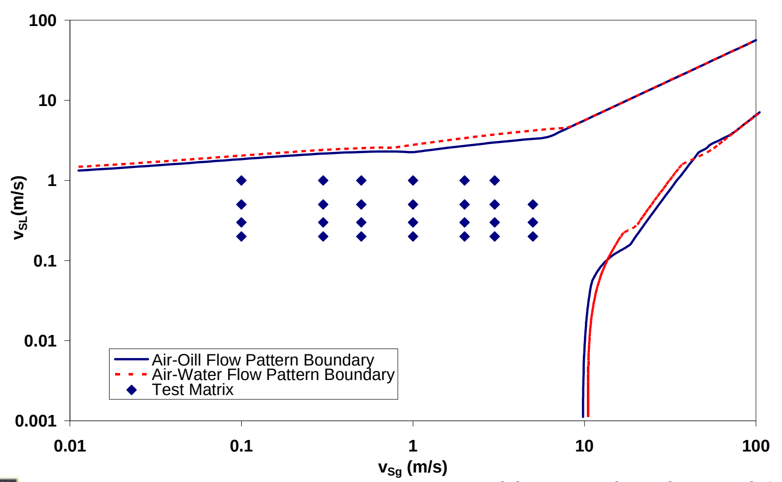
Testing Ranges …

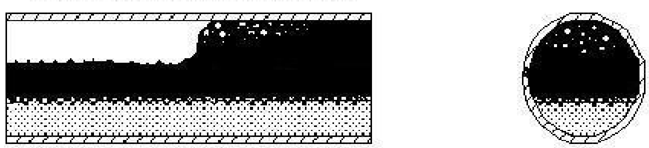
Unified Flow Pattern Map for +5° Inclination Pipe



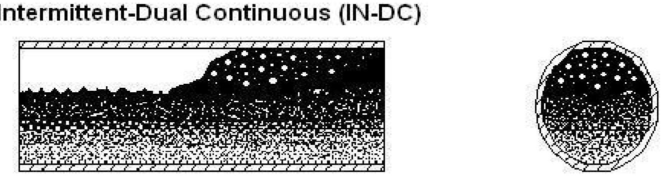
Testing Ranges …

Gas-Oil-Water Slug Flow

Intermittent-Stratified (IN-ST)



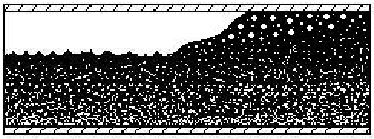
Intermittent-Dual Continuous (IN-DC)

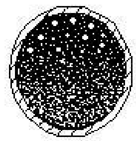


Testing Ranges …

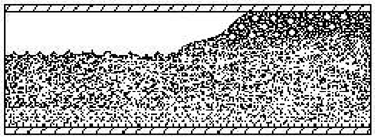
Gas-Oil-Water Slug Flow

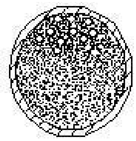
Intermittent-Oil Continuous (IN-OC)





Intermittent-Water Continuous (IN-WC)



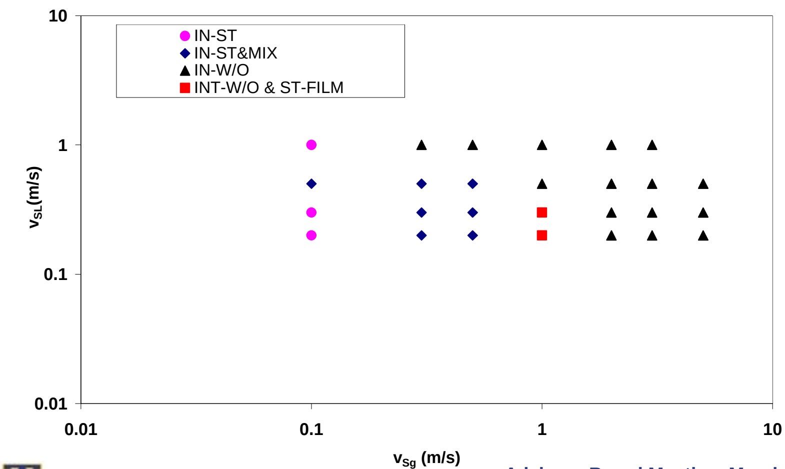


Testing Procedure

- Vary Gas Flow Rate Keeping Oil and Water Flow Rates Constant
- Repeat Above Tests for Several Oil and Water Flow Rates at Constant Water Fraction
- Repeat Above Tests with Different Water Fractions

Experimental Results

Horizontal Gas-Oil-Water Flow Pattern Map for 20% Water Cut



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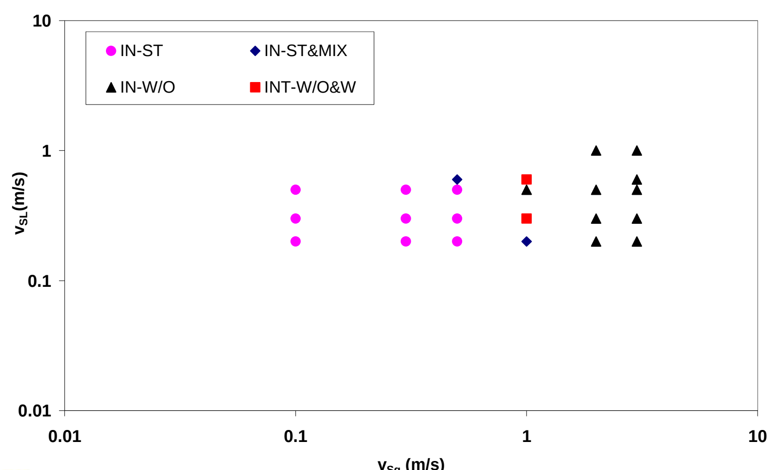
Experimental Results …

Horizontal Gas-Oil-Water Flow Pattern Map for 40% Water Cut



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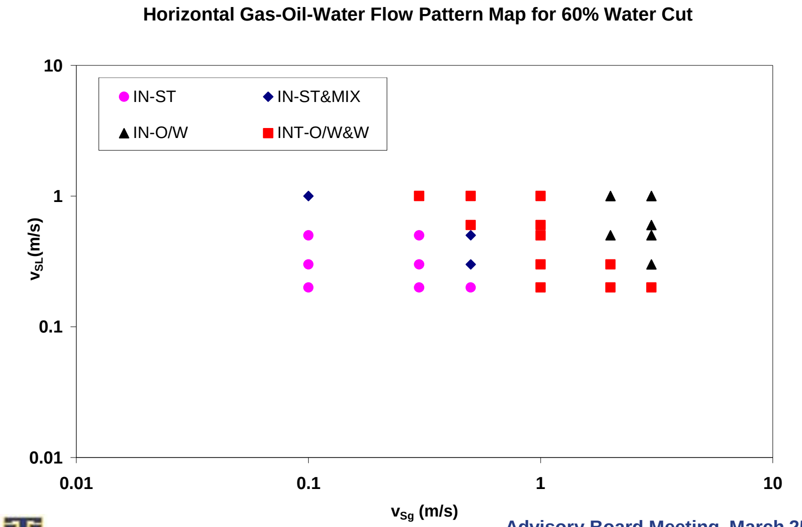
Experimental Results …



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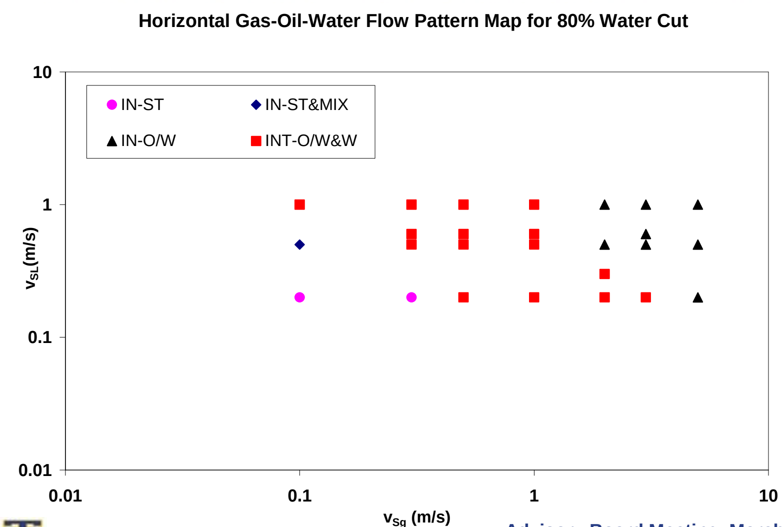
Experimental Results …



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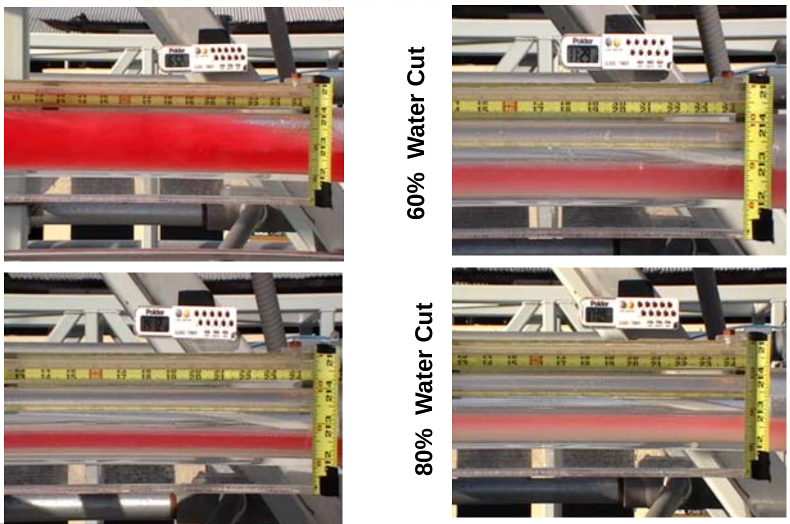
Experimental Results …

20% Water Cut

60% Water Cut

40% Water Cut

80% Water Cut



Fluid Flow Projects

V_{SG}=1m/s V_{SL}=0.5m/s

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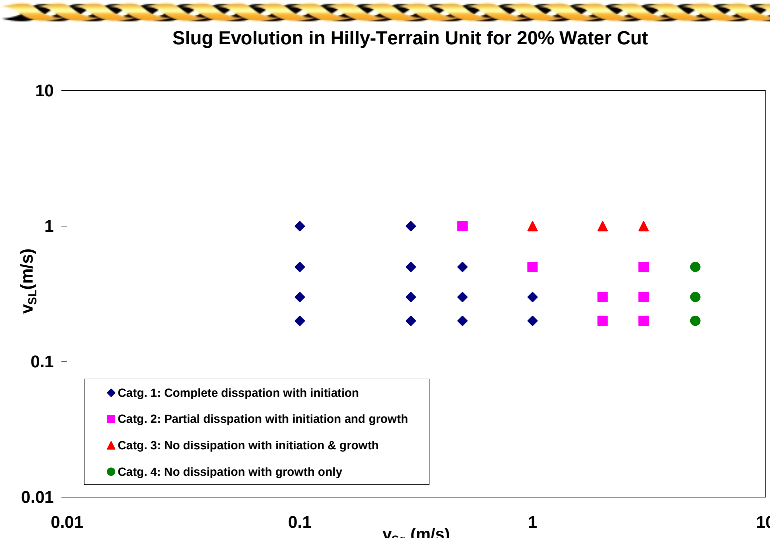
Experimental Results …



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Experimental Results …

20% Water Cut

60% Water Cut

40% Water Cut

80% Water Cut

Fluid Flow Projects

V_{SG}=1m/s V_{SL}=0.5m/s

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Experimental Results …

20% Water Cut

60% Water Cut

40% Water Cut

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Fluid Flow Projects

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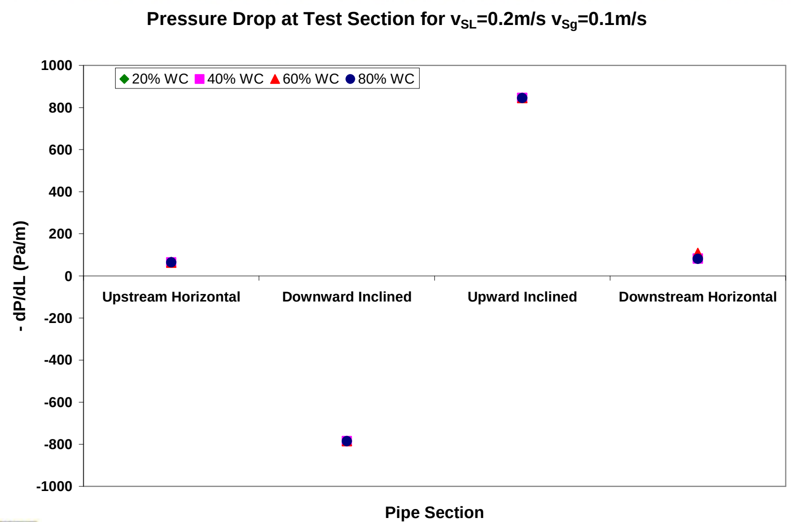
Experimental Results …



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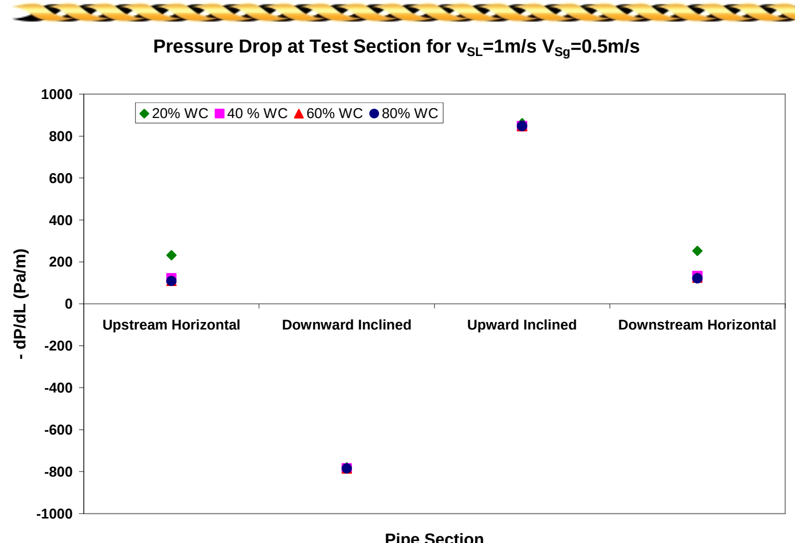
Experimental Results …



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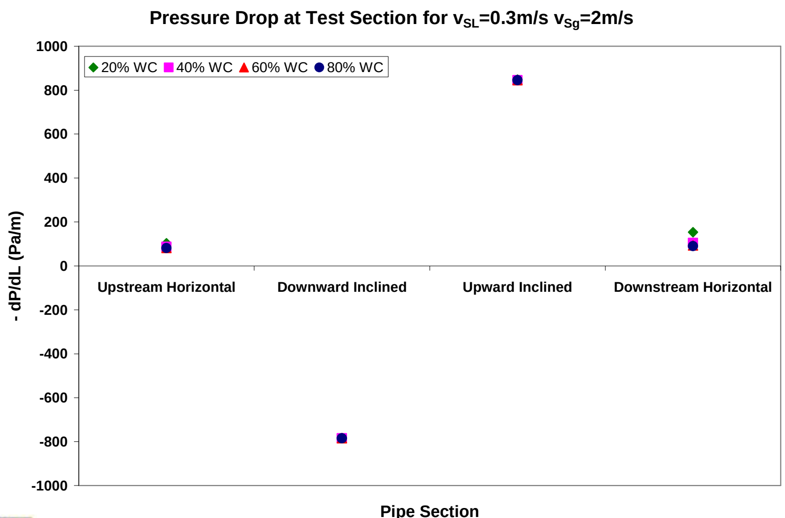
Experimental Results …



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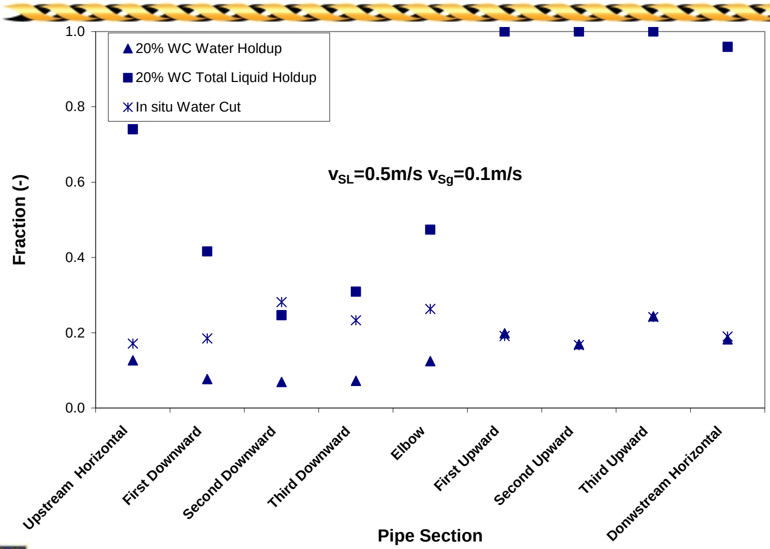
Experimental Results …



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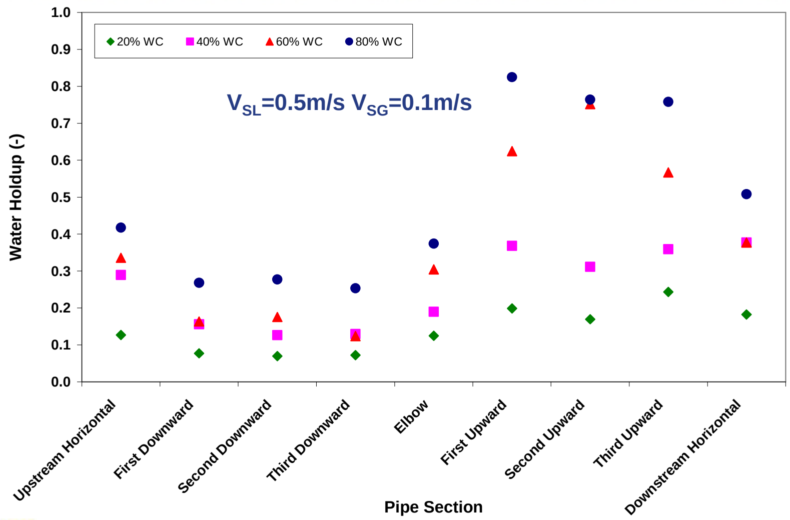
Experimental Results …



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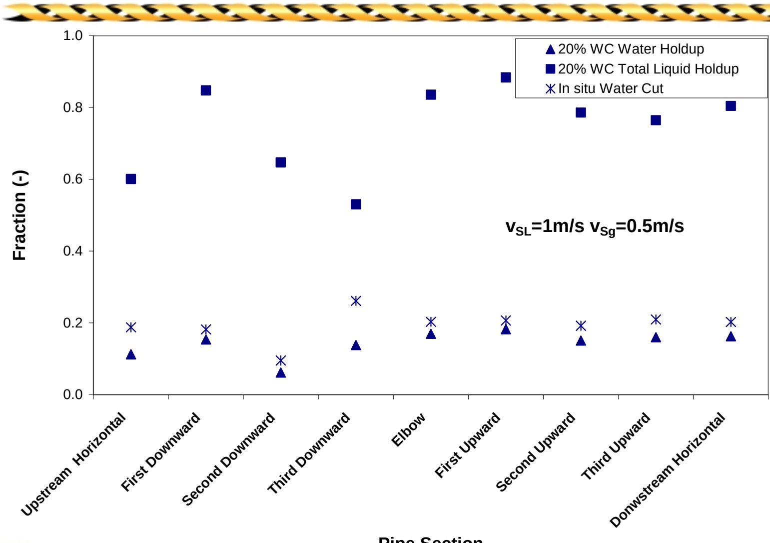
Experimental Results …



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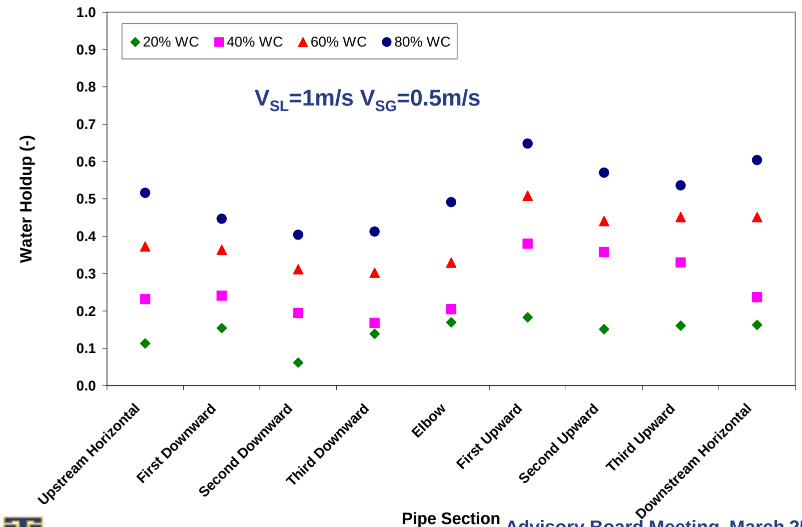
Experimental Results …



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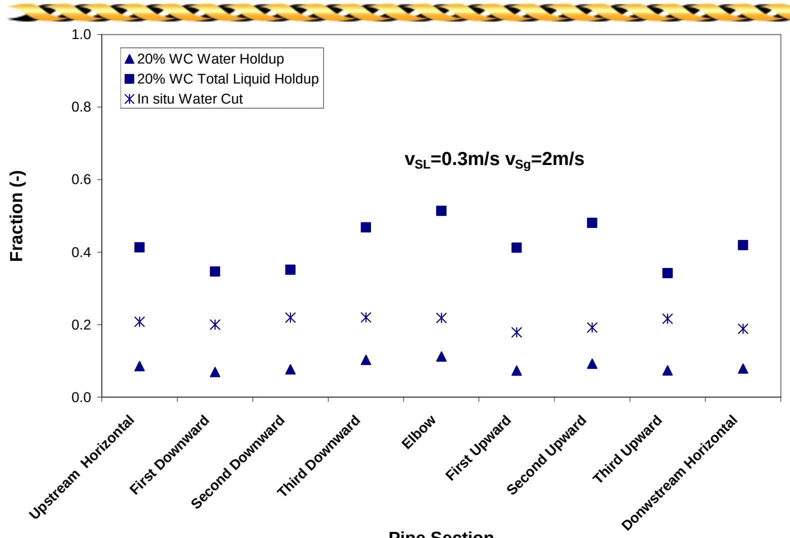
Experimental Results …



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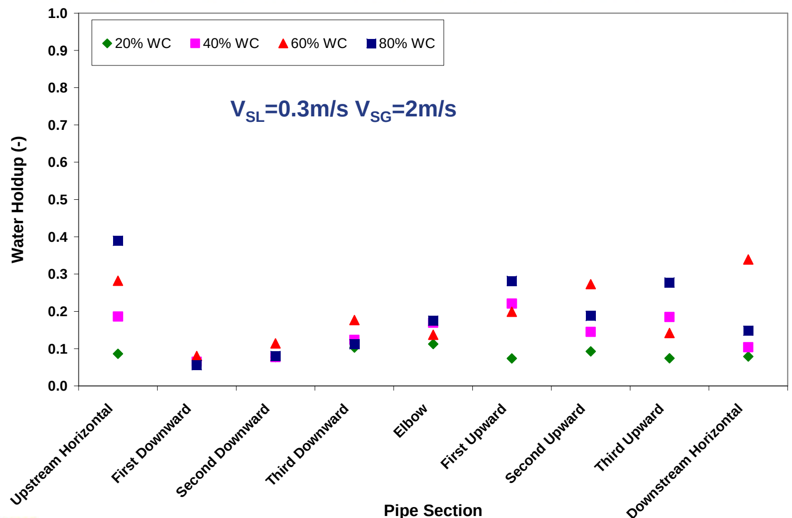
Experimental Results …



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Preliminary Modeling

- Challenges:
  - Lack of Studies Addressing Gas-Oil-Water Flow in Hilly-Terrain Pipelines
  - Significance of Experimental Data
    - Observation of Physical Phenomena
    - Validation of Models

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Preliminary Modeling …

- Testing and Modification of Existing Two-Phase Slug Initiation and Dissipation Models
  - Zhang (2000) Model Taken as Reference for Three-Phase Flow Study
  - Development of Transient Model with Entire Liquid Film as the Control Volume
    - Requires Experimental Data for Closure Relations
    - Slug Tracking Based on Slug Length

Preliminary Modeling …

- Development of Closure Models for Three-Phase Slug Flow on:
  - Slug Length/Frequency
  - Translational Velocity
  - Phase Distribution
  - Average Slug Holdup
- Identifying Flow Regions of Slug Initiation, Growth and Dissipation with Mixing Status of Liquid Phases

Preliminary Modeling …

- Investigation of Water Phase at Hilly-Terrain Unit
  - Water Level in Downhill and Uphill Sections of Hilly-Terrain Unit
  - Water Accumulation at Elbow
  - Critical Values

Preliminary Modeling …

- Investigate Effects of:
  - Phase Distribution
  - Slip Velocity
- Comparison of Developed Model with Multiphase Flow Simulator, OLGA®

Project Schedule

- Model Development June 2009
- Model Validation July 2009
- Final Report August 2009

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Questions & Comments

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Slug Flow Evolution of Gas-Oil-Water Flow in Hilly-Terrain Pipelines

Gizem Ersoy Gokcal

PROJECTED COMPLETION DATES:

Objective

The general objectives of this project are:

- to conduct experiments on three-phase gas-oil-water flow in hilly-terrain pipelines,
- to develop closure models for three-phase slug initiation, dissipation and mixing status of phases,
- to validate developed closure models with experimental results.

Introduction

In the petroleum industry, slug flow is the most complex and dominant flow pattern in horizontal and near-horizontal pipes. Numerous studies have been carried out on slug flow in pipelines. Although slug flow in horizontal and inclined pipes has been studied extensively, slug flow in hilly-terrain pipelines, which are common in both onshore and offshore production and transportation systems, is still not completely understood.

A hilly-terrain pipeline is a pipeline consisting of horizontal, upward inclined, and downward inclined sections. The standard engineering design method for hilly-terrain pipelines has been to divide the pipeline into various sections of constant slopes, and apply steady-state flow models to simulate flow behavior in each section. However, the lack of understanding of how flow characteristics change when these sections are interconnected in hilly-terrain pipelines, prevents enhancing pipeline and downstream facility designs. Some of the most common problems hilly-terrain pipelines cause are operational problems, flooding of downstream facilities, severe pipe corrosion and structural instability of the pipeline, as well as production loss and poor reservoir management due to unpredictable wellhead pressures.

With challenging field conditions, three-phase gas-oil-water flow becomes more common in oil production. The understanding of three-phase flow is crucial for flow assurance problems such as hydrates, emulsions and paraffin deposition. Corrosion and erosion also depend on the characteristics of three-phase flow in pipes. However, very limited amount of work on three-phase flow has been conducted due to the difficulties of oil-water and gas-liquid flow characterizations.

In the open literature, no studies addressing three-phase slug flow in hilly-terrain pipelines could be found. Since slug flow is such a frequently encountered flow pattern in three-phase flow, a study of slug characteristics for three-phase flow in hilly-terrain pipelines is very crucial for production and pipeline transportation. However, the complexity of slug flow increases from two-phase to three-phase flow. The increased complexity in slug flow necessitates transient solutions, supported by closure models. These closure models should focus especially on the phase distribution throughout the flow, and oil-water interactions, as well as the slug flow characteristics. In this study, these models will be examined and studied.

Experimental Study

Experimental Facility and Flow Loop

The experimental work is being conducted using the TUFFP facility for gas-oil-water flow located at the University of Tulsa North Campus Research Complex. The gas-oil-water facility was previously used by Atmaca (2007) for characterization of oil-water flow in inclined pipes. The facility consists of a closed circuit loop with storage tanks, progressive cavity pumps, heat exchangers, metering sections, filters, test section and separator.

For oil and water phases, there are two storage tanks equipped with valves to control the flow rates. Two progressive cavity pumps are used to maintain the liquid flow rates. There are manual bypass valves after the pumps to obtain low flow rates, and pressure relief valves for excessive pressure control. Copper-tube type heat exchangers are used to control the temperature of the fluids during the tests. After the heat exchangers, manual bypass valves allow the fluids to be pumped back to the respective tanks.

Two separate metering sections are equipped with Micro Motion™ Coriolis flow meters to measure mass flow rates and densities of the fluids, and with temperature transducers for monitoring the temperatures of the fluids. Oil and water flow through filters after the metering section. At the inlet of the test section, gas, oil and water flow through the mixing tee to form the gas-oil-water three-phase co-current flow. After the fluids flow through the test section, the mixture is directed to the separator where pressure is set at 20 psig.

The test section is attached to an inclinable boom that makes inclined flow in the loop possible. However, during the three-phase hilly-terrain study, the boom will not be used and the part of the flow loop that is mounted on the boom stay horizontal.

Significant modifications were needed to the flow loop to make enough space for the hilly-terrain section and instrumentation. The original gas-oil-water flow loop consisted of two 21.1-m (69.3-ft) long runs connected with a U-shaped bend to reduce the disturbance of the flow pattern due to a sharp turn. The current test section consists of a 21.1-m (69.3-ft) long upstream branch and a 46.7-m (153.2-ft) long downstream branch connected with a 1.2-m (4-ft) long U-shaped PVC bend as shown in Fig. 1. Both of the branches are made of transparent pipes with 50.8-mm (2-in.) diameter.

The upstream branch of the test section consists of a 13.8-m (45.3-ft) long flow developing section (L/D=272.0), two pressure drop sections 1.17-m (3.83-ft) and 2.79-m (9.3-ft) long, one long pressure drop section combining the two short sections, and one 3.1-m (10.2-ft) long fluid trapping section (L/D=108). The entire upstream branch is placed on the boom.

The downstream branch of the test section consists of a 13.8-m (45.3-ft) long flow developing section (L/D=272.0), a 6-m (19.7-ft) long horizontal section with two short pressure drop sections 4.2-m (14-ft) and 2.13-m (7-ft) long, in addition to a 21-m (68.9-ft) long hilly-terrain section (L/D=413.4) followed by a 6-m (19.7-ft) long horizontal section.

The hilly-terrain section simulates a hilly-terrain unit of 9.5 m (31.3 ft) downhill followed by a 1.9 m (6.2 ft) horizontal and 9.5 m (31.3 ft) uphill sections. The inclination angles are ±1°, ±2° and ±5° for the valley configurations.

The horizontal section immediately downstream of the hilly-terrain section was designed and built similar to the horizontal section immediately upstream of the hilly-terrain section.

The 21.1-m long section of the downstream branch is placed on the inclined boom as in the original gas-oil-water facility. The rest of the downstream branch, which is 25.6 m long, is supported by an aluminum base. A schematic diagram of the test section is given in Fig. 2.

Some hazards have been identified through a facility hazard analysis. Polycarbon protective glass is installed around the test section to provide protection in case of a rupture. In addition, the existing equipment such as pumps, flow meters, separator and storage tanks were checked and made operational.

Instrumentation and Data Acquisition

Instruments on the transparent pipes measure the operating temperature, pressure, differential pressure, total liquid holdup and spatial distribution of the phases.

The facility is divided into four segments. The horizontal section at the upstream branch is the first segment. The horizontal section before the hilly-terrain unit, the hilly-terrain unit and the horizontal section after the hilly-terrain unit are segments two, three and four, respectively. Capacitance sensors, quick-closing valves, laser sensors, and pressure and differential pressure transducers are installed on each segment of the facility. Two temperature transducers are also installed at the inlet of the flow loop and at the beginning of the hilly-terrain unit.

Absolute and differential pressure transducers are used to monitor the flow behavior. Absolute pressure transducers are located at the inlet, before and after the PVC bend and before and after the hilly-terrain unit. The aim of the pressure transducers before and after the PVC bend is to monitor and examine the effects of the bend on the flow. Although early studies on gas-oil-water facility showed that the effects of the PVC bend are negligible, an additional developing section for the flow at the downstream branch is included in this study. There are three differential pressure transducers installed on the horizontal section, at the upstream branch and at the hilly-terrain unit. On each of the other segments, two differential pressure transducers are installed. Pressure gradients over segments are measured with the high-speed data acquisition system to compare the results with laser and capacitance sensors for each test.

Previously developed laser sensors were modified to be used in three-phase slug flow in hilly-terrain pipelines. A new housing design was developed to use the laser sensors at outside conditions. The in-house developed laser sensors were installed on each segment of the facility to obtain translational velocity, slug frequency and slug length. The laser sensors are very sensitive to changes in flow characteristics. A preliminary testing of laser sensors was conducted to test their ability to respond to three-phase slug flow. Since the optical properties of test fluids (water and mineral oil) are very similar to each other, laser sensors were found to be applicable to determine only translational velocity, slug frequency and length. The locations of the laser sensors can be changed easily along the pipe. This enables monitoring slug initiation, slug growth and slug dissipation more easily with the change in operational conditions. The laser sensors are connected to the high-speed data acquisition system to monitor changes in three-phase slug characteristics. There are two laser sensors installed on each horizontal section of the flow loop. There are three laser sensors on the downward branch and five laser sensors on the upward branch of the hilly-terrain section. Using laser sensors with a high speed data acquisition system makes the analysis of slug characteristics easier and more accurate.

Quick-closing valves (QCV) are used for liquid trapping to measure phase fractions and obtain holdup for each flowing condition. The liquid trapped by the quick-closing valves is drained into graduated cylinders to measure the volumes of water and oil phases. There are two quick-closing valves placed in sections one, two and three of the flow loop. The hilly-terrain test section is divided into seven trapping sections to observe the change in liquid holdups with inclination angles. An air tank is also added to provide the air pressure required to operate the QCV.

New capacitance sensors were designed and built in house. They work with the high speed data acquisition system. By using capacitance sensors, translational velocity, slug length and frequency can be measured. Two capacitance sensors are installed for each trapping section. The data obtained from capacitance sensors will be analyzed, along with the data from laser sensors. The locations of capacitance sensors can also be changed along the flow loop to obtain slug characteristics at different flow conditions. The capacitance sensors were made weatherproof in order to eliminate the moisture effect.

Throughout the downstream section of the flow loop, cameras are placed to investigate the details of three-phase slug characteristics in hilly-terrain pipelines. The videos are used to identify the flow patterns and determine the oil-water mixing status, in addition to capturing the details of slug characteristics in three-phase flow in hilly-terrain configurations. Two cameras take the videos through visualization boxes at the horizontal sections before and after the hilly-terrain sections. Four cameras are placed to observe the flow in downward, horizontal and upward sections of the hilly-terrain unit. They are also used to validate the responses of laser and capacitance sensors. For each test, a 10-minute video is taken for analysis.

For data acquisition, Lab View™ 7.1 is used. New hardware, including a high speed data acquisition system was installed for absolute and differential pressure transducers and laser and capacitance sensors. With the instruments connected to a high speed data acquisition system, slug flow characteristics are captured and compared more efficiently. For the high speed data acquisition, different sampling rates were used to determine an optimum sampling rate. For most of the test matrix, a sampling rate of 100 sample/s was found to be acceptable. For high flow rates, the sampling rate can be increased based on the three-phase slug characteristics. The existing program for the low speed data acquisition was updated for three-phase gas-oil-water flow in hilly-terrain studies. A sampling rate of 1 sample/s was selected to collect data for this data acquisition system. The data logging for each test is 10-minutes.

Test Fluids

For the experiments of three-phase flow in a hilly-terrain pipeline, fresh water, air and refined mineral oil were chosen as the testing fluids. The refined oil, Tulco Tech 80, was chosen based on its easy separation. The physical properties of Tulco Tech 80 are given below:

- API gravity: 33.2°
- Density: 858.75 kg/m^3 @ 15.6°C
- Viscosity: 13.5 cp @ 40°C
- Surface tension: 29.14 dynes/cm @ 25.1°C
- Interfacial tension with water: 16.38 dynes/cm @ 25.1°C
- Pour point temperature: -12.2°C
- Flash point temperature: 185°C

The properties of Tulco Tech 80 were measured by Chevron labs. Figures 3 and 4, show the density and viscosity changes with temperature at three different flow rates.

Preliminary Testing

After the facility construction was completed, a preliminary experimental study was conducted in order to test the installed instruments on the facility. Both laser and capacitance sensors were used for the first time for three-phase gas-oil-water experiments.

The laser sensors were tested for their sensitivity to the test fluids. After mounting the sensors on the facility, dispersed cases of oil and water with different three-phase flow operational conditions were tried. In addition to the change in operational conditions, data acquisition frequency was changed to determine the optimum frequency. Six cameras stationed through out the facility were used to compare and validate the responses of laser sensors. Based on the experiments, the targeted test matrix can be divided into three sections.

The first region corresponds to the operational conditions with the slowest flow rates. At this region, mostly elongated bubbles and slug flow with low slug frequency occur at horizontal section. The liquids are segregated. Slugs dissipate at the entrance of the downward section at the hilly-terrain unit. Liquid fallback and slug generation are observed for the upward branch of the hilly-terrain section. Liquid height increases with an increase in liquid velocities. The laser sensors outputs are compared with the videos taken from cameras. Both measurements gave almost identical results at the same conditions. Therefore, both laser sensors and cameras can be used to measure three-phase slug flow characteristics for this region.

As expected, slug frequency increases with an increase of superficial gas velocity. It is observed that a dispersion of w/o also forms due to the mixing. This flow region is accepted as the second flow region. The slugs still decay at the downward section and create waves at the liquid-gas interface. At the dip section, there is still liquid accumulation. At the upward section of the hilly-terrain, slug growth takes place. However, the slug frequency is much higher than in the first flow region. When the responses of laser sensors for this region are analyzed, it has been observed that sensors cannot capture the chaotic behavior. The frequency of data acquisition was also increased to see if any improvements can be obtained. However, the laser sensors would still not work properly. Due to the increase in turbulence and mixing in the flow, the laser sensors can not differentiate between the slug body and film liquid regions. The entrained gas bubbles and dispersion of liquids affect the performance of sensors. In conclusion, only the camera system can be used to measure slug flow characteristics in this region.

As the superficial velocities of all phases increase, slug flow with fully mixed liquids is observed. The slug frequency is the highest among all of the three flow regions. The slug flow that is developed in the horizontal section still existed in the downward section. However, no increase in height of liquid is observed at the dip section of the hilly-terrain unit. Most of the operational conditions were so high that slug characteristics did not change significantly through the hilly-terrain section. For this high flow rate region, the performances of neither laser sensors nor camera system were satisfactory. Due to the limitations of instruments and no significant changes in slug flow, this flow region was omitted from the test matrix. The test matrix was revised to focus more on changes of slug characteristics with low flow rates.

Experimental Ranges

In this study, 108 tests were conducted for three-phase air-oil-water flow in hilly-terrain pipelines with an inclination angle of ±5° for thevalley configuration. Although the facility can be modified to run at ±1°, ±2° and ±5° for the valley configurations, the inclination angle for the hilly-terrain unit was set to ±5˚ due to time constraints. This inclination angle was selected to observe the most significant changes in three-phase slug flow. The testing ranges for the three-phase hilly-terrain experiments on the gas-oil-water flow loop are as follows:

Superficial gas velocity: 0.1-5.0 m/s

Superficial oil velocity: 0.04-1 m/s

Superficial water velocity: 0.02-1 m/s

Water fraction: 20, 40, 60 and 80 %

The lower limits of superficial velocities were decided on by the accuracies of the Micro Motion™ flow meters. The higher limits were set by the pressure gradient and facility limitations.

The test matrix was arranged in order to include both the flow pattern transition from stratified to slug flow and phase distributions from low water cut to high water cut. For each water cut value, 27 data points were taken.

The Unified flow pattern maps showing the test matrix for inclination angles of 0˚,-5˚ and 5˚ are shown respectively in Figs. 5 - 7. The test matrix was selected such that slug flow would exist in the horizontal 2-in. pipe. It should be noted that these flow pattern maps were drawn for each section of the pipeline with a fixed inclination angle. In a hilly-terrain pipeline, the flow pattern in an upstream section may persist in a downstream section for a considerable distance, while the flow pattern prediction may be different.

The observed three-phase flow patterns based on the test matrix for this study are Intermittent-Stratified (IN-ST), Intermittent-Oil Continuous (IN-OC) and Intermittent-Water Continuous (IN-WC) as described in Keskin et al. (2007).

Test Program

A typical test program for gas-oil-water flow in a hilly-terrain pipeline starts with varying the gas flow rate, keeping the oil and water flow rates and water fraction constant. Then, tests are repeated for several oil and water flow rates at constant water fraction, and continued with various water fractions.

Experimental Results

Experiments on hilly-terrain effects on three-phase slug flow characteristics were conducted. Experimental data contain visual observations, differential pressure, average holdup, slug frequency and length. In the following subsections, the changes in some of these variables vs. water cut are presented and discussed.

Three-Phase Flow Patterns

Three-phase gas-oil-water slug flow experiments have been conducted for 20, 40, 60 and 80% water cuts at various flow rates. Three-phase flow patterns have been observed using the video system at the facility. The flow pattern maps for the operational water cuts at the downstream horizontal section are shown in Figs. 8 – 11. The three-phase flow patterns are classified based on three-phase flow patterns by Keskin (2007). Oil-water phase mixing status is analyzed for the operational conditions.

It has been observed that for lower water cuts and flow rates, oil and water phases are completely segregated and slug flow exists between air and oil phases. As the flow rates are increased, more mixing can be observed. At moderate flow rates, a layer of segregated water phase with dispersion on top can be seen clearly regardless of the water cut value. Especially when the slug frequency is low, dispersions of w/o and o/w can segregate. The segregation can be observed especially in the film region. As the flow rates are increased, oil and water phases mix more, and more homogeneous dispersions exist.

In the previous studies of two-phase hilly-terrain pipelines, different cases of flow were identified for slug dissipation, initiation and growth along the hilly-terrain section (Al-Safran, 2003). Analysis of these cases is expected to give a better understanding of the physical phenomena. The observed flow cases are: (1) complete dissipation in the downhill section with slug initiation at the dip, (2) partial dissipation in the downhill section with initiation and growth at the dip, (3) no dissipation in the downhill section with initiation and growth at the dip, and (4) no dissipation in the downhill section with growth only at the dip. Figure 12 shows these cases for 20% water cut. The first category is the most common case. It is observed mainly at low and moderate flow rates. In this category, slug flow completely dissipates along the downhill section of the hilly-terrain unit. At the elbow, slug initiation takes place. For the second category, the slug flow still survives in the downhill section. However, slug frequency decreases. At the elbow section of the hilly-terrain unit, liquid accumulates and results in either slug initiation or slug growth of the surviving slugs. When flow rates are relatively higher, no dissipation takes place in the downhill section. This category is specified as category three. At the elbow, slug initiation and growth takes place. The fourth category is when the superficial gas velocity is increased and there is no slug dissipation in the downhill section and slug initiation at the elbow.

Pressure Gradient

The pressure gradients in the horizontal sections before and after the hilly-terrain unit, and in the downward and upward inclined sections of the hilly-terrain section are shown for low, moderate and high flow rates in Figs. 13-15. The effect of water cut can be seen in the horizontal sections before and after the hilly-terrain unit. However, no significant effect of water cut can be observed in the inclined sections.

Liquid Holdup

The changes in average water, total liquid holdup and in situ water cut for an operational case where superficial liquid and gas velocities are low are shown in Fig. 16. Oil and water phases are segregated and slugs dissipate completely in the downhill section of a hilly-terrain unit. The water and total liquid holdup decreases significantly in the downhill section. At the elbow, they start to increase due to liquid accumulation that is caused by liquid flowing from the downhill and liquid falling back from uphill sections of a hilly-terrain unit. Due to the initiated slugs, holdup values increase in the uphill sections. The in-situ water cut increases in the downhill sections and reaches a maximum value at the elbow. In the inclined sections, water cut decreases back to 20%. Figure 17 shows the changes in average water holdup for different inlet water cuts. The trends of the water holdup are similar for the same operational conditions.

When the superficial velocities are increased, slug flow still exists in the downward inclined section. Therefore, the change in water holdup and water cut along the test section is less dramatic, as shown in Fig. 18. Figure 19 shows the holdup trends with varying water cuts. The decrease in water holdup values in the downward inclined section is less severe, due to the surviving slugs.

For moderate flow rates, water is completely dispersed in oil. There was no discernable change in water cut and water holdup values (Fig. 20). When the water cut values are increased for the operational conditions, the trend in water holdup values becomes more similar to previous cases due to the existence of a segregated water layer flowing at the bottom of the pipe (Fig. 21).

Slug Frequency

Slug frequency is determined by dividing the number of slugs detected by the laser or capacitance sensors by the test duration. Times for the slug front and back to travel from the first laser sensor to the second one can be obtained. Since the distance between the two sensors is known, the slug front and back velocities can be calculated. The slug translational velocity can be obtained by taking the average of the slug front and back velocities. If the time difference between a slug front and back passing one of the laser sensors can be determined, slug length can be calculated using the translational velocity. Slug frequency analysis is still ongoing and will be presented in the fall of 2009.

Preliminary Modeling Study

As briefly discussed in the previous section, different cases of flow were identified for slug dissipation, initiation and growth along the hilly-terrain unit (Al-Safran, 2003). As a first step to model three-phase effects on slug growth and initiation mechanisms, these flow cases will be improved by including the three-phase flow patterns.

At low flow rates, it has been observed that phases are mostly mixed in the slug body and segregated fluids exit in the film region. With an increase in water cut, wave growth and coalescence mechanisms in the film region will be analyzed and slug initiation models will be investigated. Although the slug tracking model by Taitel and Barnea (1999) can predict the two-phase slug length distribution along a hilly-terrain pipeline, it requires measured slug length distributions at the entrance and at the lower dip to simulate the flow behavior. In an earlier study by Zheng (1991), the effects of a hilly-terrain pipeline configuration on two-phase flow characteristics were investigated. A simple slug-tracking model was proposed that follows the behavior of all individual slugs for a rather simple geometry consisting of a single hilly-terrain unit. His model is based on a sink/source concept at the pipeline connections, where an elbow accumulates liquid as a sink and releases liquid as a source. However, the model requires improvement in the slug dissipation mechanism. Yuan (2000) investigated two-phase slug dissipation in the downward flow section of a hilly-terrain pipeline. Experiments were done for inclination angles of 1°, 2°, 5°, 10° and 20°. Slug flow in downward inclined pipes was grouped in terms of change in slug frequency. Based on previous studies, Zhang (2000) used the entire liquid film and gas pocket in the film zone as a control volume for developing two-phase slug flow. This enabled momentum exchange between the slug body and liquid film. Unsteady continuity and momentum equations were derived for the control volume. Continuity and momentum equations were derived for the control volume relative to a coordinate system moving with translational velocity, V_t. It was assumed that, the slug frequency of an unsteady slug flow may change while the slug length remains constant. The liquid mass in the disappearing slug units were distributed to each liquid film of the remaining slugs. This two-phase modeling approach for developing slugs will be compared with three-phase experimental data, and the continuity and momentum equations will be modified and rearranged for three-phase flow. The analysis of this model is expected to improve the TUFFP unified hydrodynamic model.

For the model development, closure relationships such as liquid holdup in the slug body are needed. Three-phase gas-oil-water slug flow data will be analyzed considering the changes in water cut. Using the experimental findings, the closure models for slug length and frequency, translational velocity, and slug holdup will be investigated and developed if possible. These closure models will be integrated into three-phase momentum equations.

Accumulation of water at low spots in pipelines can cause serious corrosion and hydrate problems. As an auxiliary study, water level in downward and upward flow in the hilly-terrain section and segregation of the water phase will be observed and analyzed. At the elbow of the hilly-terrain unit, water accumulation and critical values of mixture velocity to sweep the water phase will be studied with different water cuts and mixture velocities.

The experimental database on three-phase flow is very limited and the literature shows a lack of studies that address modeling of three-phase gas-oil-water flow in hilly-terrain pipelines. Therefore, the experimental work plays a significant role in the modeling study. The resulting models will be validated with experimental data and compared with a multiphase flow simulator, OLGA®.

Near Future Activities

The modeling study, including the model validation, is expected to be finished by July 2009. The final report will be submitted in August 2009.

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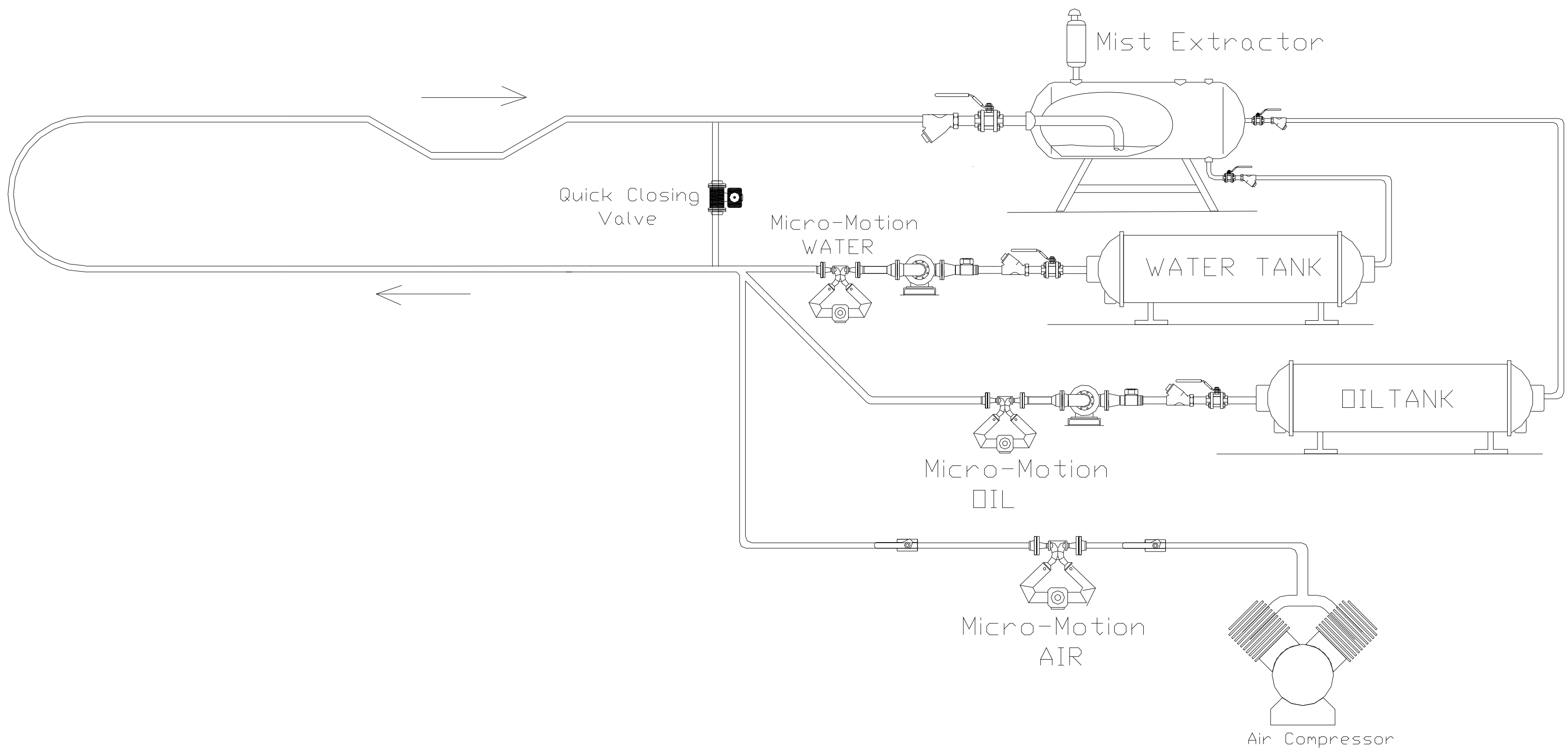

Figure 1: Gas-Oil-Water Facility Schematic

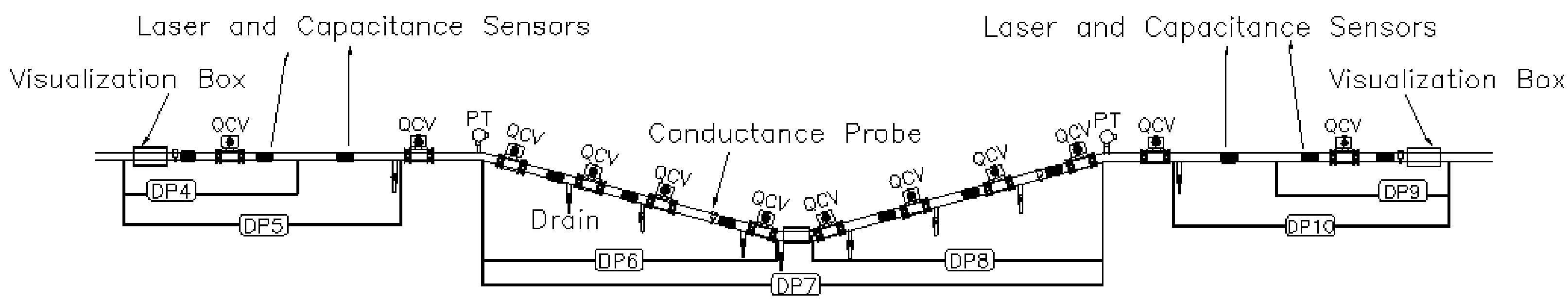

Figure 2: Schematic of Downstream Branch of Test Section

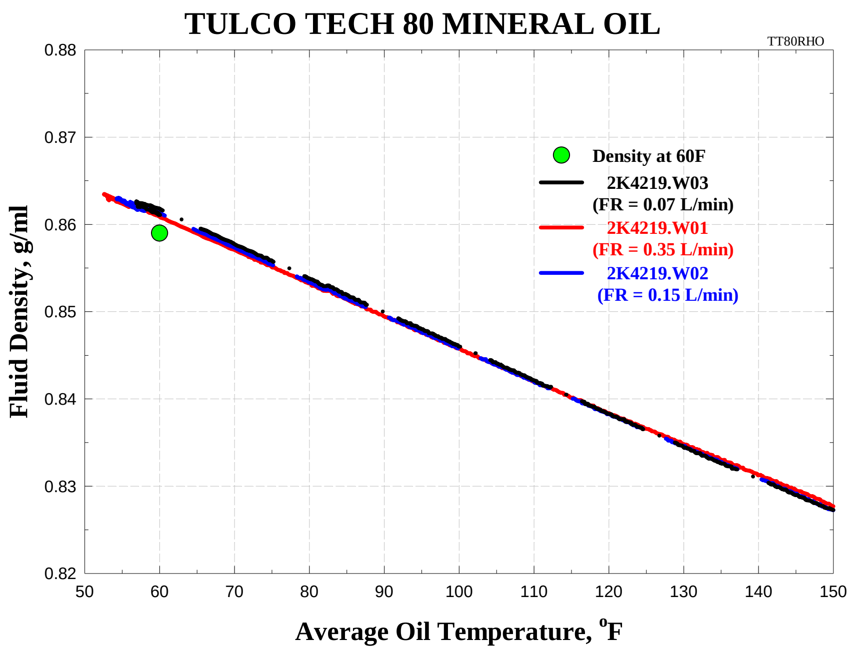

Figure 3: Tulco Tech 80 Oil Density vs. Temperature

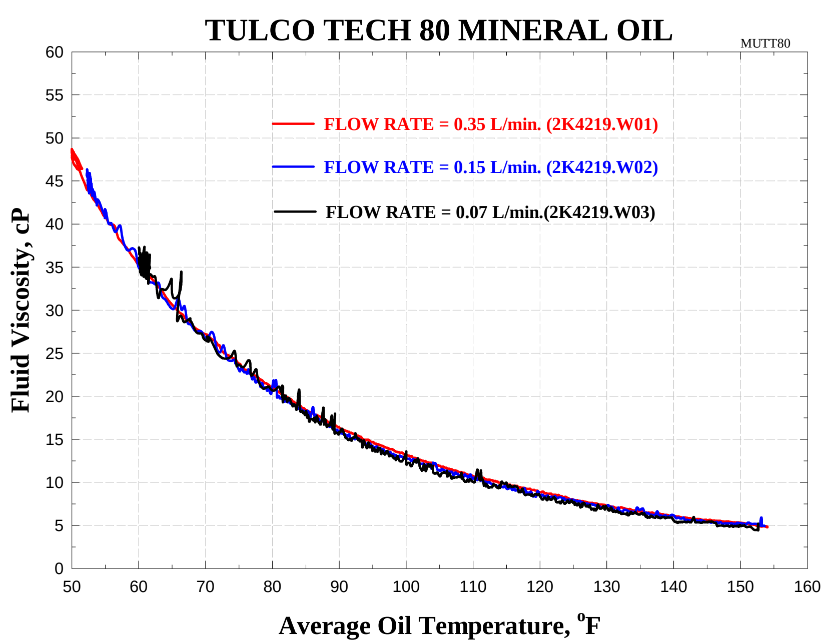

Figure 4: Tulco Tech 80 Oil Viscosity vs. Temperature

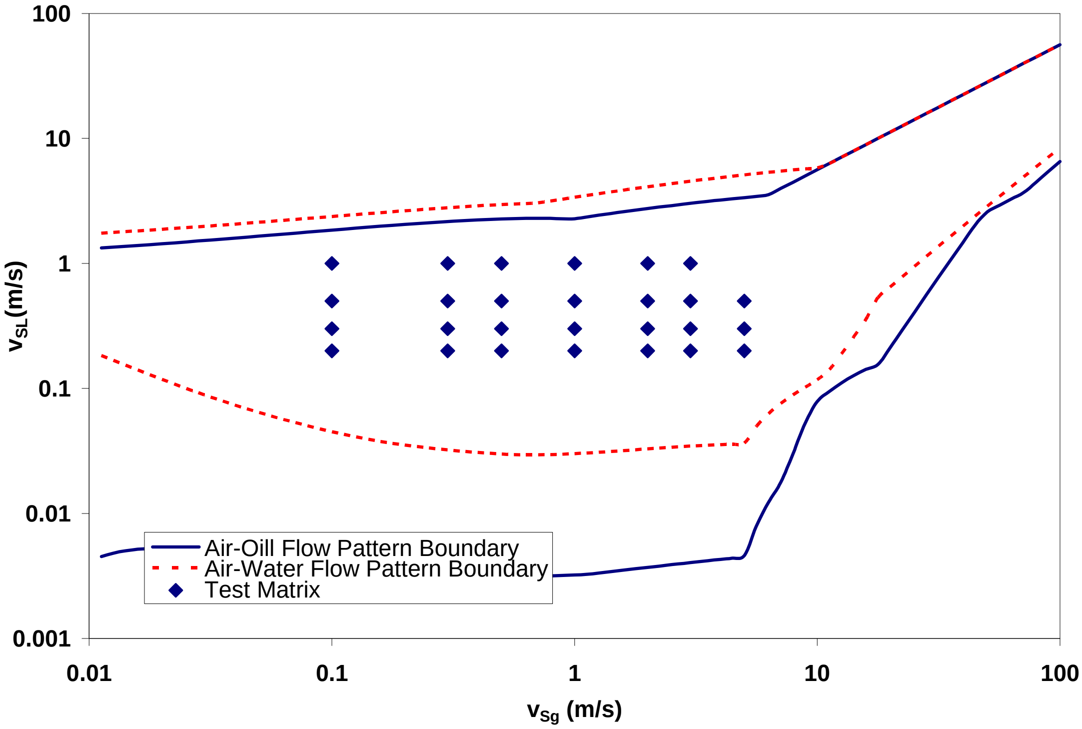

Figure 5: Unified Horizontal Flow Pattern Map in 2-in ID Pipe

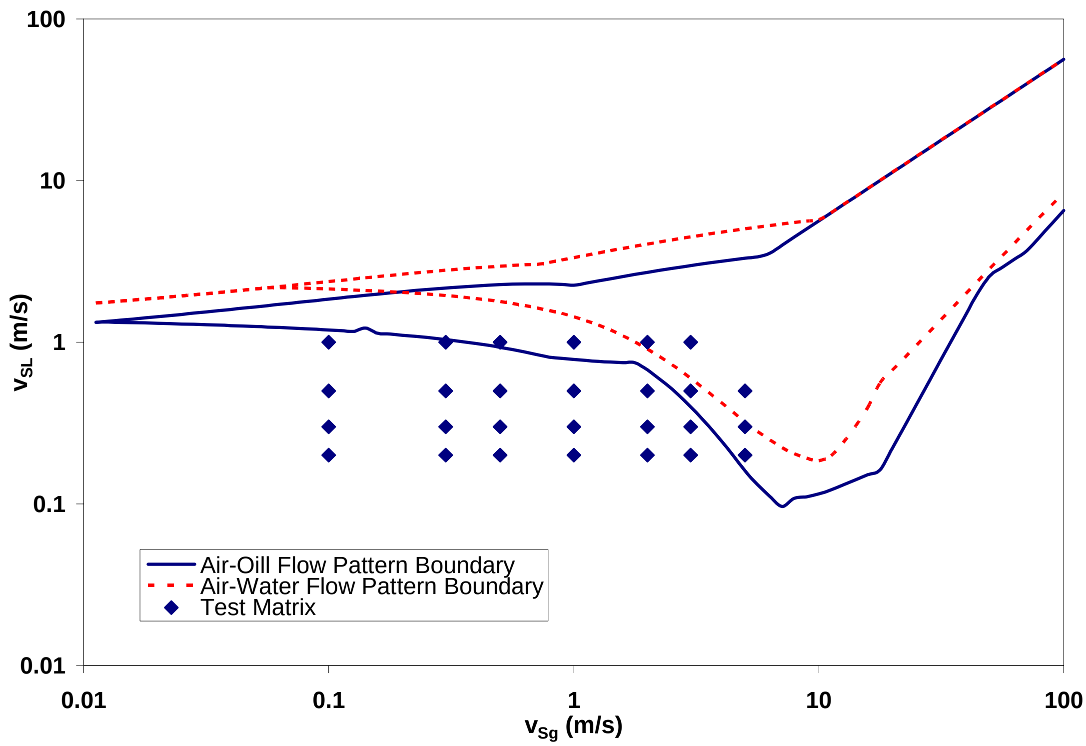

Figure 6: Unified Flow Pattern Map in 2-in ID Pipe (-5°)

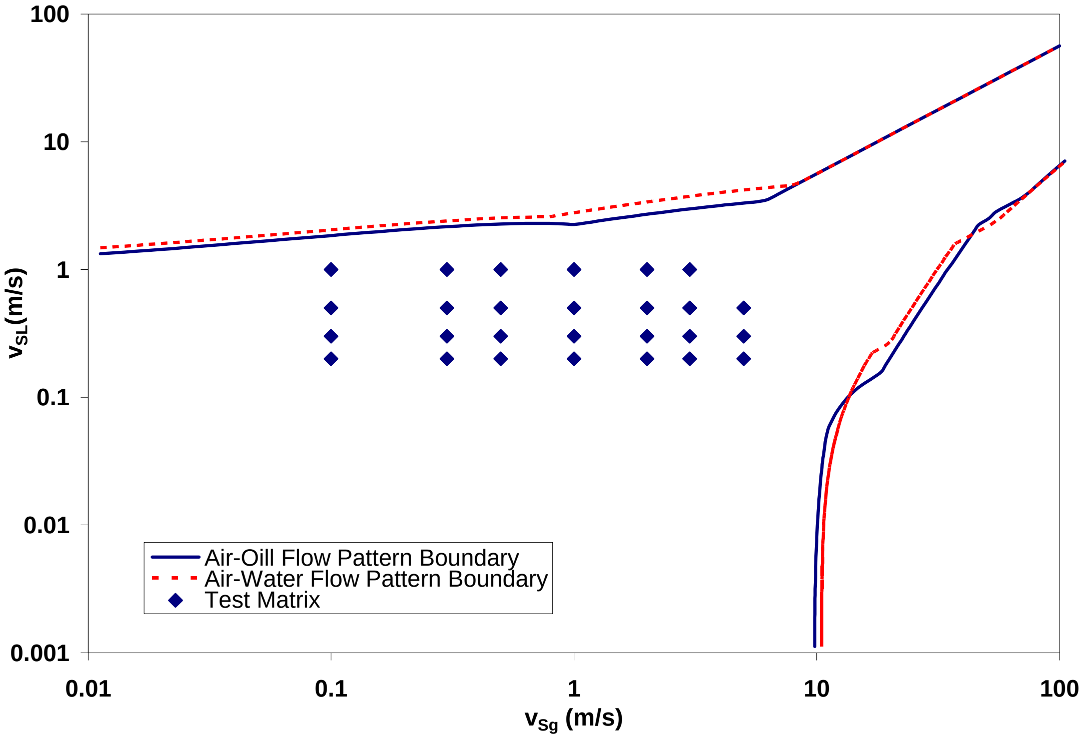

Figure 7: Unified Flow Pattern Map in 2-in ID Pipe (5°)

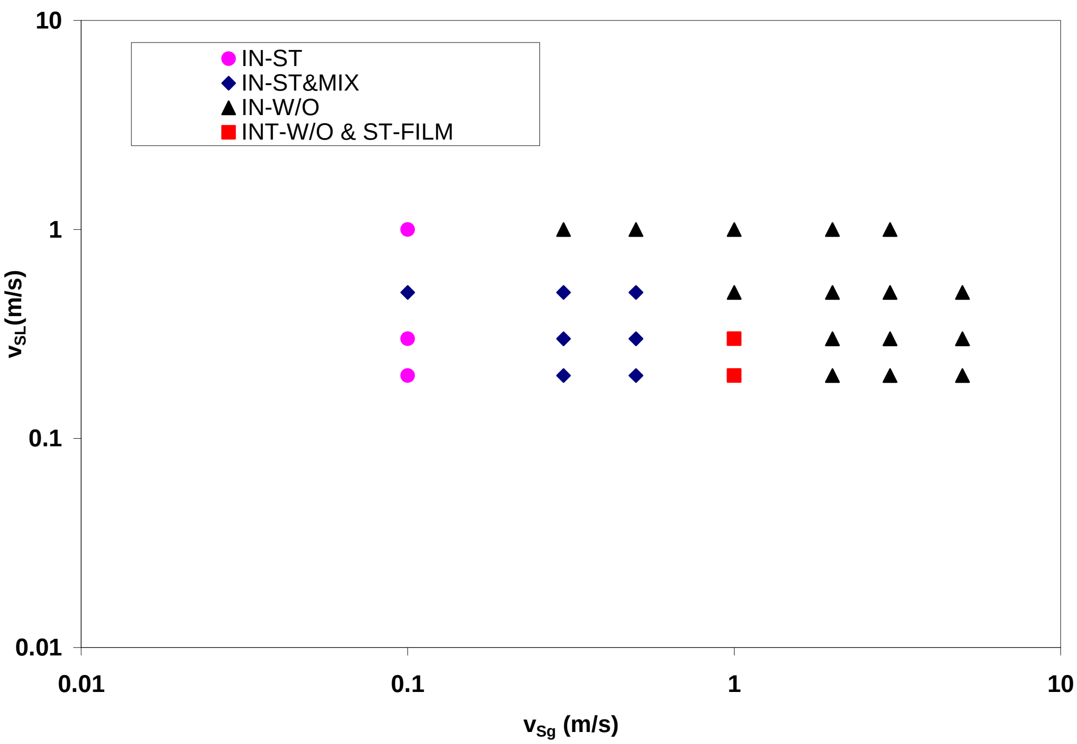

Figure 8: Horizontal Gas-Oil-Water Flow Pattern Map for 20% Water Cut

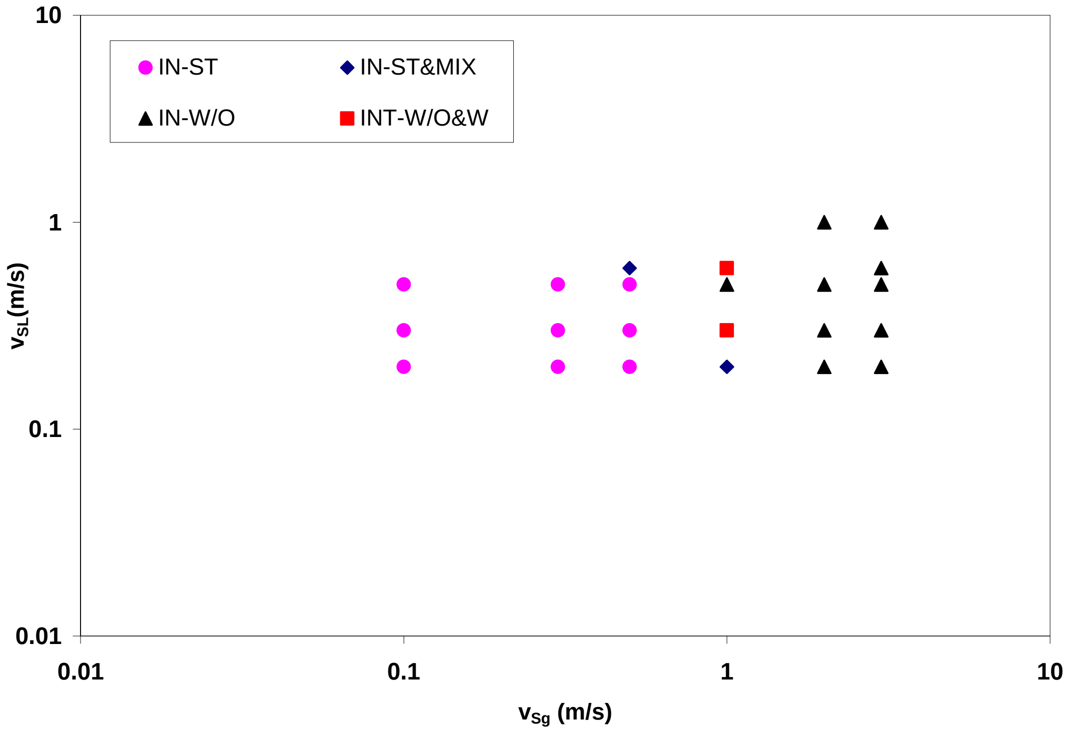

Figure 9: Horizontal Gas-Oil-Water Flow Pattern Map for 40% Water Cut

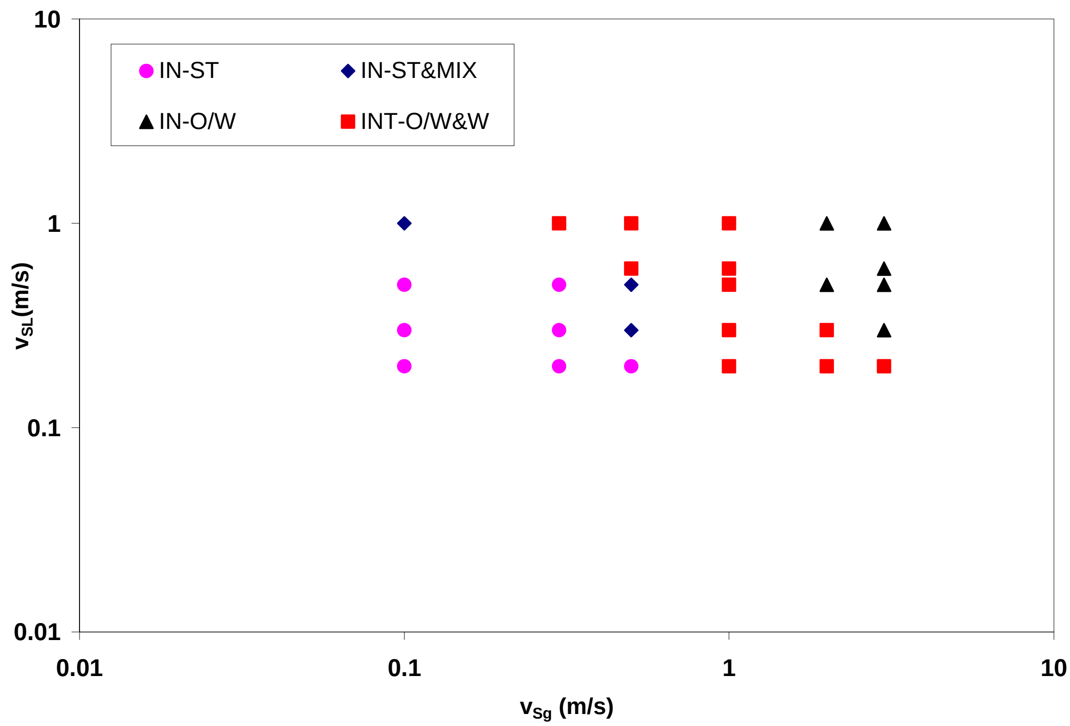

Figure 10: Horizontal Gas-Oil-Water Flow Pattern Map for 60% Water Cut

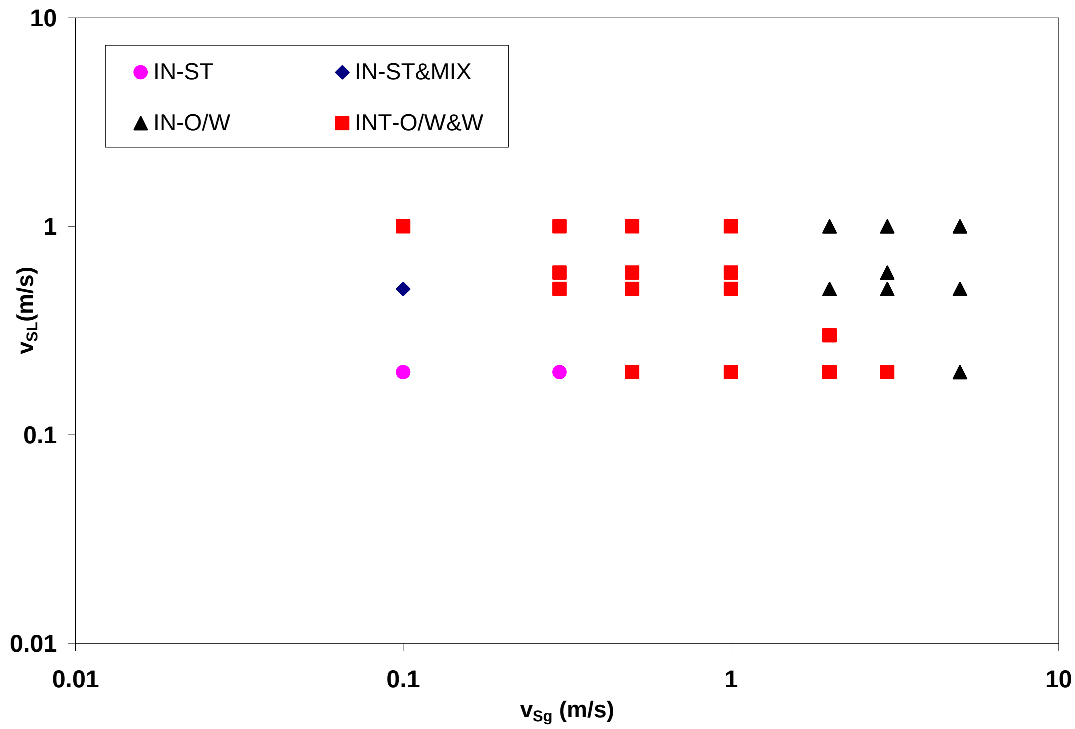

Figure 11: Horizontal Gas-Oil-Water Flow Pattern Map for 80% Water Cut

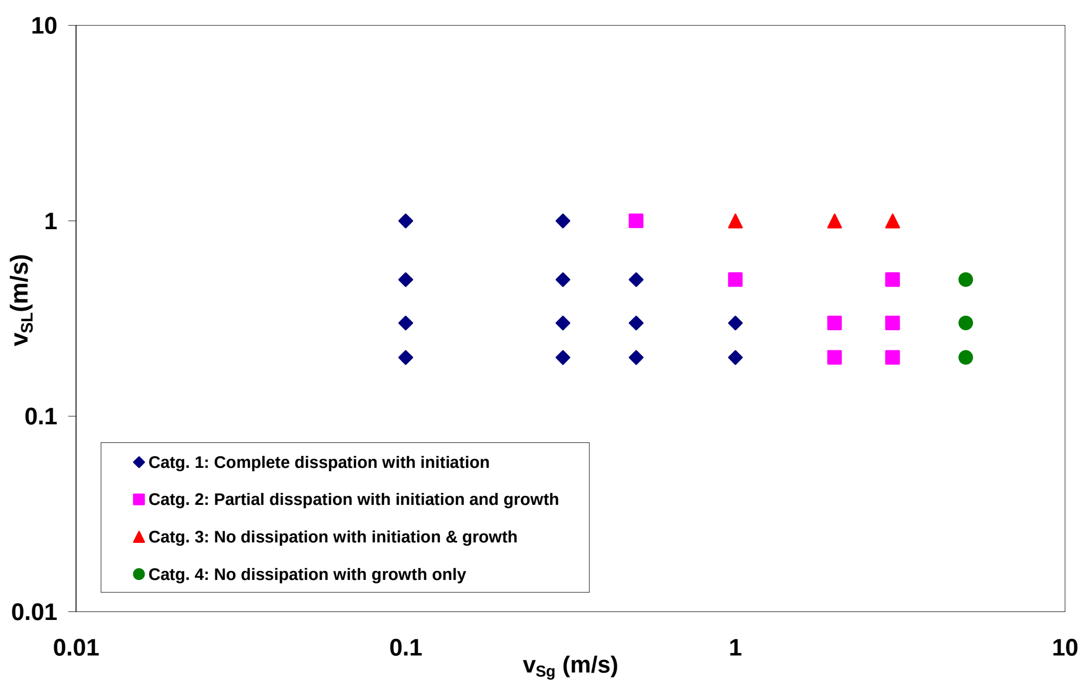

Figure 12: Slug Evolution in Hilly-Terrain Unit for 20% Water Cut

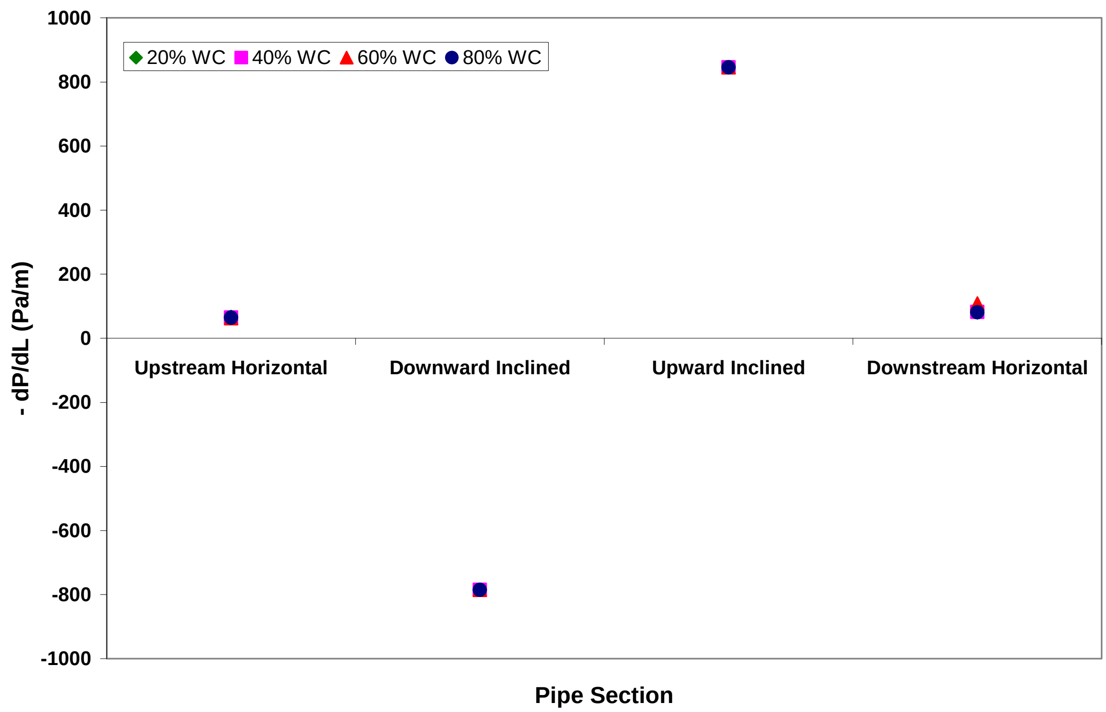

Figure 13: Pressure Drop along the Test Section for v_{SL} = 0.2 m/s, v_{Sg} = 0.1 m/s

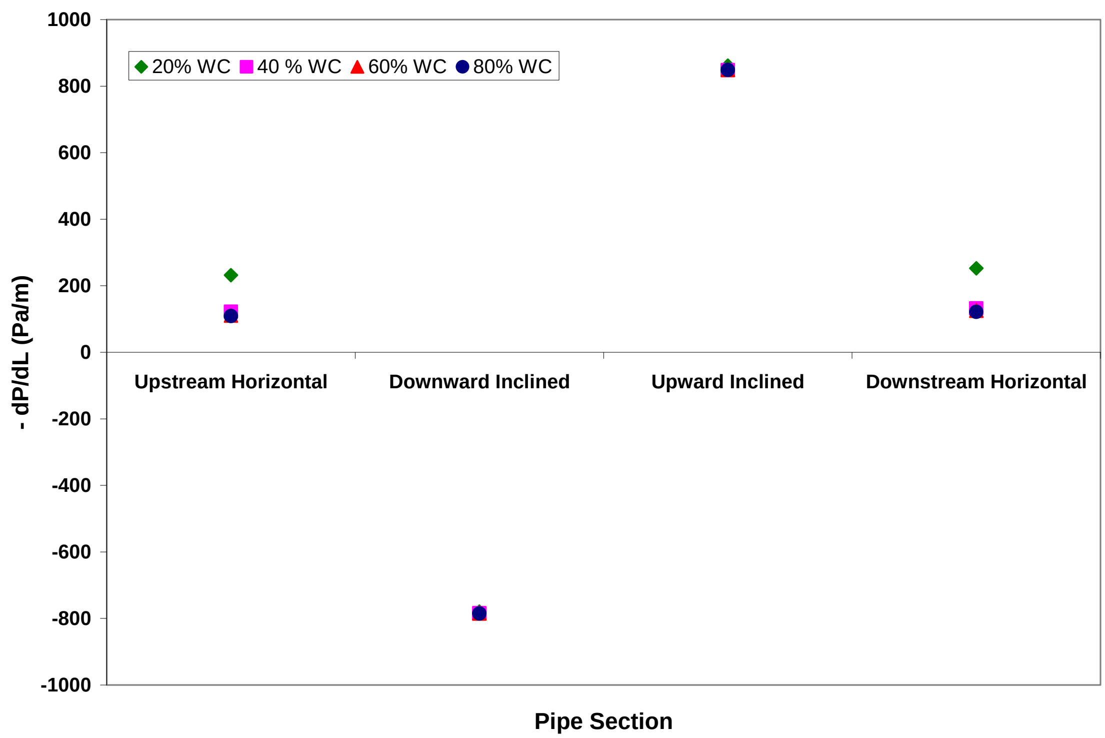

Figure 14: Pressure Drop along the Test Section for v_{SL} = 1 m/s, v_{Sg} = 0.5 m/s

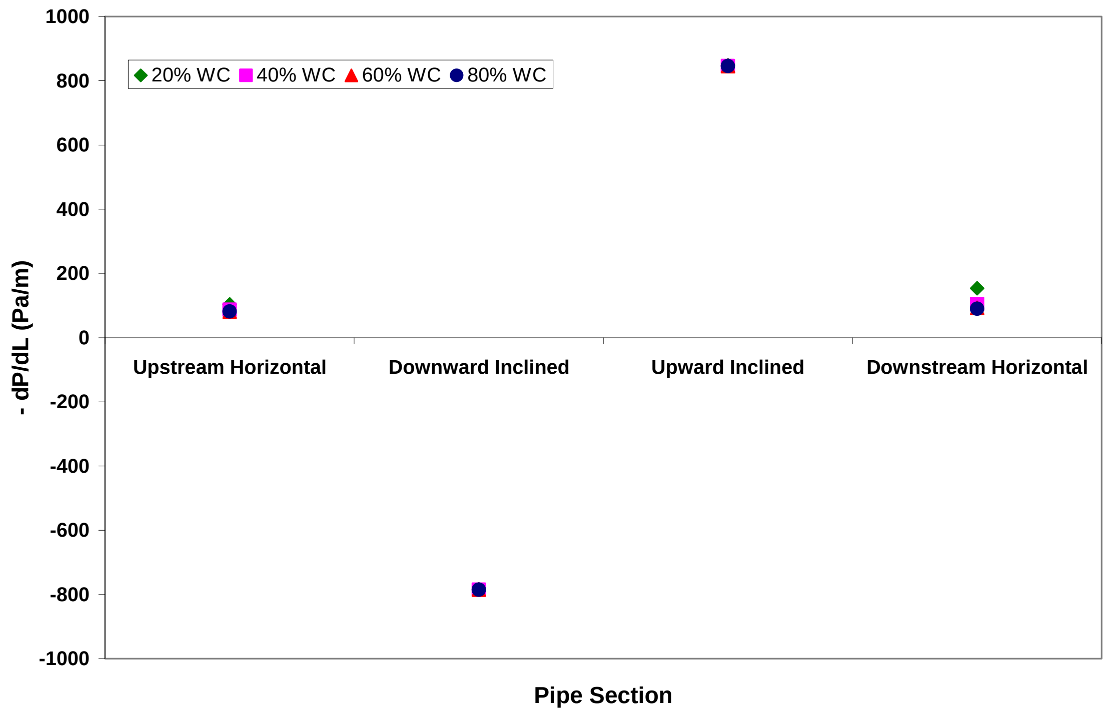

Figure 15: Pressure Drop along the Test Section for v_{SL} = 0.3 m/s, v_{Sg} = 2 m/s

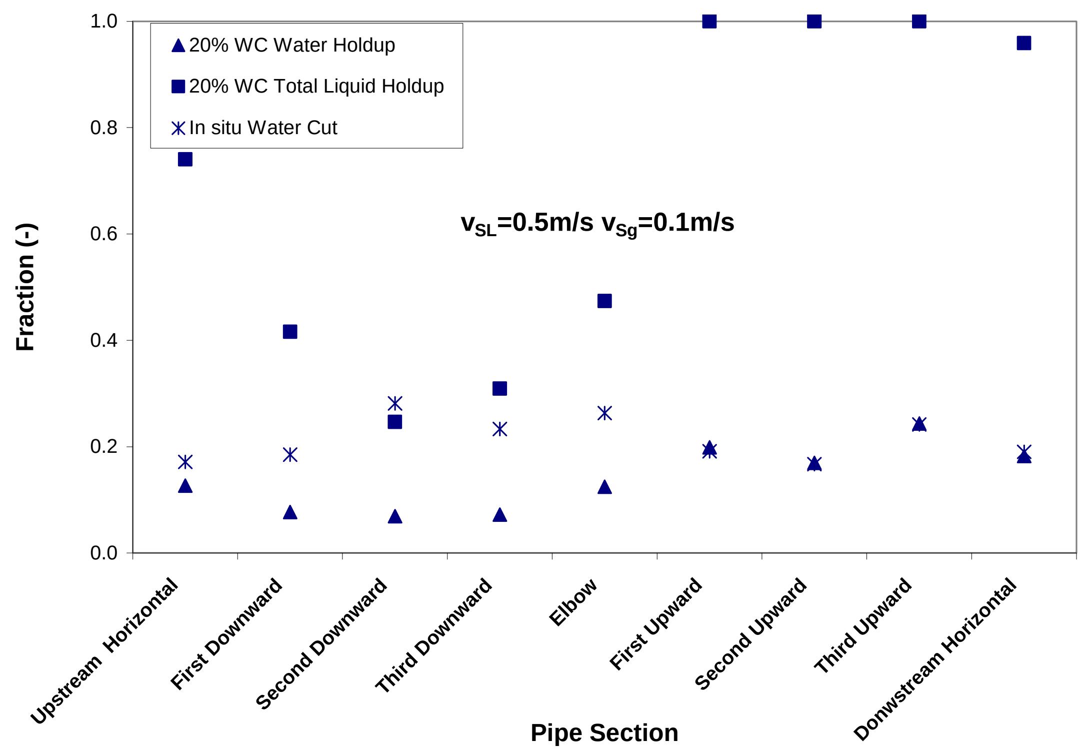

Figure 16: Liquid Fraction Values for v_{SL} = 0.5 m/s, v_{Sg} = 0.1 m/s at 20% Water Cut

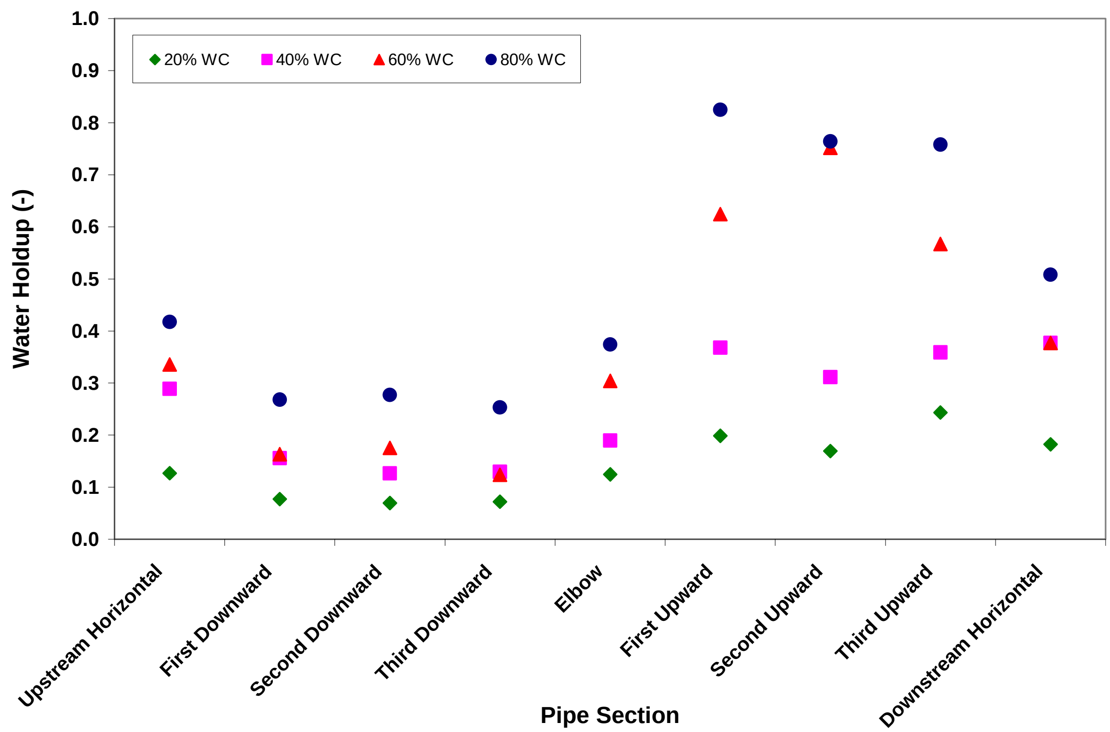

Figure 17: Water Holdup Values for v_{SL}=0.5m/s, v_{Sg}=0.1m/s

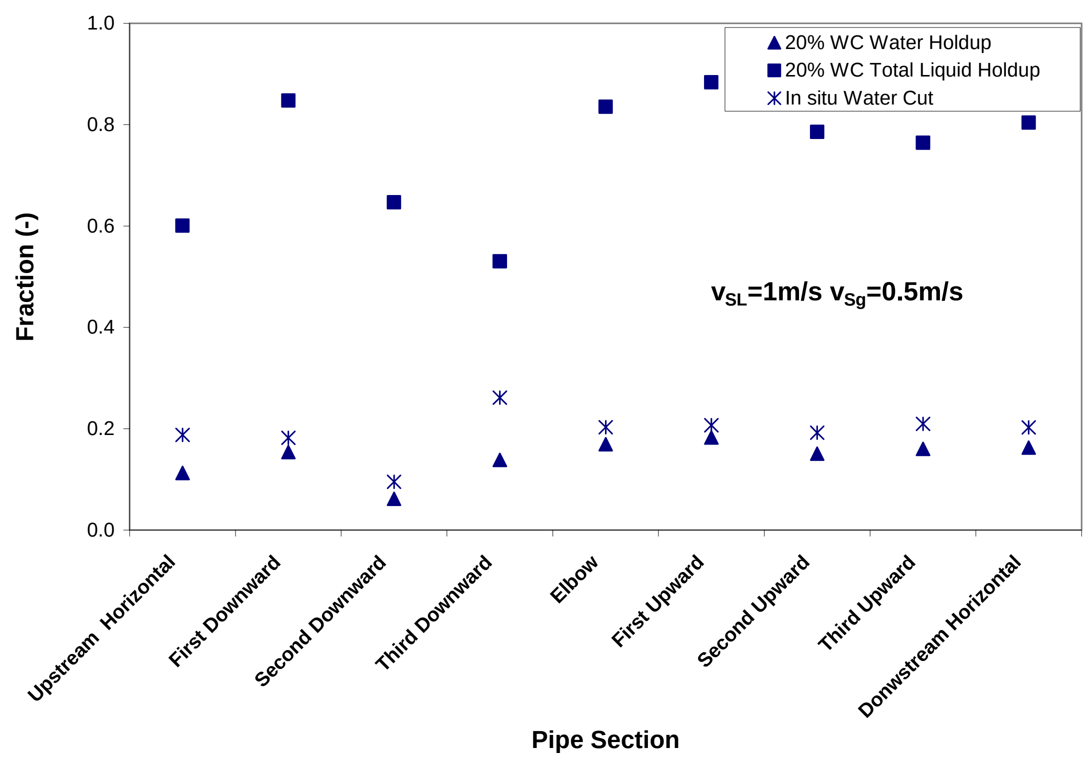

Figure 18: Liquid Fraction Values for v_{SL} = 1 m/s, v_{Sg} = 0.5 m/s at 20% Water Cut

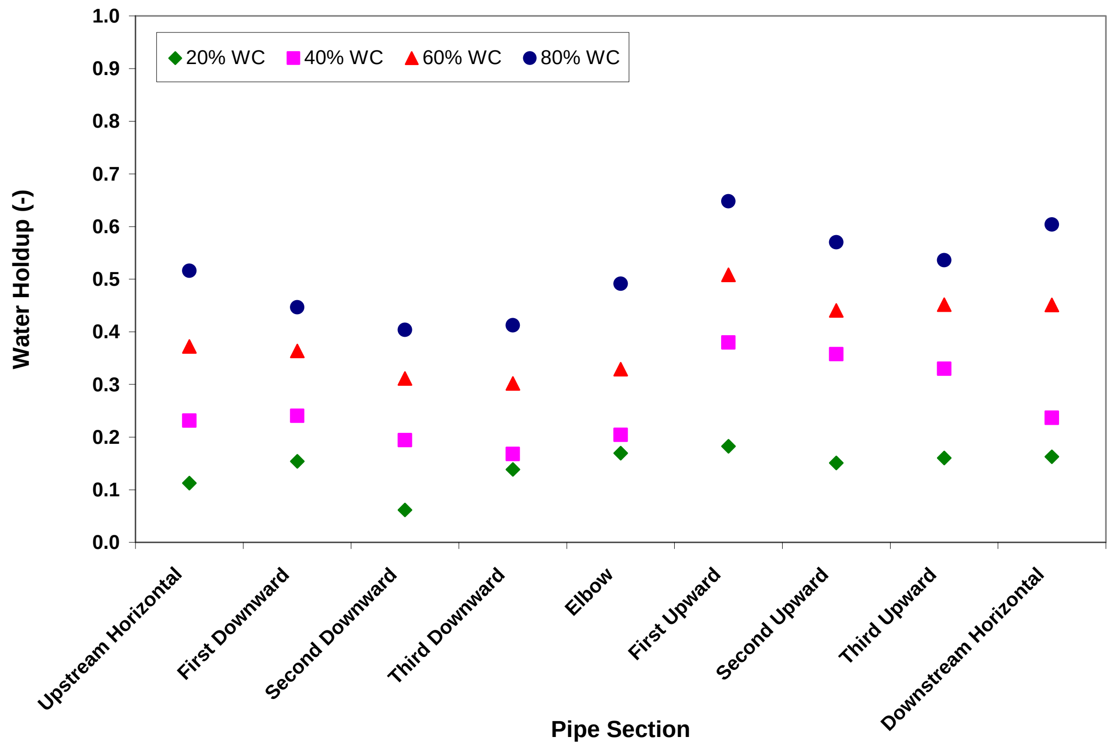

Figure 19: Water Holdup Values for v_{SL}=1m/s, v_{Sg}=0.5m/s

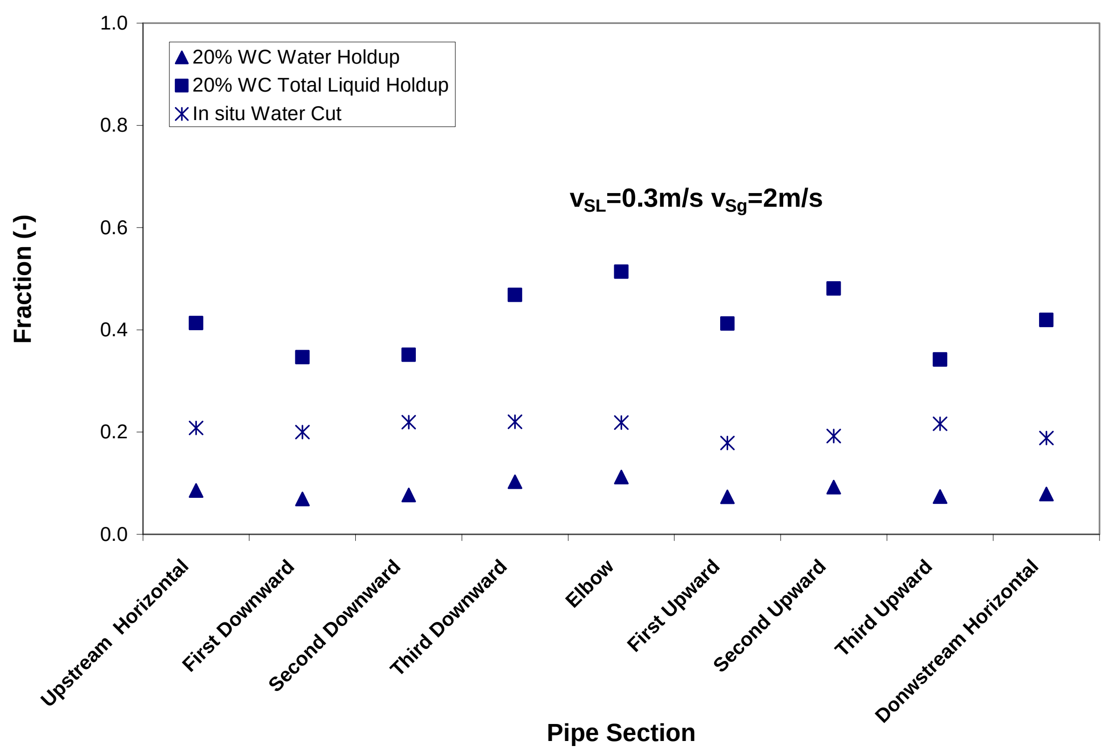

Figure 20: Liquid Fraction Values for v_{SL} = 0.3 m/s, v_{Sg} =2 m/s at 20% Water Cut

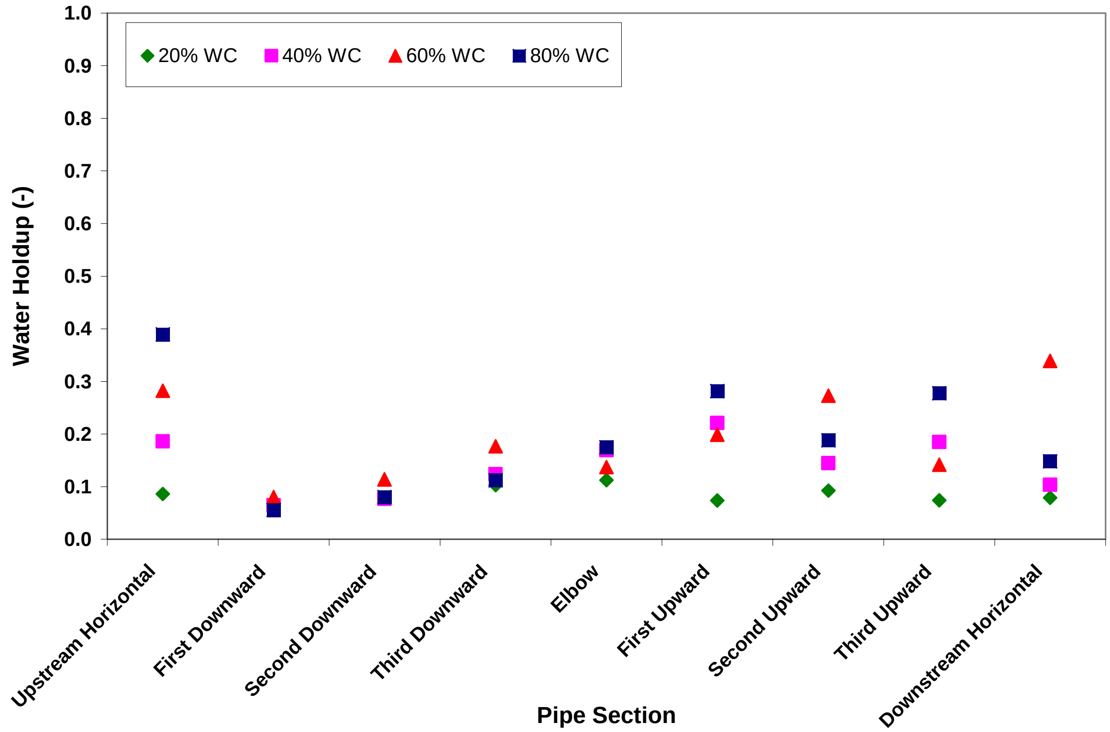

Figure 21: Water Holdup Values for v_{SL}=0.3m/s, v_{Sg}=2m/s

High Viscosity Multiphase Flow

- Significance
  - Discovery of High Viscosity Oil Reserves
- Objective
  - Development of Better Prediction Models
- Past Studies
  - First TUFFP Study by Gokcal (2005)
    - Existing Models Perform Poorly for Viscosities Between 200 and 1000 cp
    - Significantly Different Flow Behavior
      - Dominance of Slug Flow
  - Recent Study by Gokcal (2008)
    - New Drift Velocity and Translational Velocity Closure Models
    - New Slug Frequency Correlation

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High Viscosity Multiphase Flow …

- Current Study (Status)
  - Slug Liquid Holdup Closure Relationship Development
  - CFD Study on Drift Velocity
  - Slug Length Closure Relationship Development

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High Viscosity Multiphase Flow …

- Slug Liquid Holdup
  - Literature Review Mostly Complete
  - Liquid Holdup Measurement Methods are Under Study
    - Capacitance
    - Differential Pressure
    - Fast Acting Gamma-Densitometer
    - Quick Closing Valves

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High Viscosity Multiphase Flow …

- CFD Study on Drift Velocity
  - Dr. Al-Sarkhi Conducted CFD Experiments Using Fluent VOF
    - Results Agreed with Gokcal Experimental Observations
    - Benjamin Model Was Shown to Be Inaccurate
  - Next, Diameter Effect Will Be Investigated

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High Viscosity Multiphase Flow …

- Slug Length Study
  - Shorter Slug Lengths are experimentally Observed
  - Detailed Probabilistic Slug Length Study is Underway

Fluid Flow Projects

Effects of High Oil Viscosity on Slug Liquid Holdup in Horizontal Pipes

Ceyda Kora

Advisory Board Meeting, March 25, 2009

Outline

- Objectives
- Introduction
- Literature Review
- Experimental Facility
- Experimental Study
- Near Future Tasks
- Project Schedule

Objectives

- Investigate Slug Liquid Holdup for High Viscosity Oil and Gas Flow
- Develop Closure Models for Slug Liquid Holdup

Introduction

- Increase in Consumption of Hydrocarbon Resources
- Decline in Discoveries of Low Viscosity Oils
- Previous Studies Based on Low Viscosity Oils

Introduction

- Gokcal (2005, 2008) Study
  - Intermitted Flow Observed as Dominant Flow Pattern
  - Significant Effect of High Viscosity Oil on Slug Flow Characteristics Observed
  - TUFFP Unified Model Modified for High Viscosity Oil-Gas Flow

Literature Review

- Theoretical Studies
- Experimental Studies
- Liquid Viscosity Effect Studies

Theoretical Studies

- Kordyban (1961)
  - Mechanistic Model for Slug Flow
  - Assumed Slugs Glide Over Film
- Dukler and Hubbard (1975)
  - Better Slug Flow Description
  - Proposed Mathematical Model
- Barnea and Brauner (1985)
  - Developed a Mechanistic Model to Determine Slug Liquid Holdup

Theoretical Studies …

- Taitel and Barnea (1990)
  - Comprehensive Mechanistic Model for Steady State Slug Flow
- Xiao *et al.* (1990)
  - Comprehensive Mechanistic Model for Two-Phase Flow in Horizontal and Near Horizontal Pipelines
- Zhang *et al.* (2002)
  - Unified Mechanistic Model to Predict Slug Liquid Holdup Considering Dynamics of Slug Flow

Experimental Studies

- Gregory *et al.* (1978)
  - Empirical Correlation for Horizontal Flow
- Kouba (1986)
  - Applied Brauner and Barnea (1985) Model to His Experimental Data
- Felizola (1992)
  - Empirical Correlation for Inclined Pipes
- Gomez *et al*. (2000)
  - Dimensionless Correlation for Slug Liquid Holdup for Horizontal to Vertical Flow

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Liquid Viscosity Effect Studies

- Crowley et al. (1984)
  - Investigated Effect of Liquid Viscosity on Slug Flow in Horizontal and Near Horizontal Pipes
- Kago et al. (1985)
  - Developed Empirical Correlation for Liquid Holdup Considering Liquid Viscosity

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Liquid Viscosity Effect Studies …

- Nadler and Mewes (1995)
  - Investigated Changes of Liquid Slug Holdup with Increasing Liquid Viscosity
- Gokcal (2005, 2008)
  - Studied Effects of High Viscosity Oil on Average Liquid Holdup

Literature Review Summary

- Available Multiphase Flow Models Developed for Low Viscosity Liquids
- Entrained Gas Volume in Liquid Slug Ignored
- Few Studies Include Liquid Viscosity Effect
- Limited Experimental Data on High Viscosity Oil

Experimental Facility

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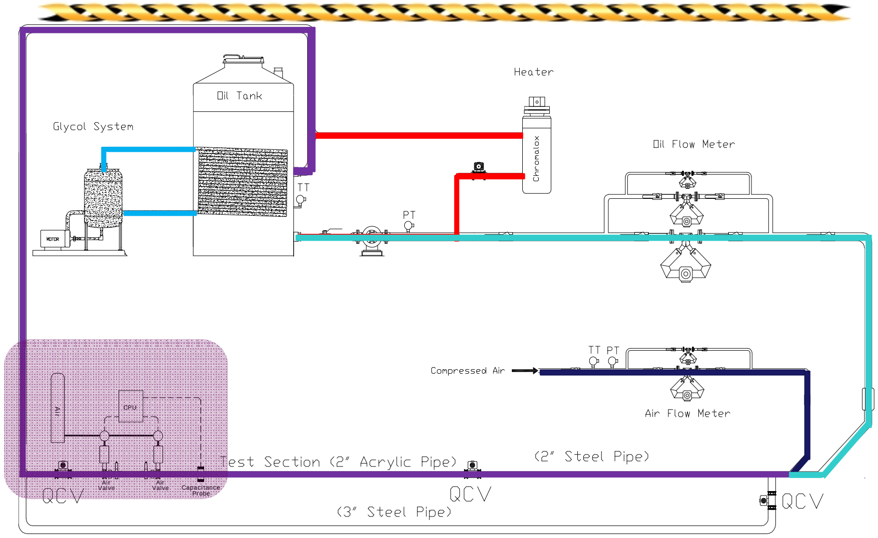
Experimental Facility



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Test Fluids

- Citgo Sentry 220
  - Mineral Oil
  - °API Gravity: 27.6
  - Viscosity: 0.22 Pa·s @ 40 °C
  - Specific Gravity: 0.89 @ 25 °C
- Air

Previous Measurement Techniques

- Impact Probe
  - Dukler and Hubbard (1975)
- Capacitance Sensor
  - Gregory et al. (1978)
  - Kouba (1986)
  - Felizola (1992)

Previous Measurement Techniques …

- Multibeam Gamma Ray Densitometer
  - Nadler and Mewes (1995)
- Conductance Probe
  - Andreussi and Bendiksen (1989)
  - Azzopardi (2004)

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Previous Measurement Techniques …

- Wetted Perimeter
  - Gokcal (2005, 2008)
- Hot-Film Anomometry
  - Wael et al. (2006)

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Considered Techniques

- In-House Capacitance Sensor
- Differential Pressure Sensor
- Fast Acting Gamma-Densitometer
- Quick Closing Valve System

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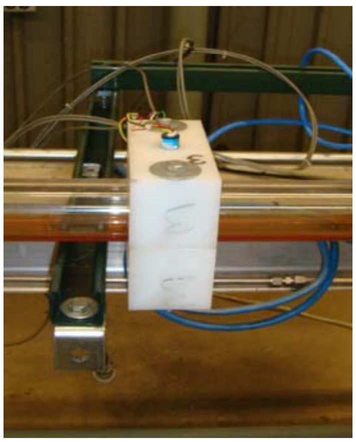
In-House Capacitance Sensor



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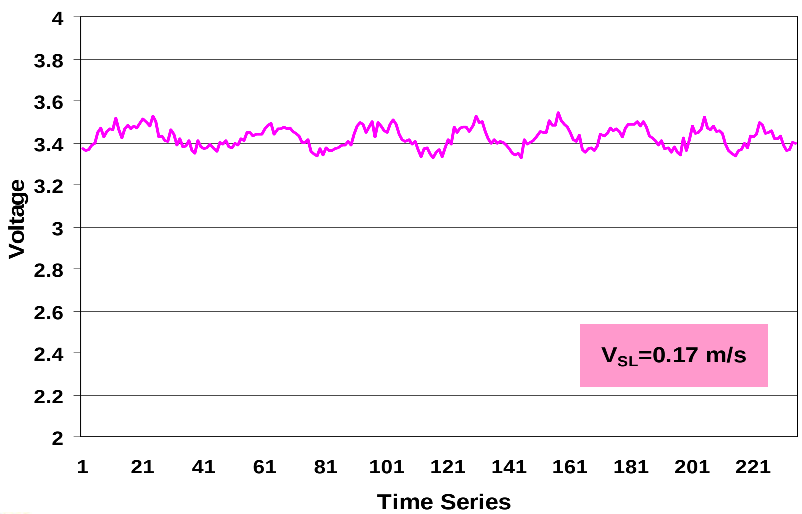
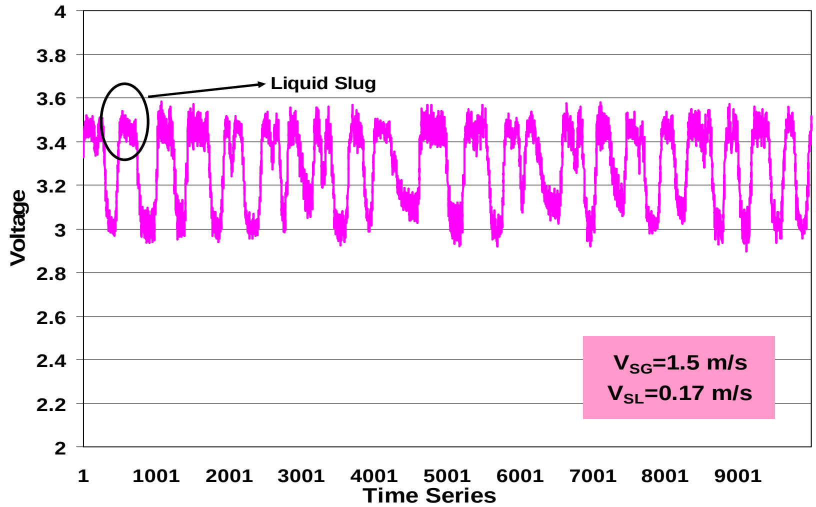
In-House Capacitance Sensor …

- Capacitance Sensor
  - Copper Ring, Electronic Circuit, Housing
  - Used for Translational Velocity Measurements

Capacitance Sensor Response to Single-Phase Oil Flow



Capacitance Sensor Response to Two-Phase Slug Flow



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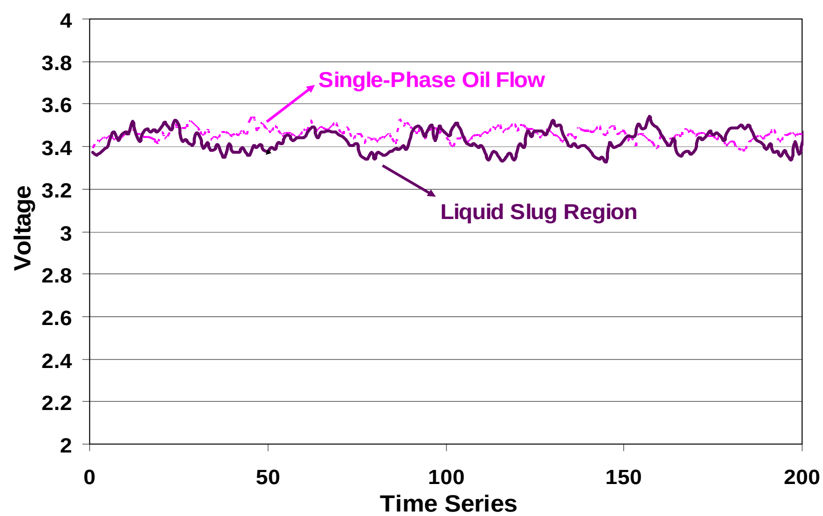
Capacitance Sensor Response to Liquid Slug Region



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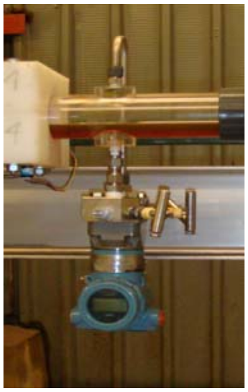
In-House Capacitance Sensor …

- No Significant Difference Observed in Output Signals Between Single Phase Oil Flow and Liquid Slug Region
- Dynamic Calibration Needed

Differential Pressure Sensor



Differential Pressure Sensor …

- Differential Pressure Sensor
  - Already Present in Facility
  - Theoretically Proper for Slug Liquid Holdup Measurements

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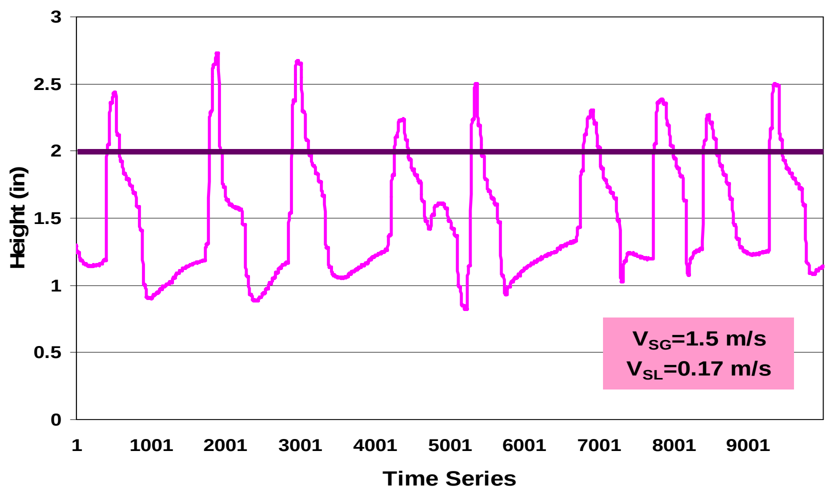
Differential Pressure Sensor Response to Slug Flow



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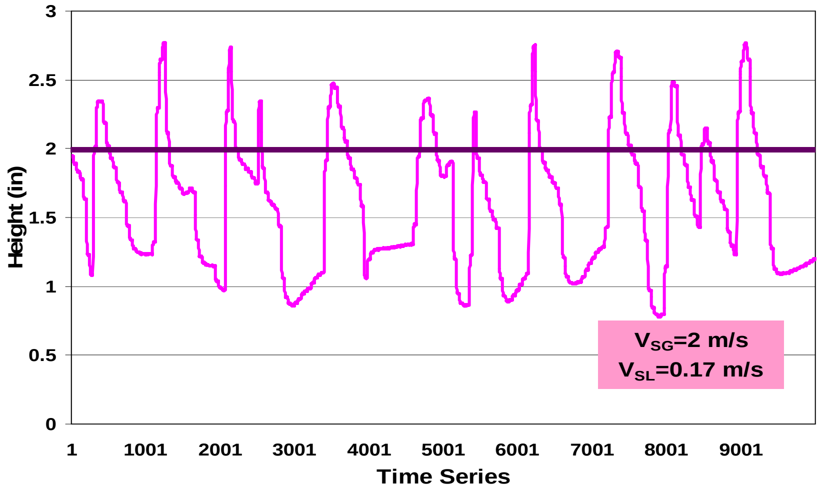
Differential Pressure Sensor Response to Slug Flow …



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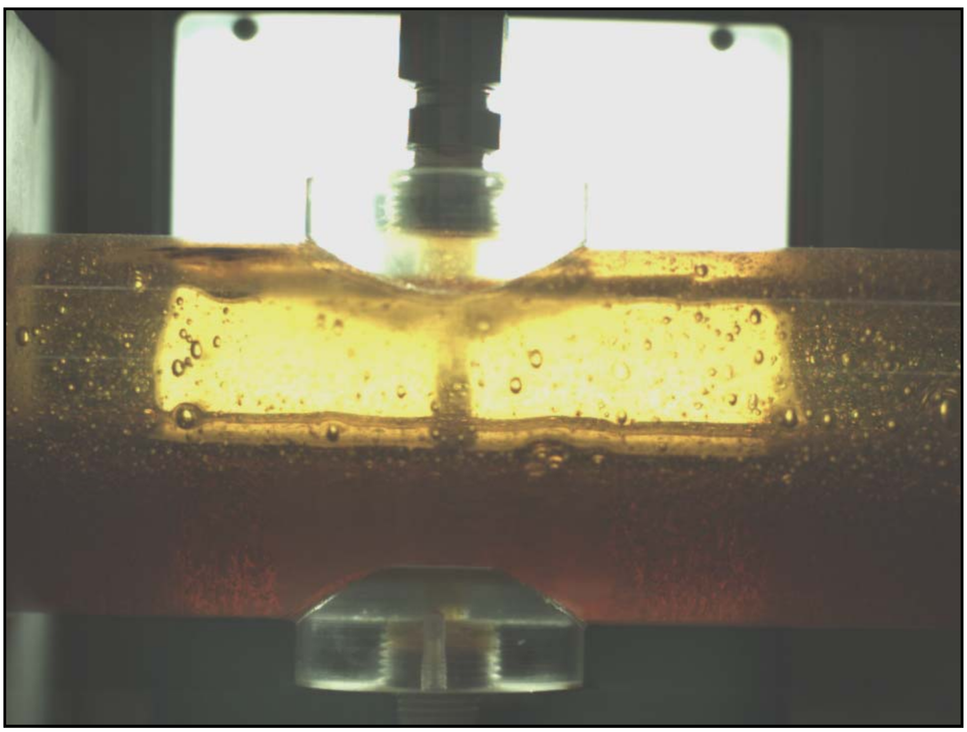


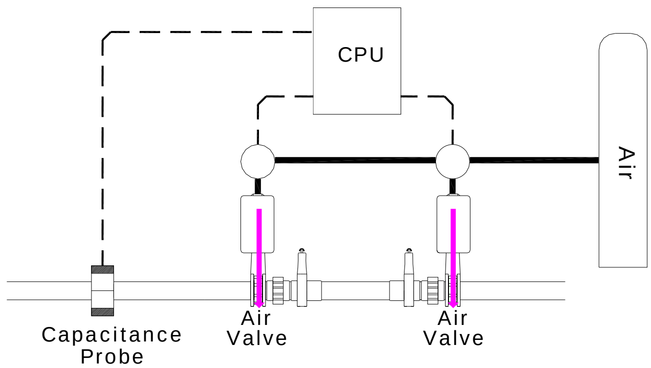
Differential Pressure Sensor …

- Calibration Problems due to Capillary Effect in Impulse Line
- Modification Needed for Impulse Line

Fast Responding Gamma-Densitometer

- IFE has been Contacted
  - Needs Significant Construction Time
  - Costly, i.e., over $120,000

Quick-Closing Valve System



Quick-Closing Valve System …

- Advantages
  - Non-Intrusive
  - Easy to Construct
  - Affordable
- Disadvantages
  - Proper Trapping of Liquid Slug
  - Labor Intensive

Near Future Tasks

- Completion of Facility Modifications for New Quick-Closing Valve System
- Conduct Preliminary Experiments with Quick-Closing Valve System
- Evaluation of Acquired Data From New Quick-Closing Valve System
- Improve and Calibrate Differential Pressure Sensor And Capacitance Sensor

Project Schedule

Literature Review	May 2009
Facility Modifications	May 2009
Preliminary Testing	June 2009
Testing	September 2009
Data Evaluation	November 2010
Final Report	May 2010

Questions & Comments

Fluid Flow Projects Advisory Board Meeting, March 25, 2009

Effects of High Oil Viscosity on Slug Liquid Holdup in Horizontal Pipes

Ceyda Kora

PROJECTED COMPLETION DATES:

Literature Review .... May 2009
Facility Modifications ....May 2009
Preliminary Testing ....June 2009
Testing ....September 2009
Data Evaluation ....November 2009
Final Report ....….May 2010

Objectives

The main objectives of this study are,

- Investigation of slug liquid holdup for high viscosity oil and gas flow,
- Development of closure models for slug liquid holdup.

Introduction

High viscosity oils are produced and transported from many fields all over the world. Because of the increased consumption of hydrocarbon resources and decline in discoveries of low viscosity oils, the importance of high viscosity oil has increased. It is important to design a proper production system in order to eliminate operational problems for high oil viscosity fields. Available multiphase flow models are primarily developed for low viscosity liquids. TUFFP has been studying the high viscosity oil multiphase flow in a systematic way since 2005 and has made significant progress towards the improvements in high viscosity oil multiphase flow prediction.

The first experimental study at TUFFP on high viscosity oil was completed by Gokcal (2005). The effects of high oil viscosity on oil-gas two-phase flow behavior were investigated and significant changes in flow behavior were encountered. Gokcal (2005) observed intermitted flow (slug and elongated bubble) as the dominant flow pattern for high viscosity oil and air flow. Slug characteristics should be examined in detail for better understanding of high liquid viscosity effect. In particular, determination of the liquid holdup in the slug body is important for calculating pressure drop.

Previous models were based on experimental data with low viscosity oils. Gokcal (2005) compared his experimental data against Xiao (1990) and TUFFP Unified (2003) models. The results revealed that these models do not adequately perform for high viscosity oils. After this study, two modifications were made to improve the TUFFP Unified Model. In the Unified Model, the slug liquid holdup prediction was based on a turbulent flow assumption. However, mostly laminar flow was observed in the liquid phase for high viscosity oil-air two-phase flow. The first modification was made in translational velocity calculation by adding a laminar flow option in the model. Secondly, the momentum term in the slug liquid holdup model was modified considering the high viscosity flow behavior. If the Reynolds number is less than 5000, the momentum term for gas entrapment is multiplied by Re/5000. Although the TUFFP Unified model gives better results after these modifications, additional improvements are needed from experimental and modeling studies on slug characteristics.

An experimental and theoretical investigation of slug flow for high oil viscosity in horizontal pipes was completed by Gokcal in 2008. Gokcal (2008) developed models for drift velocity, transitional velocity and slug frequency by taking into account the viscosity effect. Slug liquid holdup was not studied due to a lack of proper instrumentation. Gokcal was only able to measure average liquid holdup in his study. A new investigation of slug liquid holdup for high viscosity oils is needed.

Literature Review

There has been significant research reported in the literature on slug holdup. Most of the studies focused on low oil viscosity. There exist some efforts investigating the effects of viscosity on slug liquid holdup. Following is a summary of the literature review to date in three categories: theoretical, experimental and liquid viscosity effect studies. The literature review will be an ongoing effort.

Theoretical Studies

Kordyban (1961) presented the earliest mechanistic model for horizontal gas-liquid slug flow. In this model, the slug velocity is strongly dependent on the gas velocity and the slug is assumed to slide over the top of the film without mixing at the slug front. This mechanistic model was followed by an experimental study of Kordyban and Ranov (1963). Their experimental data was not in good agreement with the model. Kordyban's homogenous liquid slug assumption might be the reason for this discrepancy. Gas entrainment can not be neglected, especially at high gas flow rates.

Dukler and Hubbard (1975) presented a comprehensive study of horizontal gas-liquid slug flow. This was the first accurate description of the flow for each part of the slug unit. They proposed a mathematical model to calculate the pressure drop in a high velocity slug section by using the in-situ gas-liquid mixture density, taking into account the gas fraction in the slug. However, they were not able to measure the liquid holdup in the slug. This did not affect their model because the predicted results were not sensitive to the liquid fraction in the slug. The liquid holdup was used as an input value to their model.

Barnea and Brauner (1985) published a mechanistic model to determine liquid holdup in a slug for horizontal and vertical pipes. Only the influence of the mixture velocity on liquid slug holdup was taken into account. They assumed that gas in developed liquid slug resembles dispersed bubbles and the liquid slug will have the same liquid holdup as fully dispersed bubble flow moving with the same mixture velocity. The gas carrying capacity of the liquid slug was determined by analyzing the turbulent forces and buoyancy forces due to gravity or surface tension.

Taitel and Barnea (1990) proposed a comprehensive mechanistic model for steady state slug flow. The slug holdup part of this model was based on a mass balance and they stated that liquid and gas velocities in the slug are not the same. An average liquid holdup formula for a slug unit was presented based on the combination of liquid holdup in the slug and the liquid film. However, this formulation was not verified by experimental data. Empirical formulas were used to calculate slug liquid holdup. Moreover, they did not consider any fluid property effects in their model.

Xiao *et al.* (1990) presented a comprehensive mechanistic model for gas-liquid slug flow in horizontal and near-horizontal pipes. A pipeline data bank which contains 426 field and laboratory data sets was used to compare measured and predicted pressure drop values. The data bank did not provide experimental liquid holdup values; instead, the Gregory *et al. (1978)* empirical correlation was used to calculate liquid holdup in the slug, which is an input for their model.

Zhang *et al.* (2002) developed a unified mechanistic model to predict slug liquid holdup considering the overall dynamics of slug flow. Slug flow characteristics were taken into account while solving momentum and continuity equations for slug flow. In this study, TUFFP data were used to check the validity of the slug liquid holdup model at different inclination angles. In all these experiments, kerosene and air were used as test fluids. The experimental data and the predicted results of the model generally fit well for low viscosity oils.

Experimental Studies

Gregory *et al.* (1978) conducted experiments on liquid holdup in slugs with two different horizontal diameters pipe: 2.58 cm and 5.12 cm. They proposed an empirical correlation for the liquid slug holdup. In this correlation, they assumed the slug to be homogenous. However, they suggested that the gas fraction in a slug changes with position. Moreover, they mentioned that this correlation has to be modified to include the effects of fluid properties.

Kouba (1986) applied the Brauner and Barnea (1985) model to his liquid holdup data and observed that the slug liquid holdup is not only a function of mixture velocity, but there is also a linear relationship between the weighted mean liquid velocity in the slug and slug liquid holdup. Kouba (1986) used air as the gas phase and 1.5 mPas and kerosene as the liquid phase for his experiments.

Felizola (1992) measured average liquid holdup in the slug for pipe inclinations of 0° to 90° degrees from horizontal. He developed a new empirical correlation for slug liquid holdup which depends on mixture velocity and pipe inclination.

Gomez *et al*. (2000) proposed a new dimensionless correlation for liquid holdup in the slug body for horizontal to upward vertical flows. This correlation included the mixture velocity, liquid viscosity, pipe diameter, and inclination angle. They used the slug liquid holdup experimental data of Kouba (1986), Rothe et al. (1986), Brandt and Fuches (1989), Kokal (1987), Felizola (1992), Schmidt (1977) to develop their correlation. They established their empirical correlation by considering inclination angle and Reynolds number. Although this empirical correlation considers the liquid viscosity through a Reynolds number, it was not validated for high viscosity oils. All experimental data used in this study were conducted with low oil viscosity.

Liquid Viscosity Effect Studies

Crowley *et al.* (1984) studied the effect of liquid viscosity on slug flow in horizontal and near horizontal pipes. Water, glycerin-water and water-polymer solutions were used as the liquid phase, and air and Freon 12 were used as the gas phase in a 171-mm inner diameter pipe. They concluded that, as liquid viscosity increases, average liquid holdup increases.

Kago *et al.* (1985) developed an empirical correlation for gas holdup. Experiments were conducted in the slug flow pattern with a horizontal loop constructed with an inner diameter of 51.5 mm. Water, water polymers and slurries were chosen as the liquid phase and air for the gas phase. The viscosity ranged from 0.0008 Pa.s (water) to 0.055 Pa.s (slurry). Kago *et al*. stated that when liquid viscosity increases 45-55 times, the liquid holdup increases 50-60%.

Nadler and Mewes (1995) investigated the changes in slug holdup with increasing liquid viscosity. Their experiments were carried out in a 59-mm inner diameter horizontal pipe. For two–phase air-liquid slug flow, air, water and oil with viscosities in range from 0.014 to 0.037 Pa.s were used. After comparing the liquid holdup data of oil-air and water-air slug flows, considerable difference was observed especially in the slug unit and the film zone. Nevertheless, only a slight difference was observed in the liquid slug. Since, void fraction in the liquid slug could not be determined accurately, they ignored entrained bubbles in the liquid slug.

Gokcal (2005 and 2008) studied the effects of high viscosity oil on average liquid holdup. A 50.8-mm inner diameter horizontal high viscosity facility was used to conduct the experiments. Air was selected for the gas phase and mineral oil with viscosities of 0.587, 0.378, 0.257 and 0.181 Pas were used as the liquid phase. Even though significant change in liquid holdup could not be measured with the existing instrumentation, the change was visually observed as oil viscosity increases. The entrained gas volume in the slug increased with increasing liquid viscosity. Moreover, entrained gas bubbles in the liquid slug have a different size distribution, and the gas amount is not negligible as it is in low viscosity liquids.

In general, few studies include oil viscosity effects in empirical correlations. It is necessary to develop a new closure model for slug liquid holdup that include effects of liquid viscosity. Moreover, almost none of the slug liquid holdup models were adequately validated with high viscosity oil experimental results. Therefore, it is critical to acquire high quality slug liquid holdup data with high viscosity oil.

Experimental Study

Facility

An existing indoor TUFFP high viscosity facility will be used for this study (Fig. 1). This facility was previously used by Gokcal (2005 and 2008) to investigate the effects of high oil viscosity on slug flow characteristics. The previously used high viscosity oil (Citgo Sentry 220) was selected again for this study. Following are typical properties of the oil:

Gravity: 27.6 °API
Viscosity: 0.220 Pa·s @ 40 °C
Density: 889 kg/m3 @ 15.6 °C

The oil viscosity vs. temperature behavior is shown in Fig. 2.

There are four main parts of the facility: metering section, test section, heating system and cooling system. The test section was designed as an 18.9-m (62-ft) long, 50.8-mm (2-in.) ID pipe including a steel pipe section and a transparent acrylic pipe section. A 9.15-m (30-ft) long transparent acrylic pipe section is used to observe the flow behavior visually. This section is connected to a 76.2-mm (3-in.) ID return pipe with a flexible hose. The return pipe sends the oil back to the oil storage tank. The oil is pumped by a 20-hp screw pump from the oil storage tank. A dry rotary screw air compressor delivers compressed air to the system. Before entering the test section, two fluids were mixed at a mixing tee. Micro Motion™ mass flow meters are used to meter the mass flow rates and densities of oil and air. There is no special separation system. Air and oil are gravity segregated in the oil tank, and separated air is released to the atmosphere through a ventilation system. The inclination of the test section can be set from -2° to 2° from horizontal by adjusting the heights of the stands.

The test oil viscosity is very sensitive to temperature changes. Therefore, it is crucial to conduct experiments at a constant temperature. Existing heating and cooling systems will be used to control temperature. Resistance Temperature Detector (RTD) transducers already exist in the facility to measure temperatures during experiments. Pressure transducers and differential pressure transducers are located at various points to monitor the pressure and pressure drop during experiments.

The most challenging part of this study is to measure the gas void fraction in liquid slugs. In previous studies, several instruments and methods were used, such as impact probes, quick-closing valves, conductance sensors, capacitance sensors, hot-film anemometry and gamma ray densitometers.

Previous Measurement Techniques

Dukler and Hubbard (1975) used an impact probe to measure liquid slug holdup. A 1/8-in impact tube, filled with water was connected to two pressure transducers. When slug passed through, an abrupt increase in momentum flux and, therefore, an impact pressure was recorded. Significant fluctuations in impact pressure were observed with time. Therefore, they were not able to measure the liquid holdup in slugs.

Gregory *et al.* (1978) used in-situ capacitance sensors, designed by Gregory and Mattar (1972) to acquire data for liquid holdup in a slug in horizontal pipes. Kouba (1986) and Felizola (1992) also measured instantaneous liquid holdup with parallel ring and helical wrap capacitance sensors. A quick-closing valve system was used for dynamic calibration of these capacitance sensors. After trapping the air-kerosene mixture, the mixture was drained and actual liquid holdup is calculated based on the drained volume for low viscosity oil.

Nadler and Mewes (1995) used a multibeam gamma ray densitometer to measure the liquid holdup in the film zone and slug, and the average liquid holdup. After conducting static and dynamic tests, it was declared that the absolute error for average liquid holdup is less than 3%. They concluded that results were limited, and more investigations were needed to study viscosity effects in the slug zone.

Andreussi and Bendiksen (1989) measured mean void fraction in water-air slug flow with a conductance probe. Azzopardi (2004) also used conductance ring probes and electrical capacitance tomography to measure void fraction in a liquid slug. Water and air were selected as two-phase fluids for these experiments. The conductivities of high viscosity oil and air are very close to each other. Therefore, it is not possible to distinguish air in oil with a conductivity sensor. Due to the nonconductive properties of these fluids, this method can not be applied for this study.

Gokcal (2005 and 2008) measured average liquid holdup using a graduated tape mounted outside of the pipe after trapping the gas-oil mixture between quick-closing valves. It was not possible to measure the slug liquid holdup with this technique.

Wael et al. (2006) tried hot-film anemometry to measure averaged local profiles of void fraction of air and low viscosity oil. This instrument can distinguish the elongated bubble and liquid slug region. However, they ignored the void fraction in the liquid slug.

Current Effort

A capacitance sensor and a differential pressure sensor developed in-house were tested since the last ABM. Unfortunately, the responses of these instruments were not satisfactory with the current designs. It is believed that, with proper change in design of both devices liquid holdup in the slug can be measured to the desired accuracy. The following summarizes our efforts to use existing capacitance and differential pressure sensor devices.

A few tests were conducted to check the accuracy of the capacitance sensor. Static calibrations were done for a full pipe and an empty pipe. The test section was filled with test oil and voltages were recorded from the capacitance sensor. The same procedure was repeated for an empty pipe. A calibration curve was created with these recorded voltages. Slugs were than flowed with different gas-oil superficial velocities. For each test, the capacitance sensor was used to measure slug liquid holdup. No significant differences in the output data were observed during slug flow. Capacitance sensor responses were plotted for single phase oil flow and for slug conditions (Figs. 3-5).

Secondly, the feasibility of measuring differential pressure drop measurement across the cross section of the pipe was tested for liquid holdup measurement. To our knowledge, this method has not been tried before to measure slug liquid holdup in gas-oil two-phase flow. Theoretically, this method is very promising. However, some mechanical problems with the impulse lines were encountered during tests.

Calibration of the differential pressure sensor was performed by recording the full pipe and empty pipe voltages and creating a calibration curve with respect to these reference voltages. After converting voltage signals to the liquid height, it was observed that, for the higher gas flow rates, the converted heights are larger than the pipe diameter, which is physically impossible (Figs. 6-7). Initially, the reason for this was thought to be a capillary effect in the impulse line. The impulse line was changed to a larger diameter one to reduce the capillary effect. Unfortunately, the increase in impulse line diameter did not affect the results. The flow was recorded with a high speed video system to identify the problem at the point where differential pressure sensor is placed. Drainage of oil and penetration of gas bubbles into the impulse lines were observed while slugs were passing. In order to solve this problem, a differential pressure sensor with a capillary system and preventing impulse line liquid drainage can be tried in future.

As another option, following Scand Power Petroleum Technology (SPT)'s suggestion, Institute for Energy Technology (IFE) has been contacted to construct a fast responding gamma-densitometer for this project. They require a very long construction time for a new gamma-densitometer and the cost of construction is very high compared to other methods. Moreover, the accuracy of this method is still questionable and needs to be verified with an independent method.

All of the methods considered have some issues and should be checked with a non-intrusive method to reduce the uncertainty of the measurements. For this purpose, a new separable quick closing valve system was designed for this study. Gokcal (2008) proposed that the slug length for high viscosity oil is changing between 8D-13D. Therefore, the optimum distance between two valves was determined as 12 inches in order to capture only the liquid slug. A dedicated CPU will be developed for this system. A capacitance sensor will be placed before the first valve. When the capacitance sensor detects the slug front, after a pre-determined time period, the dedicated CPU will close the quick-closing-valves. Since the drainage of high viscosity oil is a slow process, the trapped section will be separated and weighed to measure the liquid holdup. A by-pass system will be added to the system for continuity of flow. A schematic view of this system is shown in Fig. 8.

The challenging part of this method is trapping only the liquid slugs none of the film region or elongated bubble. In order to control what is trapped in this section, a high speed video system will be used. Before using the data, videos will be checked. Although this method needs a significant amount of time for each experiment, it is worth using since the result of this method is a direct measurement. Moreover, the results can be used to study and calibrate other instruments.

Near Future Tasks

The main tasks for the future are:

- Completion of facility modifications for the new quick-closing valve system,
- Preliminary experiments with quick-closing valve system,
- Evaluation of the acquired data from new quick-closing valve system.
- Improvement and calibration of differential pressure sensor and capacitance sensors.

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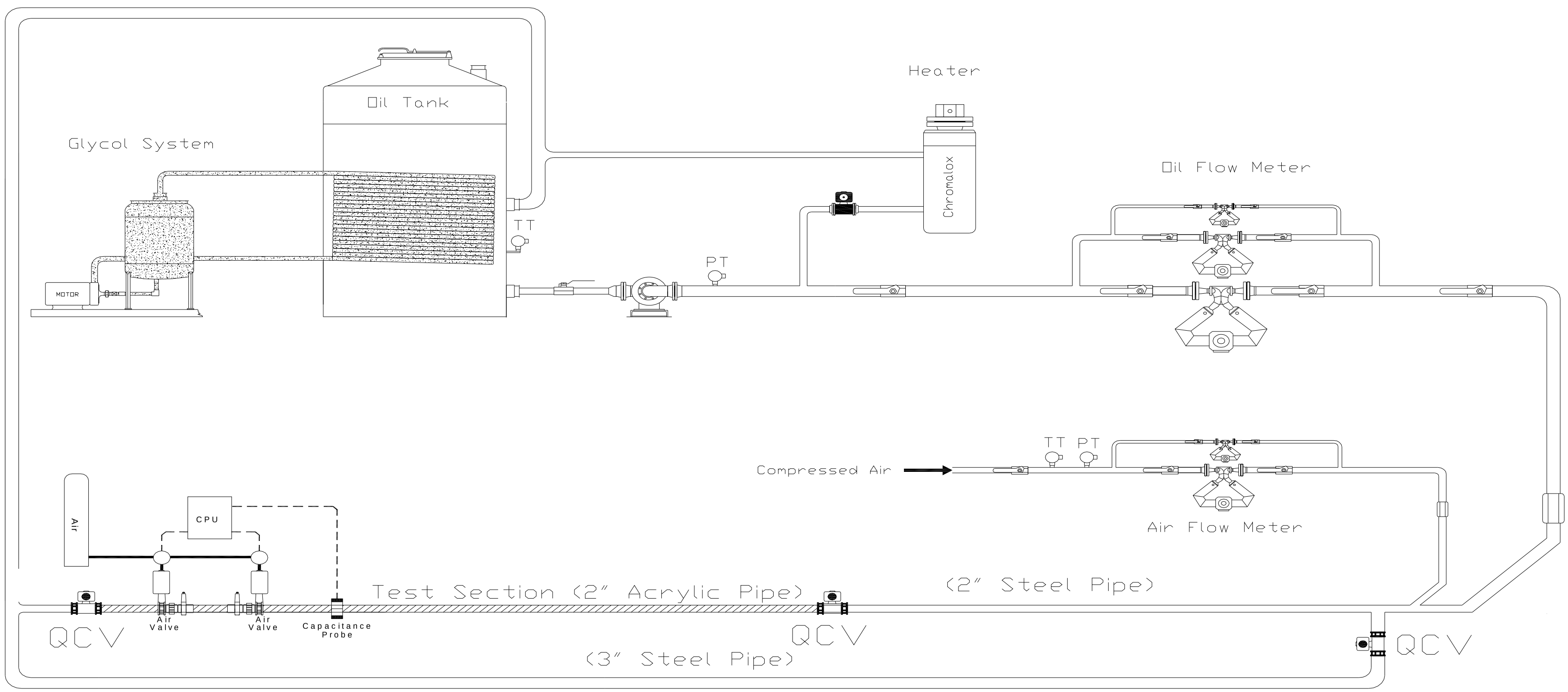


Figure 1: Schematic of High viscosity facility of the Tulsa University Fluid Flow Projects (TUFFP)

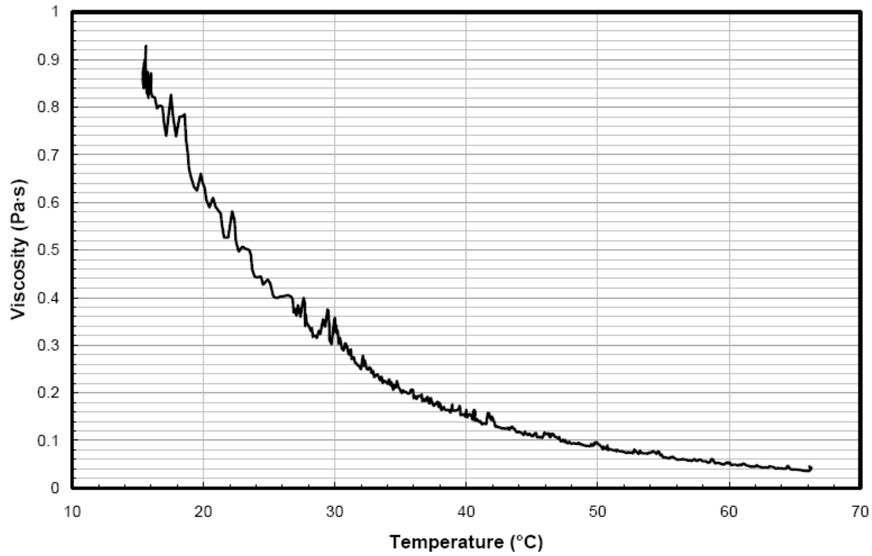


Figure 2: Viscosity vs. Temperature for Citgo Sentry 220 Oil

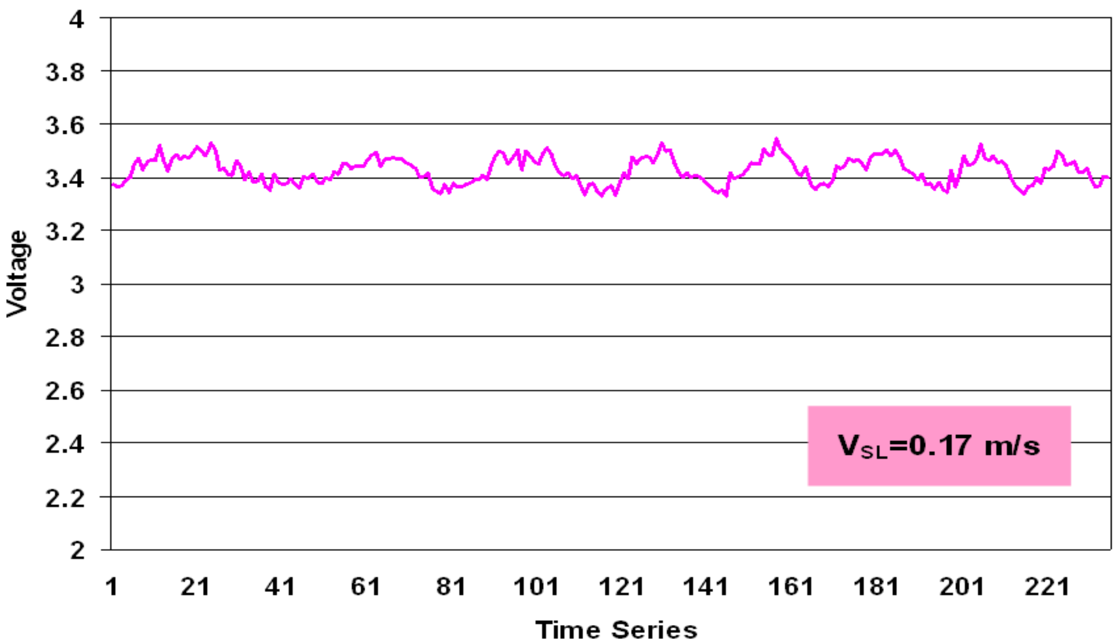


Figure 3: Raw Output Signal of Capacitance Sensor (Single Phase Oil Flow)

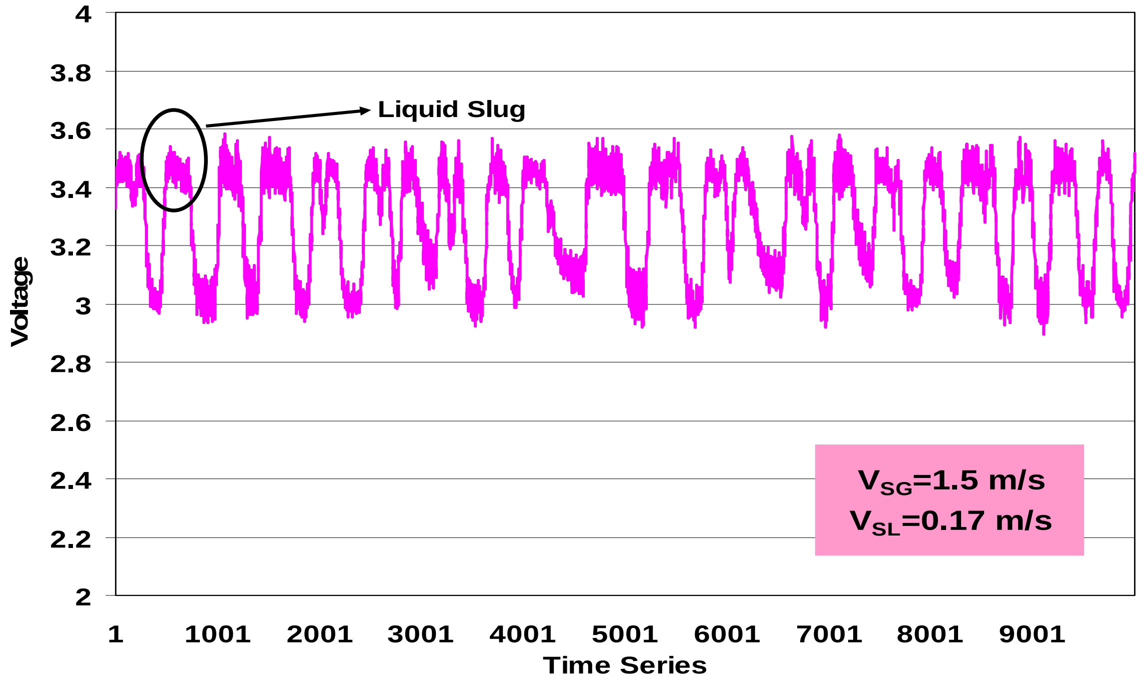


Figure 4: Raw Output Signal of Capacitance Sensor for Slug Flow

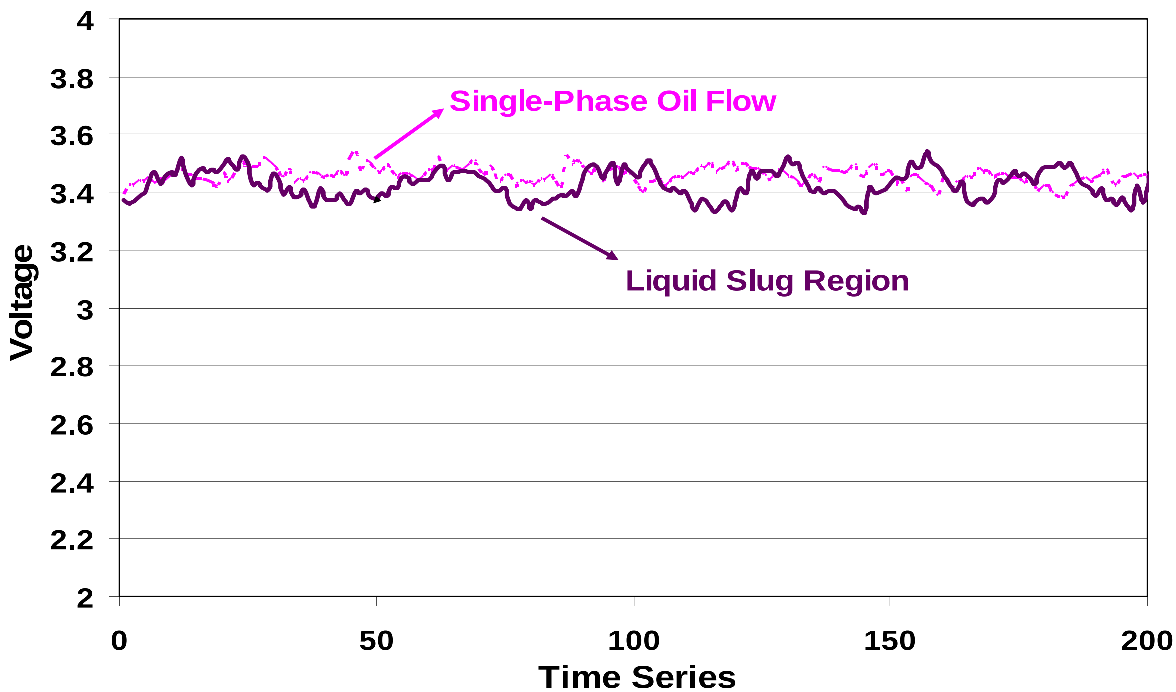


Figure 5: Raw Output Signal of Capacitance Sensor for Liquid Slug Region

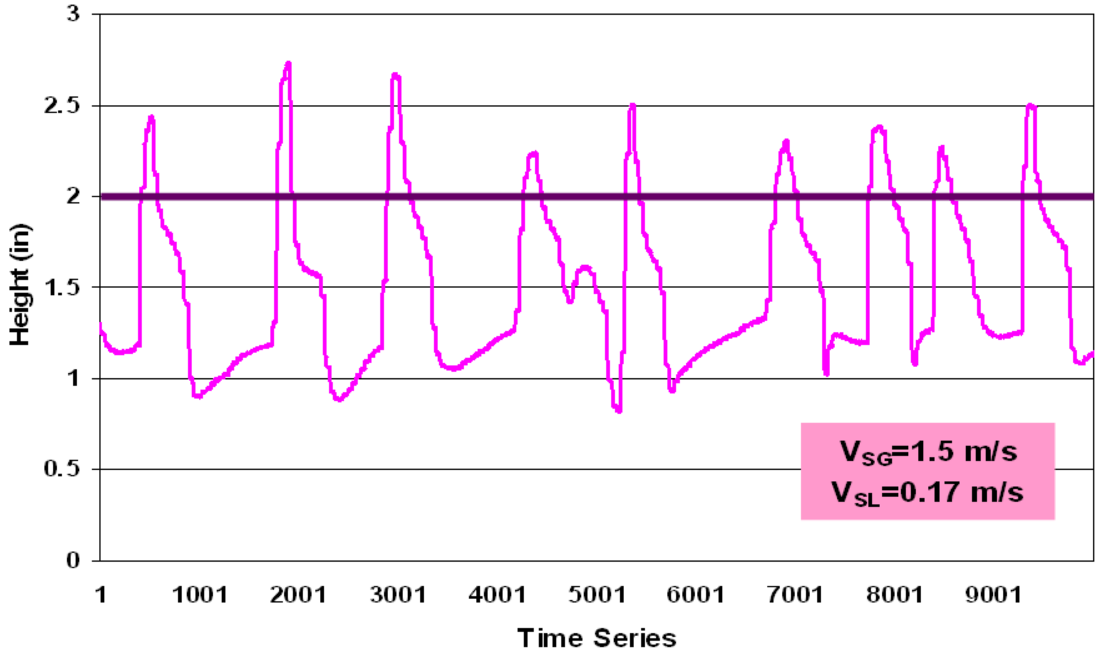


Figure 6: Raw Output Signal of Differential Pressure Sensor

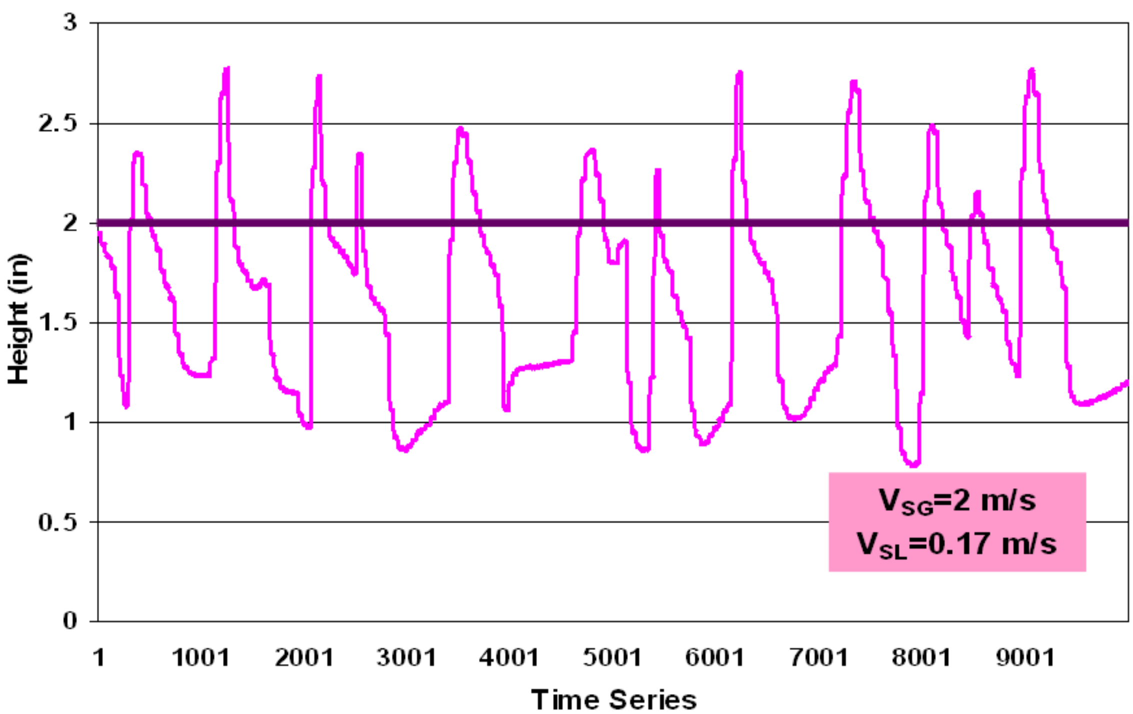


Figure 7: Raw Output Signal of Differential Pressure Sensor

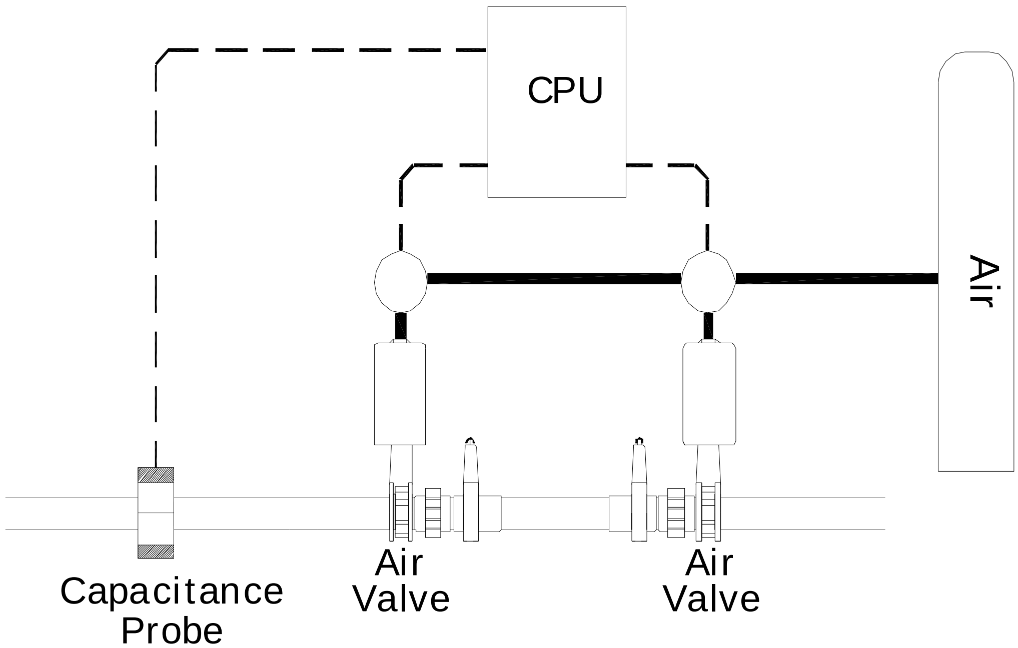


Figure 8: Schematic of New Quick-Closing Valve System

Fluid Flow Projects

CFD Modeling of Drift Velocity for High Viscosity Oil-Gas Two-Phase Horizontal Flow

Abdel Al-Sarkhi and
Cem Sarica

Advisory Board Meeting, March 25, 2009

Objectives

- Investigation of Drift Velocity Using Computational Fluid Dynamics (CFD)
- Verification of Previous Experimental and Modeling Studies
- Improvement of Drift Velocity Closure Relationship

Introduction

- Nicklin's Translational Velocity

$$v_t = C_s v_s + v_d$$

- Benjamin's Drift Velocity for Inviscid Flow in Horizontal Pipes

$$v_d = 0.542\sqrt{gD}$$

Introduction …

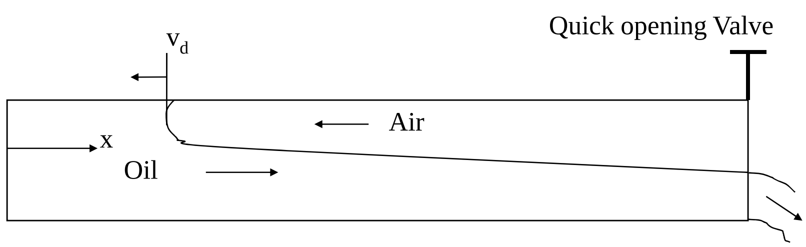
- Zukoski (1966)
  - Experimentally Investigated Liquid Viscosity, Surface Tension, and Pipe Inclination Effects
  - Concluded That Viscosity Effects are Negligible for Re_d>200

Introduction …

- Gokcal (2008)
  - Experimental and Modeling Study
  - Developed a Model Based on Benjamin's Approach
  - Drift Velocity was Related to h/D rather than Viscosity
  - Experimental Relationship between h/D and Viscosity then Provided the Drift Velocity Viscosity Relationship

CFD Study

- Drainage or Gas Penetration Problem

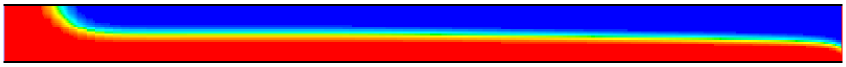


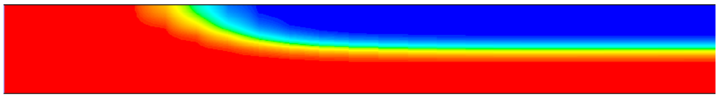
CFD Study …

- Fluent 6.1 Used
  - Volume of Fluid Model (VOF)
- VOF Method
  - Solves Single Momentum Equation

Results

- Water as Liquid

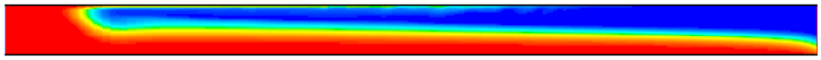




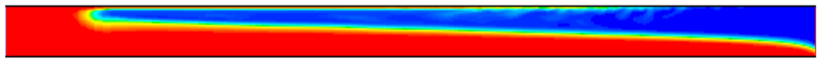
Contours of volume fraction
(Red color: 100% water, Blue color: 100% air.

Results …

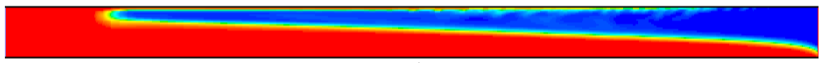
Two-Dimensional Results



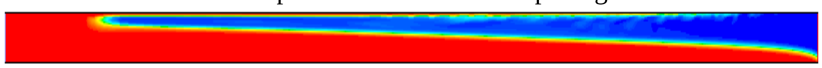
15 cp at t=5 s after valve opening



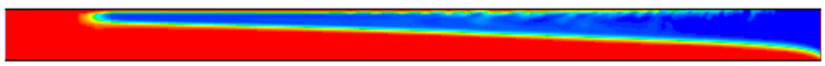
120 cp at t=6.5 s after valve opening



250 cp at t=8.5 s after valve opening



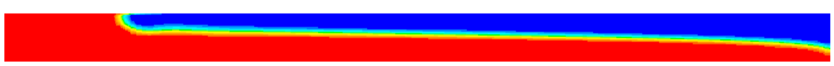
600 cp at t=16 s after valve opening



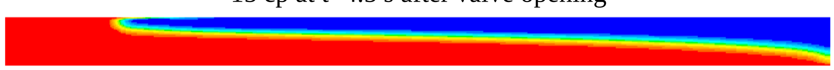
1200 cp at t=30 s after valve opening

Results …

Three-Dimensional Results

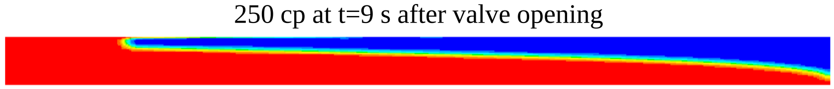


15 cp at t=4.5 s after valve opening

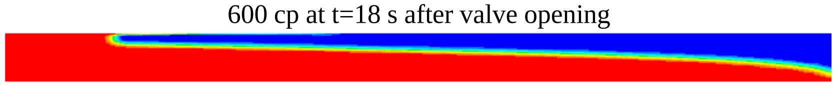


120 cp at t=7.5 s after valve opening

250 cp at t=9 s after valve opening

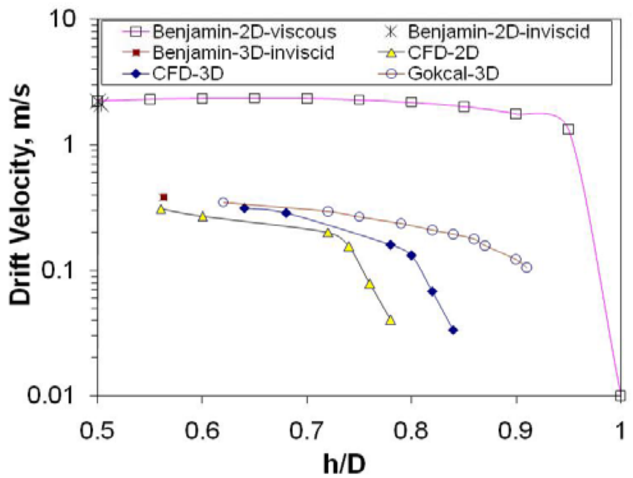


600 cp at t=18 s after valve opening



1200 cp at t=36 s after valve opening

Drift Velocity vs. h/D



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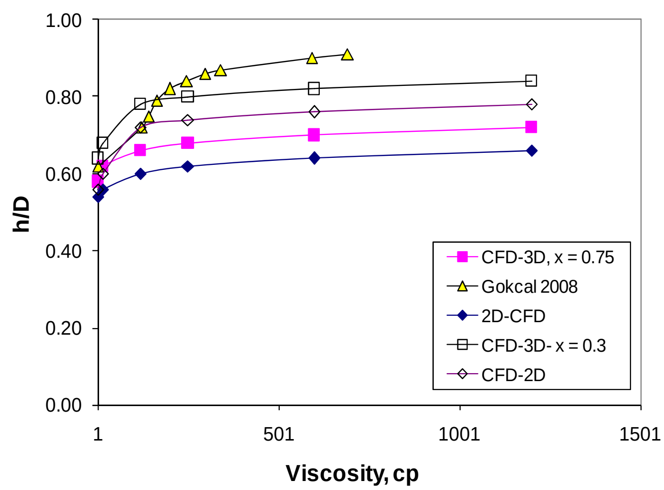
h/D vs. Viscosity



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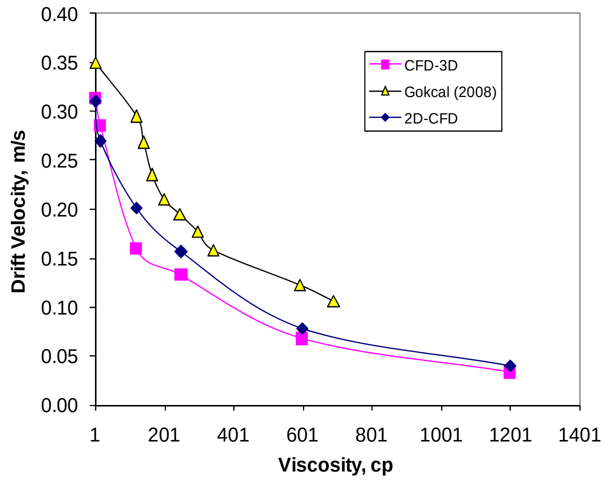
Drift Velocity vs. Viscosity



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Concluding Remarks

- CFD Simulation Results Agree with Gokcal (2008) Results
- Benjamin's (1968) Energy Loss Model Significantly Over Predicts Drift Velocity for High Liquid Viscosities
- Liquid Film Height Increases with Liquid Viscosity
- Liquid Height is not Constant for High Viscosity Liquids

Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

CFD Modeling of Drift Velocity for High Viscosity Oil Gas Two-Phase Horizontal Flow

Abdel Al-Sarkhi and Cem Sarica

Objective

The objective of this study is to investigate drift velocity for high viscosity oils using the Computational Fluid Dynamics (CFD) approach.

Introduction

High viscosity oils are produced from many oil fields around the world. Oil production systems are currently flowing oils with viscosities as high as 10000 cp. (10 Pa·s). Current multiphase flow models are largely based on experimental data with low viscosity liquids. Commonly used laboratory liquids have viscosities less than 20 cp. (0.020 Pa·s). Multiphase flows are expected to exhibit significantly different behavior for higher viscosity oils.

Gokcal et al. (2006) observed slug flow to be the dominant flow pattern for high viscosity oil and gas flow. Knowledge of slug flow characteristics is crucial to design pipelines and process equipments. In order to improve the accuracy of slug characteristics for high viscosity oils, new models for slug flow are needed, such as translational velocity.

Translational velocity is composed of a superposition of the velocity of the bubble velocity in stagnant liquid, i.e. the drift velocity v_d, and the maximum velocity in the slug body. Research efforts have been focused on drift velocity in horizontal pipes.

Nicklin et al. (1962) proposed an equation for translational velocity as,

$$v_t = C_s v_s + v_d \,. \tag{1}$$

The parameter C_s is approximately the ratio of the maximum to the mean velocity of a fully developed velocity profile. C_s equals approximately 1.2 for turbulent flow and 2.0 for laminar flow.

Wallis (1969), and Dukler and Hubbard (1975) claimed that there is no drift velocity for horizontal flow since gravity cannot act in the horizontal direction. However, Nicholson et al. (1978), Weber (1981) and Bendiksen (1984) showed that drift velocity exists for the horizontal flow and the value of drift value can exceed that for vertical flow. The drift velocity results from hydrostatic pressure difference between the top and bottom of the bubble nose.

For the drift velocity, Benjamin (1968) proposed the following relationship for horizontal pipes,

$$v_d = 0.542\sqrt{gD} \,. \tag{2}$$

Benjamin (1968) calculated the drift velocity coefficient by using inviscid (potential) flow theory (surface tension and viscosity are neglected.) The drift velocity in horizontal slug flow is the same as the velocity of the penetration of a bubble when liquid is drained out of a horizontal pipe. Bendiksen (1984) and Zukoski (1966) supported the study of Benjamin (1968), experimentally.

Bendiksen (1984) performed an experimental study for velocities of single elongated bubbles in flowing liquids at different inclination angles. He proposed the following equation for all inclination angles:

$$v_d = v_d^h \cos\beta + v_d^v \sin\beta \,. \tag{3}$$

Where, v_d^h and v_d^v are drift velocities for horizontal and vertical flow, respectively.

Zukoski (1966) experimentally investigated the effects of liquid viscosity, surface tension, and pipe inclination on the motion of single elongated bubbles in stagnant liquid for different pipe diameters. He also found that the effect of viscosity on the drift velocity is negligible for Re= $v_d \rho D/\mu$ >200.

Gokcal et al. (2008) proposed a mechanistic model to evaluate drift velocity for high viscosity oils in horizontal pipes. The model was based on Benjamin

(1968) in which the mass and energy balanced equations were solved, and matched the experimental data very well. The drift velocity model for horizontal flow was developed in terms of h/D instead of liquid viscosity. The liquid viscosity vs. liquid height from the conducted experiments was plotted. The viscosity correlation is developed as a function of h/D, based on experimental results. This relationship was based on a single pipe diameter and liquid viscosity up to 0.7 Pa·s. The relationship for the h/D resulting from the experiments was used as the link between viscosity and drift velocity. In the present study, a wider range of viscosities were tested using a CFD model with no need for h/D as an input parameter.

Mathematical Modeling and Numerical Solution

The drift velocity information can be obtained by simulating the draining of oil, and simultaneously the penetration of air (gas) phase into an initially oil filled horizontal circular pipe of diameter D. A schematic of the physical domain under investigation is shown in Fig.1.

The process is a time-dependent, two-phase fluid flow. Since the two phases are not inter-penetrating, the volume of fluid method (VOF) is the most suitable method to track the gas-liquid (air-oil) interface during this drainage process. For the range of viscosities studied in the present work, the flow in the oil phase is laminar and may be creeping at high viscosity (up to 1.2 *Pa.s*). Furthermore, the air phase flow is also laminar as the Reynolds number is expected to stay below 2000 for the cases investigated.

VOF Model

Only the major features and assumptions of the VOF model are discussed below. The reader is referred to the Fluent 6.1 Manual for details.

The VOF formulation assumes that the two phases are not inter-penetrating. In each control volume of the computational domain, the volume fraction of both phases add-up to unity. The flow field for all variables and properties are shared by both phases and represent volume-averaged values, provided that the volume fraction of each of the phases is known at each location. The volume fraction of the gas (air) is denoted by α_l. In the VOF model, given a cell inside the computational domain, three conditions are possible:

- α=0: the cell is full of liquid.
- α=1: the cell is full of gas.
- $0<\alpha<1$: the cell contains an interface between the gas and the liquid.

The local value of the volume fraction α_g determines the appropriate properties and variables appearing in the transport equations.

In the VOF formulation, a single momentum equation is solved throughout the computational domain, and therefore, the predicted velocity field is shared among the phases. In cases where large velocity differences exist between both phases near the interface, the VOF model cannot predict this difference, as all variables are locally shared across the field.

Results and Discussion

Two-Dimensional Results

This case is intended as a benchmark to compare the CFD results with Benjamin (1968) results. Figure 2 shows the CFD simulation for the case of water.

As can be seen from Fig. 2, the drained liquid height for the case of the lowest tested viscosity (water) is almost constant, and can be easily averaged to get a value that represents h/D. As the viscosity increases as shown in Figs. 3 and 4, the liquid height decreases with axial distance, and it is hard to be averaged into a single value. Therefore, in this study, the distance at which the height is calculated is emphasized.

Figure 5 shows a comparison of different models with Benjamin (1968) and the experimental results of Gokcal et al. (2008) in terms of the normalized liquid drainage height, h/D. The figure shows clearly that the drift velocity was over predicted by the Benjamin 2D model. The 3-D CFD results were the closest to the experimental data. Moreover, the differences between the data and both the 2-D and 3-D CFD predictions were not pronounced up to a normalized liquid height of 0.725.

Figure 6 shows the variation of normalized liquid height as a function of the liquid viscosity. It is clearly seen that as the viscosity increases the h/D increases and reaches an asymptotic value at very high viscosities. It can also be seen that h/D is not constant along the pipe; h/D at x= 0.3 is higher than h/d at x = 0.75.

Figure 7 shows the drift velocity variation with viscosity. The CFD results and the experimental data have the same trend. The difference between the CFD modeling and the experiments could be related to many reasons. One of the reasons is the exponential relationship between the viscosity and the drift velocity. A slight change in viscosity (especially at lower values) can cause larger change in the drift velocity. Slight changes in the temperature during experiments can cause a larger effect on the viscosity and at the end in the drift velocity. In general, the CFD modeling is in good agreement with the experiments and can be used as a tool to predict the drift velocity of heavy oil-gas flows.

Concluding Remarks

This study presents an investigation on the effect of liquid viscosity on drift velocity, or the penetration velocity of a gas phase into a horizontal pipe initially filled with liquid, using a CFD approach. CFD simulation results agree with Gokcal (2008) experimental results. Benjamin's (1968) analytical solution for two-dimensional flow with energy loss does not agree with the experimental observations. The model for the energy dissipation case is not applicable to viscous oil. Drift velocity decreases with an increase in liquid viscosity. The height of the liquid film behind the penetrating gas bubble (h/D) increases with an increase in liquid viscosity. h/D is not constant during drainage of the liquid.

Near Future Activities

The CFD study will continue with an investigation of pipe diameter effect on drift velocity.

Nomenclature

C	*Coefficient [-]*
D	*Diameter [m]*
g	*Gravity [m/s^2]*
h	*Liquid height [m]*
Re	*Reynolds number [-]*
t	*Time[s]*
v	*Velocity [m/s]*
x	*Axial coordinate*

Greek symbols

α	*Gas void fraction [-]*
β	*Pipe inclination angle from horizontal [degrees]*
μ	*Viscosity [Pa·s]*
ρ	*Density [kg/m^3]*

Superscripts

h	*Horizontal*
v	*Vertical*

Subscripts

d	*Drift*
s	*Mixture*
t	*Translational*

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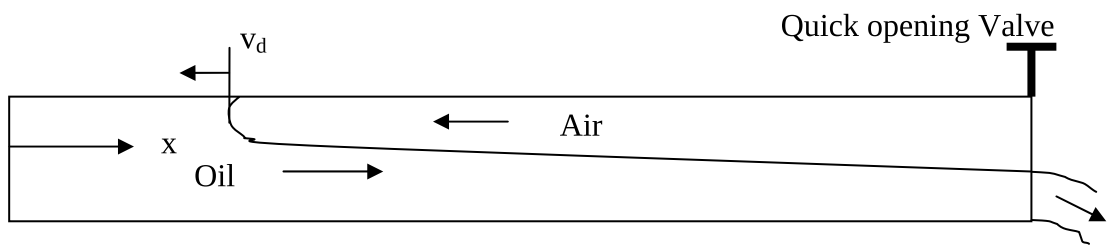

Figure 1: Schematic of gas penetration into a liquid filled pipe.

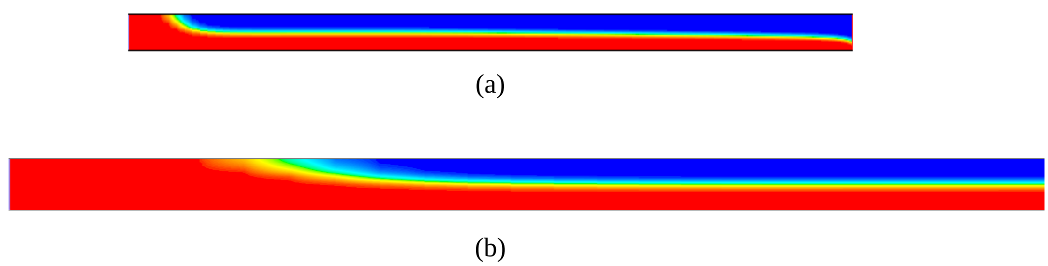

Figure 2: Contours of volume fraction (Red color: 100% water, Blue color: 100% air) from water drainage at time t=4.5sec, (a) 2D channel 1.5 m long, separated by a distance of 5 cm (quick opening valve located at the right end) (b) zoom-in view of the left end.

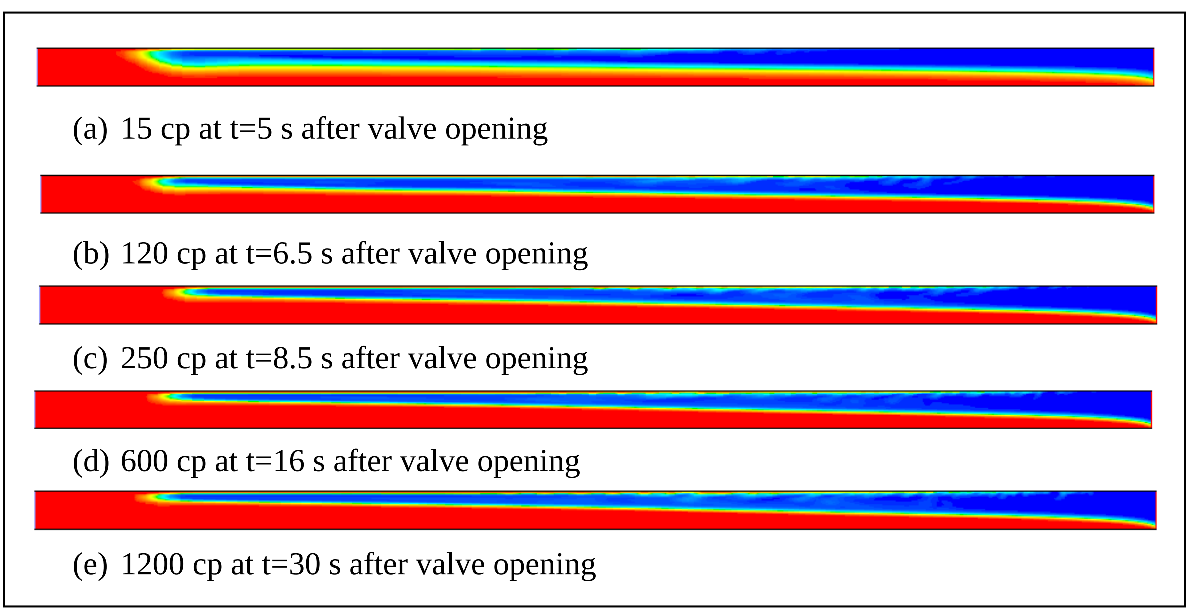

Figure 3: 2D Drift velocity simulation to approximately the same length for different oil viscosities.

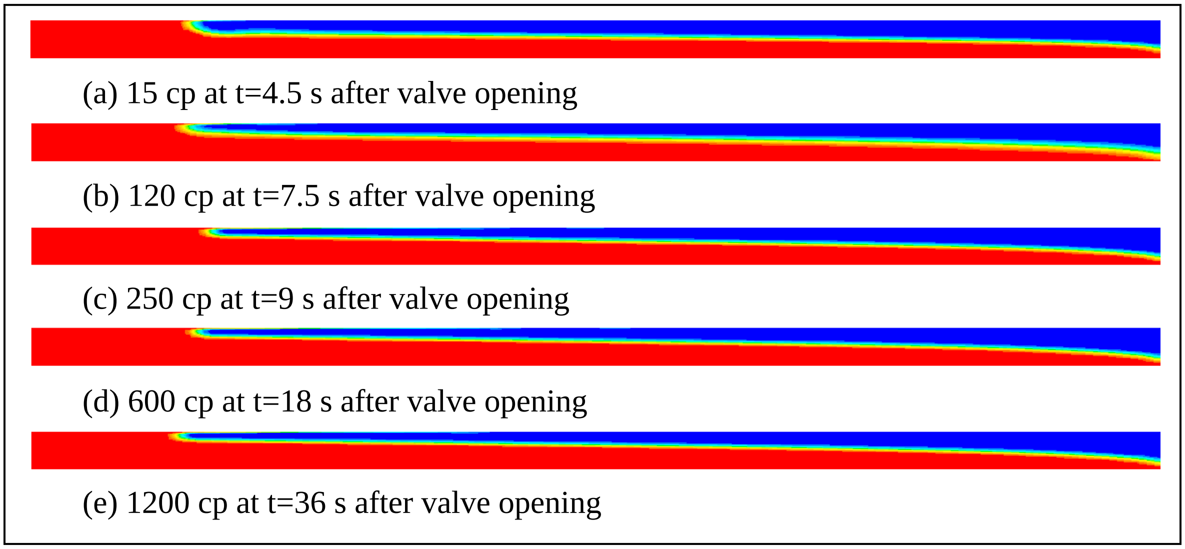

Figure 4: 3-D Drift velocity simulation to approximately the same length for different oil viscosities.

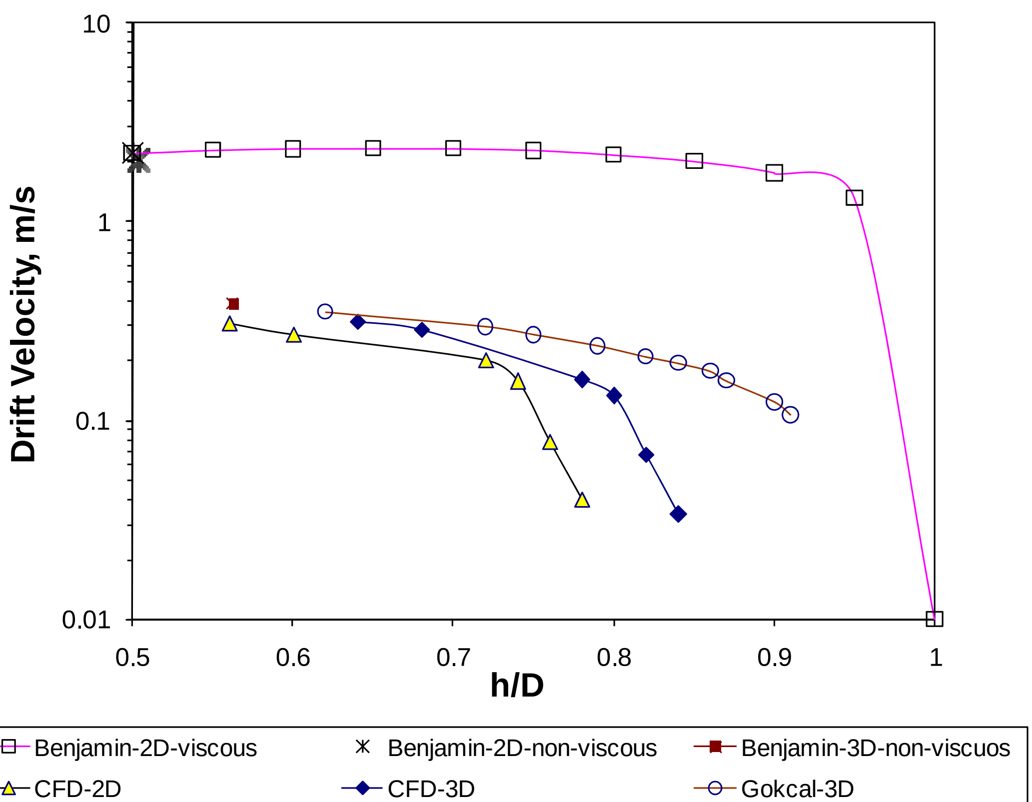

Figure 5: Comparison of different model predictions with experiment data of Gokcal et al. (2008) in terms of normalized liquid drainage height.

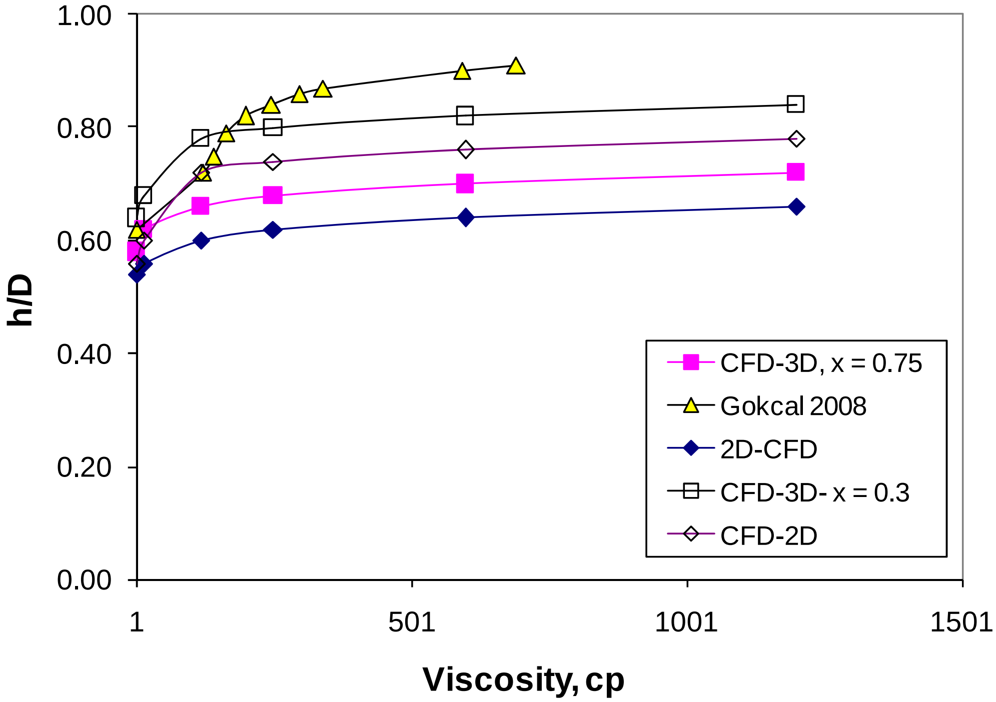

Figure 6: Variation of normalized liquid height with viscosity

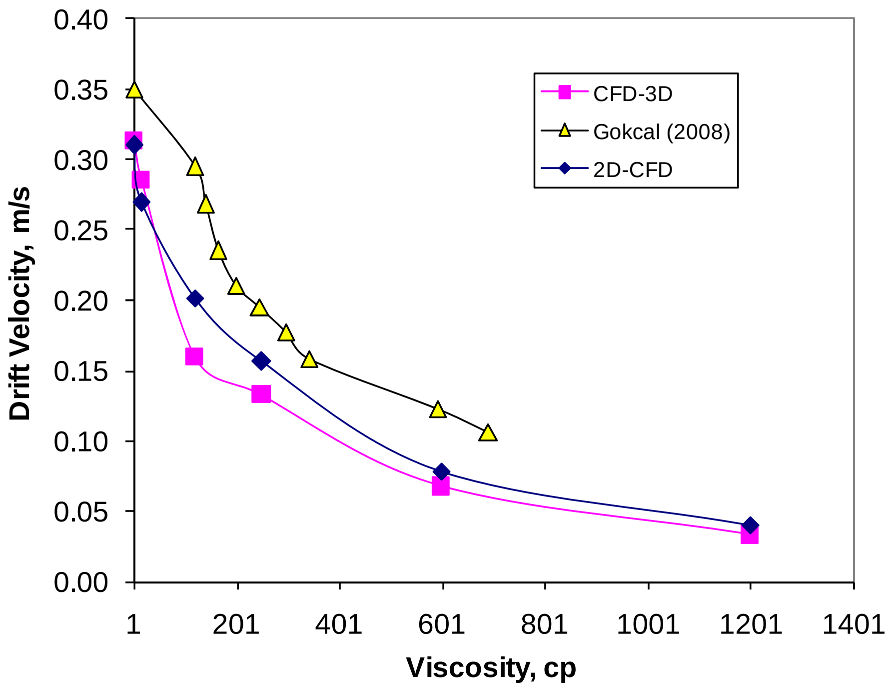

Figure 7: Variation of drift velocity with oil viscosity

Fluid Flow Projects

Investigation of Heavy Oil Two-Phase Slug Length in Horizontal Pipes

Eissa Alsafran (KU/KOC)

Advisory Board Meeting, March 25, 2009

Outline

- Introduction
- Experimental Analysis
- Theoretical Considerations
- Modeling Possibilities
- Concluding Remarks

Significance

- Pipeline Design (Sizing and Routing)
  - Pressure Drop
  - Liquid Volume
- Facility/Equipment Design
  - Instantaneous Liquid Rate at Pipe Outlet is 5-20X of Average Rate
  - Slug Catchers
  - Multiphase Pumps
  - Multiphase Meters

Significance

- Flow Assurance
  - Terrain Slugging
  - Erosion/Corrosion
- Mechanical Integrity
  - Piping System
  - System Components

Literature Review

- No Literature is Found on Heavy Oil Two-phase Slug Length
- Light Oil Slug Length is Strongly Correlated to Pipe Diameter, and Insensitive to Other Parameters
- Light Oil Slug Length
  - Smallest Near the "Center" of Slug Flow Region on Flow Pattern (FP) Map
  - L_s Increases Near Transition Boundaries

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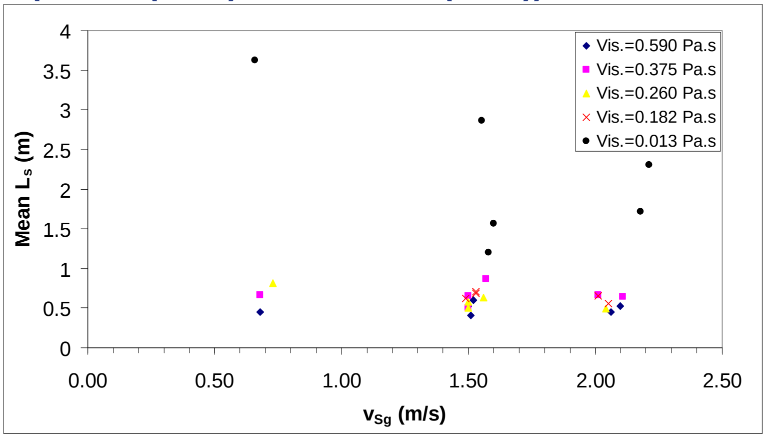
Experimental Analysis

- Comparison with Light Oil Slug Length (Gokcal(2008) and Alsafran(2003))



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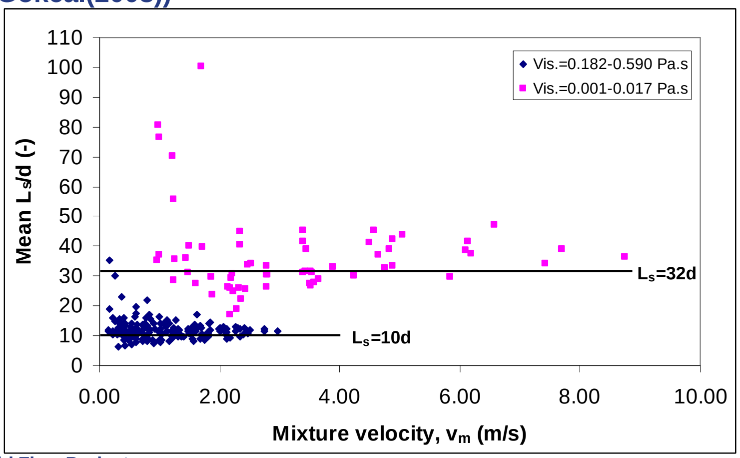
Experimental Analysis

- Comparison (Kouba(1986), BP Loop (2001), Alsafran(2003), Gokcal(2008))



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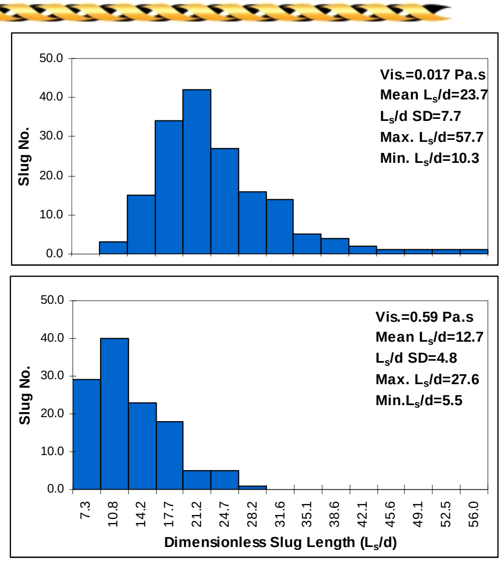
Experimental Analysis . . .

- Comparison (Light vs. Heavy)

v_{SL}=0.3 m/s

v_{Sg}=1.5 m/s



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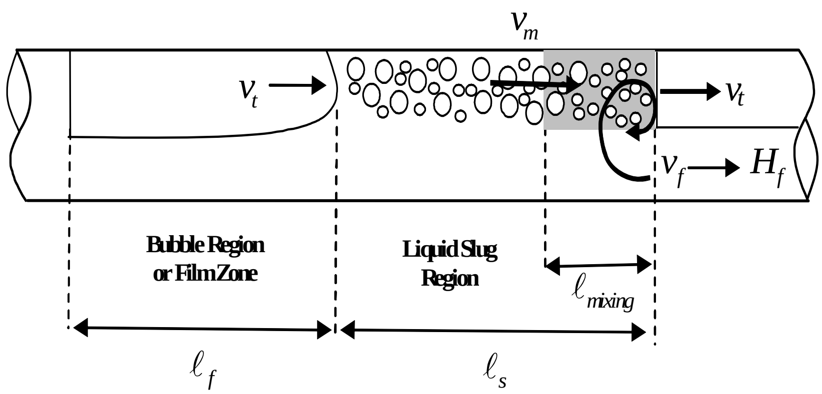
Theoretical Considerations



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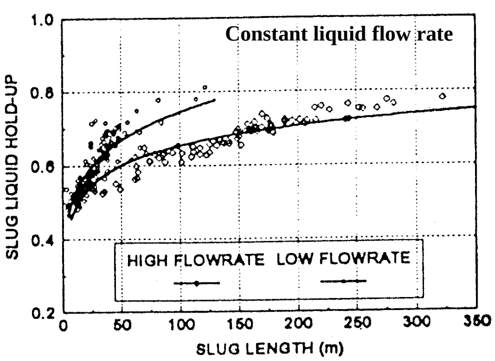
Theoretical Considerations (Light Oil)

- Relations to Slug Flow Characteristics
  - Mean L_s Increases as Slug Void Fraction Increases

- Bekapai (Indonesia) 42-km and 12-in. Pipeline (Dhulesia, BHRg 1993)

- This Trend is not Observed in Experimental Data



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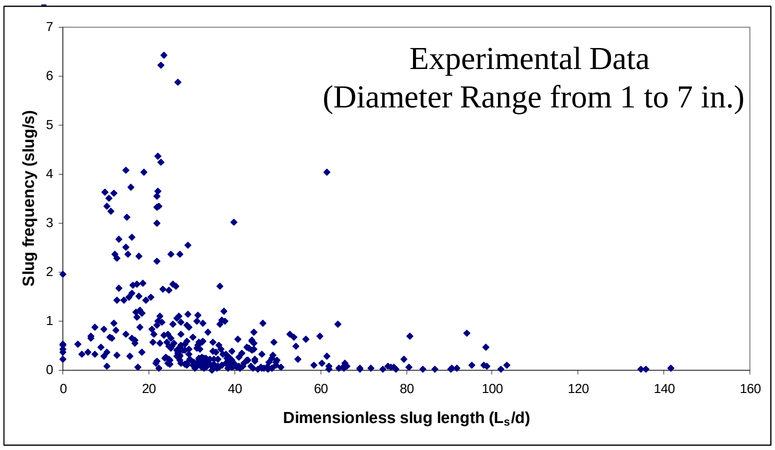
Theoretical Considerations (Light Oil) . . .

➢ Mean L_S Decreases as Slug Frequency



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Theoretical Considerations (Heavy Oil)

- Analogy
  - ➢ Viscosity Effect on Slug Length is Similar to Upward Inclination Angle Effect on Slug Length
- Viscous Forces Tend to Increase Liquid Holdup in Pipe, Increasing Liquid Film Height in Bubble Region and Slip Between Two Phases
- Thick Film Impacts Several Slug Flow Characteristics as Follows

Fluid Flow Projects

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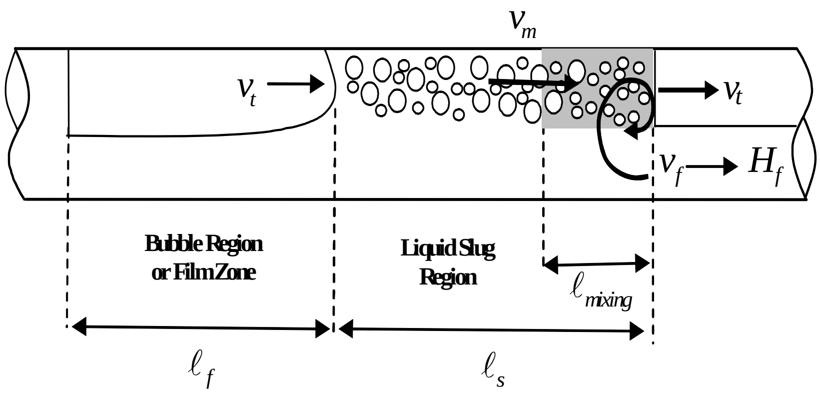
Theoretical Considerations (Heavy Oil) . . .

- Decreases Slug Liquid Holdup ➔ Reducing Slug Length



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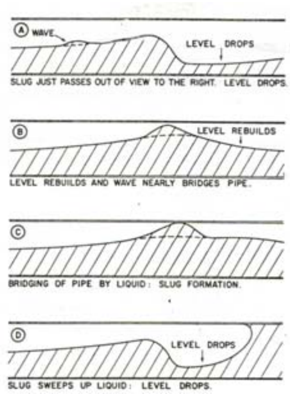
Theoretical Considerations (Heavy Oil) . . .

- Increases Slug Frequency ➔ Decreases Slug Length



Slug Initiation Process and Its Relation to Liquid Height

Fluid Flow Projects

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Theoretical Considerations (Heavy Oil) . . .

- Taylor Bubble (TB) Velocity and Interaction Between TBs
  - TBs Have Longer Wake in Heavy Oil Than Light Oil
  - Long Wake of Leading TB Promotes Acceleration the Trailing TB
  - This Lead to a Short Minimum Stable Slug Length

Modeling Possibilities

- Existing Correlations
  - Brill et al. (1981)

$$\ln\left(\overline{L_s}\right) = -2.663 + 5.441\sqrt{\ln(d)} + 0.059\ln(v_m)$$

$$L_{\max} = 4.66\overline{L_s}$$

  - Exxon (1982)

$$\ln\left(\overline{L_s}\right) = -2.099 + 4.859\sqrt{\ln(d)}$$

Modeling Possibilities . . .

- Scott et al. (1990)

$$\ln\left(\overline{L_s}\right) = -2.099 + 4.859\sqrt{\ln(d)}$$

- Alsafran et al. (2005): Full L_s Distribution

$$\mu_N^* = 5.32d^* - 17.91Q_{LF}{}^* + 1.58\Theta^* + 0.047v_m^*$$

$$\sigma_N = 0.297 - 1.027H_F + 0.995\Theta$$

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Modeling Possibilities . . .

- Alsafran et al. (2005) . . .

$$\mu_L = exp\left[\mu_N + \frac{\sigma_N^2}{2}\right]$$

$$\sigma_L^2 = \mu_L^2\left[exp\left(\sigma_N^2\right) - 1\right]$$

$$L_{s_{0.99}} = \exp\left(\mu_N - 2.58\sigma_N\right)$$

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Modeling Possibilities . . .

- Selection of Probabilistic Model
- Identification of Significant Correlation Parameters
  - Liquid Film Height
    - Already Been Used for Light Oils
    - Can Be Calculated From Mechanistic Models

Modeling Possibilities . . .

- Slip Factor
  - By Definition

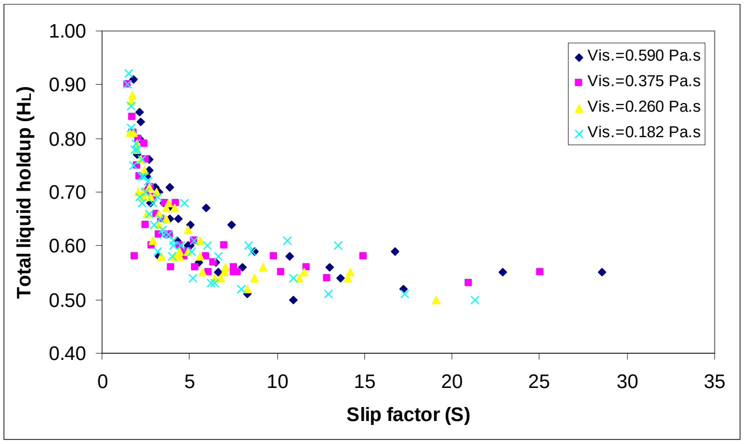
$$S = \frac{v_g}{v_L}$$

  - Correlated in Literature (Schmidt et al. (2008))

$$S = 1 + 1.95\left(\frac{\mu_L}{\mu_0}\right)^{0.258}\left(\frac{v_{Sg}}{v_{SL}}\right)^{0.749} \qquad \mu_0 = 1 \text{ Pa.s}$$

Modeling Possibilities . . .

- Slip Factor Fitted with Heavy Oil Data



Modeling Possibilities . . .

- Heavy Oil Average Slug Length is Constant

$$\overline{L_s} = 10D$$

- Modify Existing Correlations by Including Slip Factor or Liquid Height

Concluding Remarks . . .

- Significant Difference in Mean, Standard Deviation and Maximum Value of Slug Length Between Light and Heavy Oils
- High Liquid Viscous Forces Increases Liquid Holdup
- High Liquid Viscous Forces Change Slug Flow Characteristics
  - Slug Liquid Holdup
  - Slug Frequency
  - TB Velocity

Concluding Remarks . . .

- Change in These Slug Flow Characteristics Reduces Mean Slug Length
- Slug Length May Correlate With
  - Liquid Film Height
  - Slip Between Two Phases
- Existing Correlation Can Be Modified to Include Viscosity Effect

Concluding Remarks . . .

- Any Feedback is Appreciated and Crucial at This Stage of the Study
- Thanks for Your Attention

Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Low Liquid Loading Flow

- Significance
  - Wet Gas Transportation
    - Holdup and Pressure Drop Prediction
    - Corrosion Inhibitor Delivery (Top of the Line Corrosion)
- Objectives
  - Develop Better Predictive Tools

Low Liquid Loading Flow …

- Past TUFFP Studies
  - Two-phase, Small Diameter, Low Pressure
    - Air-Water and Air-Oil
    - 2-in. ID Pipe with ±2° Inclination Angles from Horizontal
  - Two-phase, Large Diameter, Low Pressure
    - Air-Water
    - 6-in. ID and ±2° Inclination Angles from Horizontal

Low Liquid Loading Flow …

- Past TUFFP Studies …
  - Three-phase, Large Diameter, Low Pressure
    - Air-Mineral Oil-Water
    - 6-in. ID, Horizontal Flow
    - Findings
      - Observed and Described Flow Patterns and Discovered a New Flow Pattern
      - Acquired Significant Amount of Data on Various Parameters, Including Entrainment Fraction
    - Remaining Tasks
      - Development of Improved Closure Relationships

Low Liquid Loading Flow …

- Current Study
  - Three-phase, Large Diameter, Low Pressure Inclined Flow
    - Air-Mineral Oil-Water
    - 6-in. ID and ±2° Inclination Angles from Horizontal
    - Objectives
      - Acquire Similar Data as in Horizontal Flow Study
      - Develop Improved Closure Relationships

Low Liquid Loading Flow …

- Status
  - On Hold Due to Insufficient Graduate Student Performance
  - Started in Spring 2009
    - New Ph.D. Student
    - Research Scholar, Professor Yuxing Li of China University of Petroleum
- Future Studies
  - Two and Three-phase, Large Diameter, High Pressure Horizontal and Inclined Flow
    - Requires New High Pressure Facility

Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Up-Scaling Studies

- Significance
  - Better Design and Operation
- Objective
  - Testing and Improvement of Existing Models for Large Diameter and Relatively High Pressures
- Past Studies
  - Low Pressure and 6-in. ID Low Liquid Loading (Fan and Dong)
  - High Pressure 2-in. ID (Manabe, 2002)

Fluid Flow Projects

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Up-Scaling Studies …

- Current Project
  - Construction of a New High Pressure, Large Diameter Facility
  - Extension of Low Liquid Loading Study to High Pressures is Envisioned as the First Study

Up-Scaling Studies …

- Status
  - Design is Complete
    - Operable with both Nitrogen and Natural Gas
    - Professional Outside Evaluation of the Design is Complete
    - P&ID Developed by EnSerca Engineering
  - City Permits for Construction Obtained
  - Equipment Purchases
    - Gas Compressor and Power Generator are Purchased
    - Tanks, Separators and Stainless Pipes are Being Ordered
  - Construction is Underway

Up-Scaling Studies …

- Near Future Activities
  - Completion of Concrete
  - Completion of Support Structures
  - Assembly of All of the Components
  - November 2009 is Targeted for Completion of Construction

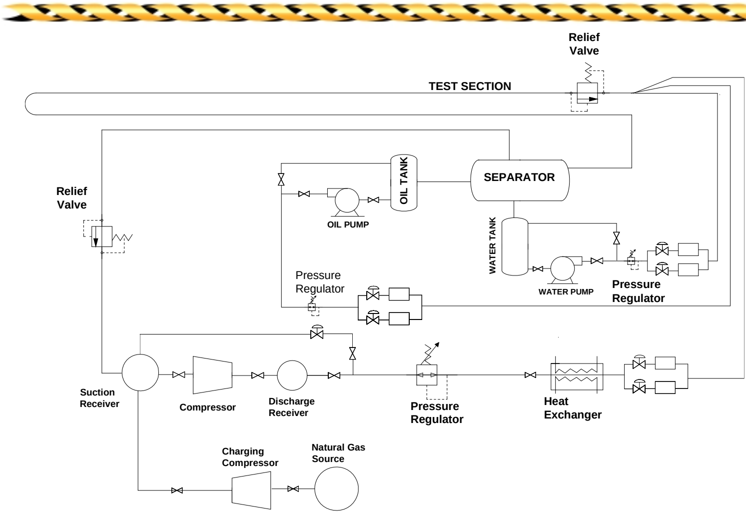
Fluid Flow Projects

6” High Pressure Flow Loop Construction Update

Scott Graham

Advisory Board Meeting, March 25, 2009

High Pressure Facility



Fluid Flow Projects

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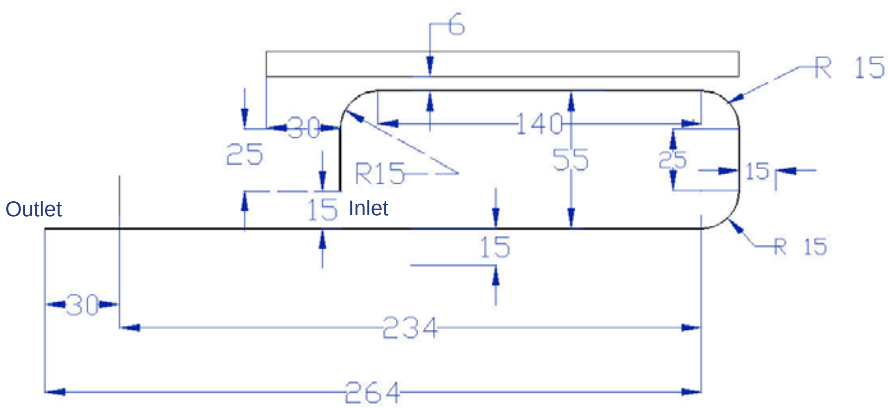
Flow Loop – Test Section (Dimensions in Feet)



HPF- Fluids

- Gas Phase
  - Nitrogen
  - Tulsa City Natural Gas
- Oil Phase - Tulco Tech-80 Mineral Oil
- Water Phase - Distilled Water

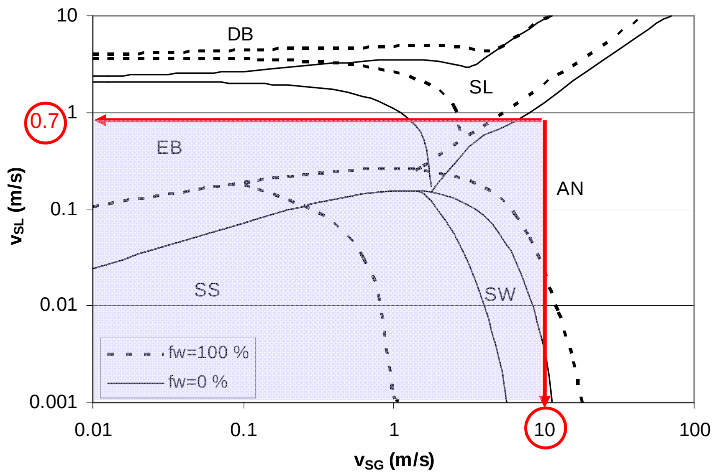
Flow Conditions

	Pressure (psig)	Capacity
Gas Flow Rate	600	18 MMSCF/D
Water Flow Rate	600	200 GPM
Oil Flow Rate	600	200 GPM
Differential Pressure	600	0 – 50 in H_2O
Pressure	600	0 – 800 psi
Temperature	600	Ambient

HPF - Operating Range

- Operating Pressure = 500 psig
- q_{gmax} = 18 MMSCF/D or $v_{Sg,max}$= 10 m/s
- q_{Lmax}= 200 GPM or $v_{SL,max}$= 0.7 m/s
- f_w between 0 and 100 %
- Separator 54 in. x 10 ft @ 600 psig

HPF - Flow Pattern Maps



Safety Considerations

- Three Steps
  - Hiring Professional Engineering Co. "Enserca Engineering"
  - Nitrogen as a Transition State Then Methane
  - Dept. of Transportation Safety Division Regulations

Safety Considerations

- Department of Transportation-Safety Division
  - While not required project was designed to meet DOT Pipeline Safety Regulations as they relate to a residential environment

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Pipe schedule

- 6" ANSI Schedule 40 304 Stainless Steel pipe
  - O.D. - 6.625 in.
  - I.D. – 6.070 in.
  - Thickness 0.28 in.
  - Weight Pound per foot -14.89 lbs
  - Pressure Rating – 1219 psi at 100°F

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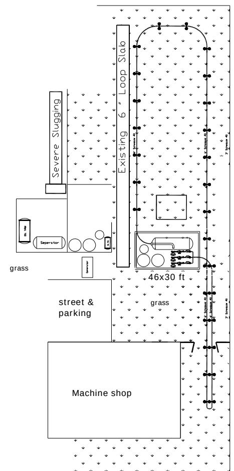
Facility Layout



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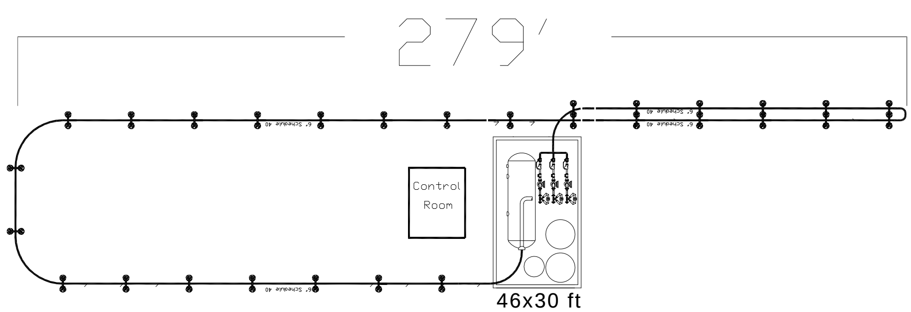
Facility Layout



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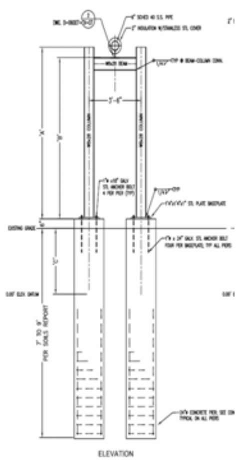
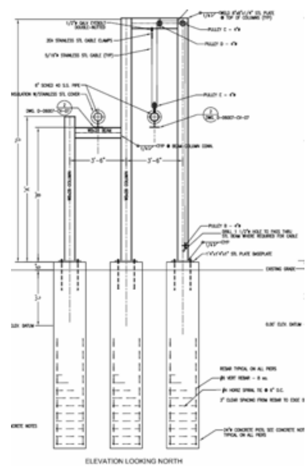
Support Detail





Fluid Flow Projects

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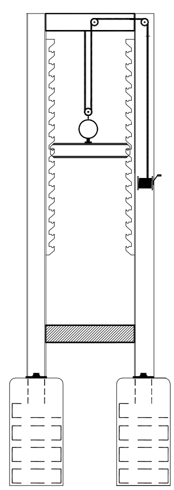
Lifting Mechanism



Fluid Flow Projects

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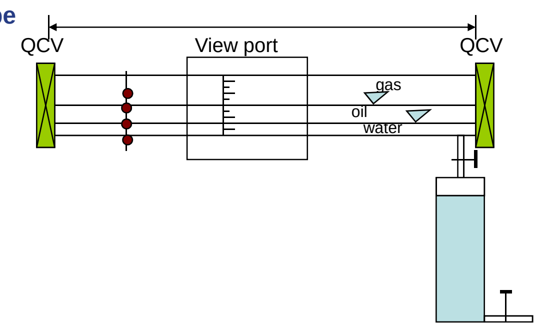
Instrumentation - Basic

	Pressure (psig)	Capacity (6 in. pipe)
Gas Flow Rate	600	18 MMSCFD
Water Flow Rate	600	200 GPM
Oil Flow Rate	600	200 GPM
Differential Pressure	500	0 – 50 in H_2O
Pressure	600	0 – 800 psi
Temperature	500	0-100 °C
Quick Closing Valves	600	6 in. ID

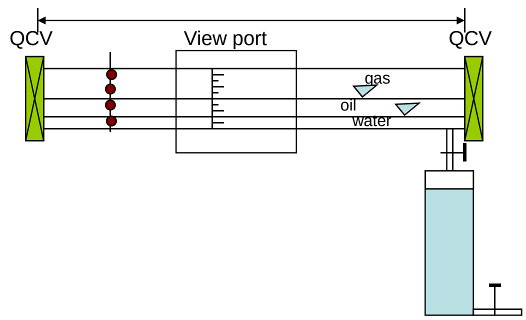
Instrumentation - Special

- **Total Liquid Holdup**
  - **Quick Closing Valve**
  - **Viewing Window (Liquid Height Measurement)**
  - **Two Lasers Sensor (Trial)**
  - **Conductivity Probe for Water Height (?)**
  - **Multipoint Densitometer**



Instrumentation- Special

- Oil/Water Holdup
  - Quick Closing Valve
  - High Pressure to Flush Liquid Out
  - Wait for Separation of Oil and Water
  - Multiple Point Densitometer to Get the Level
  - Push Liquid Back to Separator Using Gas Line



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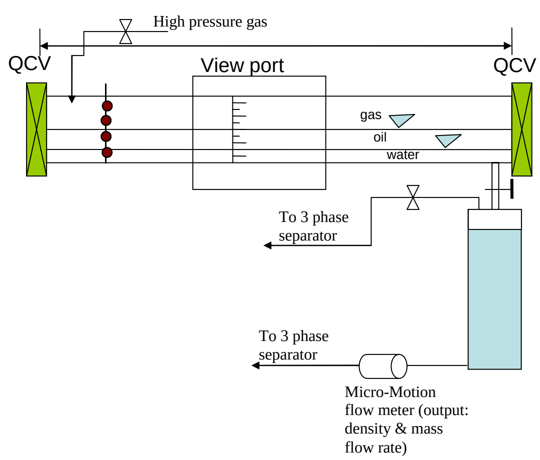
Instrumentation- Special (Suggested By Fan)

- Oil/Water Holdup



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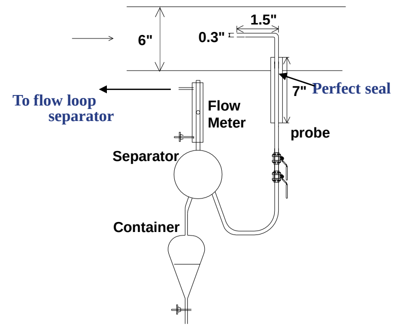
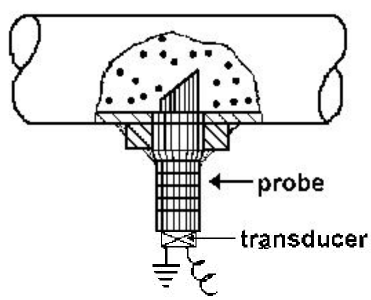
Instrumentation- Special

- **Liquid Entrainment**
  - Iso-kinetic Probe High Pressure Rating
  - Gas Outlet to Separator
  - Reaching Iso-kinetic Conditions is Expected to Be Challenging



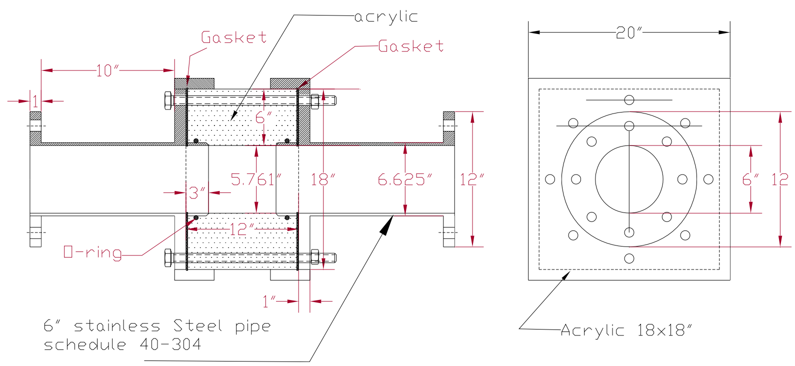
Instrumentation- Special

- **Liquid Entrainment**
  - Inline, Online Monitoring of Particle Flow
  - Detection of Minute Shock Waves by the Impact of a Particle or Droplet on a Probe
  - A Transducer Converts Waves, Which are Proportional to the Kinetic Energy, into Electrical Signals:
  - K.E. = $\frac{1}{2} m v^2$
  - Manufacturer Provides a Procedure to Accomplish Iso-kinetic Conditions



Instrumentation- Special

- Flow Pattern
  - Visual Observation/Whole Perimeter Viewing Section



  - Static Testing is Needed

Capital Cost Analysis

#	Component	Capacity	Cost	Actual Cost
1	Compressor	18 MMSCFD	242,000	243,980
2	Heat Exchanger	720,000 BTU/HR/Pass	20,000	Being quoted
3	Chiller	90 ton	67,000	Being quoted
4	Valves	2	20,000	Being quoted
5	Water pump	200 GPM	20,000	Being quoted
6	Oil pump	200 GPM	20,000	Being quoted
7	Separator	54" x 10' x 600	36,000	Being quoted
8	Water tank	1200 gallon	33,000	Being quoted
9	Oil tank	1200 gallon	33,000	Being quoted
10	Pipeline (SS)	6-in. ID, 540 ft	90,000	95,000

Capital Cost Analysis

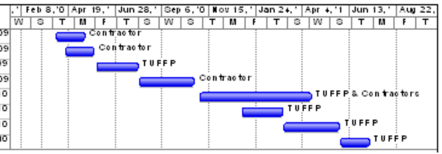
#	Component	Capacity	Cost (K $)	Actual Cost (K $)
11	Gas flow rate	18 MMSCFD	20,000	23,759
12	Water flow rate	200 GPM	20,000	8905
13	Oil flow rate	200 GPM	20,000	8905
14	Diff. pressure	0 – 50 in H2O (8)	8,000	Being quoted
15	Pressure	0 – 800 psi (8)	5,000	Being quoted
16	Temperature	0-100 C (8)	5,000	Being quoted
17	QCV	6 in ID (5)	15,000	Being quoted
18	Power generator	500 KW	65,000	64,965
19	Steel structure/Tilting		50,000	74,378
20	Pressure regulator	3 (Oil, Water & Gas)	5,000	Being quoted
21	Concrete foundation	600 ft by 6 ft	50,000	92,825
22	Comp. Surge control	Dual loop	25,000	Being quoted
23	Data Acquisition system		10,000	Being quoted

Capital Cost Analysis

#	Component	Capacity	Cost	Cost
25	Geo-Tech-Exploration	3 points	2400	2400
26	Enserca Engineering			
	P&ID and PFD		27,400	27,400
	Permit Review		6400	6400
	Civil/Structural Design		31,500	31,500
	Total		**957,000**	**???????**

Time Table

ID	Task Name	Duration	Start	Finish
1	Concrete Work	30 days	Wed 4/1/09	Tue 5/12/09
2	Fabrication of Structural Steel	30 days	Wed 4/15/09	Tue 5/26/09
3	Installation of Structural Steel	45 days	Mon 6/1/09	Fri 7/31/09
4	Construction of 6" S.S. Flow loop	60 days	Mon 8/3/09	Fri 10/23/09
5	Installation of process vessels and equipment	120 days	Mon 11/2/09	Fri 4/16/10
6	High Voltage electrical work	45 days	Mon 1/4/10	Fri 3/5/10
7	Instrumentation and Controls	60 days	Mon 3/8/10	Fri 5/28/10
8	Startup and Shake Down of facility.	30 days	Tue 6/1/10	Mon 7/12/10



6” Flow Loop

Questions?

Transient Modeling

- Significance
  - Industry has Capable All Purpose Transient Software
    - OLGA, PLAC, TACITE
  - Efforts are Well Underway to Develop Next Generation All Purpose Transient Simulators
    - Horizon, LEDA
  - Need for a Simple Transient Flow Simulator

Transient Modeling …

- Objective
  - Development and Testing of a Simple and Fast Transient Flow Simulator That Can Be Used as a Screening Tool
- Project Proposal Rated High in Recent TUFFP Questionnaire
- Past Studies
  - TUFFP has Conducted Many Transient Multiphase Studies
    - Scoggins, Sharma, Dutta-Roy, Taitel, Vierkandt, Sarica, Vigneron, Minami, Gokdemir, Zhang, Tengesdal, and Beltran

Transient Modeling …

- **Status**
  - Researcher is Identified
  - Will Start Immediately After Advisory Board meeting

Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

Unified Model

- **Objective**
  - Develop and Maintain an Accurate and Reliable Steady State Multiphase Simulator
- **Past Studies**
  - Zhang et al. Developed “Unified Model” in 2002 for Two-phase Flow
    - Became TUFFP’s Flagship Steady State Simulator
    - Applicable for All Inclination Angles
  - “Unified Model was Extended to Three-phase in 2006

Fluid Flow Projects

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Unified Model …

- Current Activities
  - Code and Software Improvement Efforts

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Unified Model …

- Future Activities
  - Continue Improvements in Both Modeling and Software Development

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Fluid Flow Projects

Business Report

Cem Sarica

Advisory Board Meeting, March 25, 2009

Membership Status

- Current Status
  - Membership Stands at 16
    - 15 Industrial and MMS
  - Tenaris and Petronas Terminated Their Memberships
  - SPT Joined
  - Efforts Continue to Increase Membership
    - CNOOC Expected to Join

Personnel Changes

- Dr. Polat Abduvayt Joins TUFFP as Post Doctoral Research Associate
- Mr. Kiran Gawas Joins TUFFP Team as Research Assistant to Purse Ph.D. Degree in Petroleum Engineering
- Drs. Eissa Alsafran and Yuxing Li as Visiting Research Scholars

Papers and Publications

- Atmaca, S., Alsarkhi, A., Zhang, H. Q. and Sarica, C.: "Characterization of Oil Water Flows in Inclined Pipes," SPE115485, Accepted for Publication in *SPE Projects, Facilities & Construction Journal,* 2009.
- Gokcal, B., Alsarkhi, A. and Sarica, C.: "Effects of High Viscosity on Drift Velocity for Inclined Pipes," *SPE Projects, Facilities & Construction Journal,* 2009
- Al-Safran, E. Kappos, L. and Sarica, C. "Experimental and Numerical Investigation of Separator Pressure Fluctuation Effect on Terrain Slugging in a Hilly-Terrain Two-phase Flow Pipeline," *Journal of Energy Resources Technology* September 2008.
- Vielma, M., Atmaca, S., Zhang, H. Q., and Sarica, C.: "Characterization of Oil/Water Flows in Horizontal Pipes," SPE109591, *SPE Projects, Facilities & Construction Journal,* December 2008.

Next Advisory Board Meetings

- Tentative Schedule
  - September 29, 2009
    - TUHOP Meeting
    - TUFFP Workshop
    - Facility Tour
    - TUHOP/TUFFP Social Function
  - September 30, 2009
    - TUFFP Meeting
    - TUFFP/TUPDP Reception
  - October 1, 2009
    - TUPDP Meeting
- Venue is The University of Tulsa

Financial Report

- Year 2008 Closing
  - TUFFP Industrial Account
  - TUFFP MMS Account
- Year 2009
  - TUFFP Industrial Account
  - TUFFP MMS Account

2008 TUFFP Industrial Account Budget Summary
(Prepared February 25, 2009)

Anticipated Reserve Fund Balance on January 1, 2008	**575,234.31**
Income for 2008	
2008 Membership Fees (16 @ $48,000 - excludes MMS)	**$768,000**
2008 Membership Fees (1 @ 38,000)	**$38,000**
Total Budget	**1,381,234.31**

Projected Budget/Expenditures for 2008

		Budget	Revised Budget 4/08	Expenses 2/11/09	Anticipated 2008 Expenses
90101	Principal Investigator - Sarica	24,392.00	25,907.00	21,494.04	27,930.48
90103	Co-Principal Investigator - Zhang	19,665.00		3,622.50	3,622.50
90600	Professional Salary - Jones	5,141.00	7,330.00	8,472.20	8,472.20
90601	Professional Salary - Li	13,505.00	28,854.00	22,567.74	24,651.07
90602	Professional Salary - Graham	5,237.00	15,785.00	15,505.93	15,505.93
90603	Professional Salary - Al-Sarkhi	32,500.00	38,750.00	16,646.15	6,458.34
90701	Technician - Miller	15,065.00		1,853.73	
90702	Technician - Waldron	6,575.00	11,428.00	11,218.27	9,869.21
90703	Technician - Kelsey	9,750.00	10,154.00	14,637.00	12,754.00
90800	Salaries - Part-time	4,290.00		518.50	
91000	Graduate Students - Monthly	50,100.00	65,000.00	60,699.99	60,745.44
91100	Students - Hourly	15,000.00	15,000.00	14,797.88	14,064.63
91800	Fringe Benefits (33%)	50,910.83	45,609.00	38,285.79	36,057.04
93100	General Supplies	3,000.00	3,000.00	283.57	500.00
93101	Research Supplies	100,000.00	100,000.00	87,678.98	100,000.00
93102	Copier/Printer Supplies	500.00	500.00	169.05	280.00
93104	Computer Software	4,000.00	4,000.00	502.18	1,200.00
93106	Office Supplies	2,000.00	2,000.00	2,070.79	3,000.00
93200	Postage/Shipping	500.00	500.00	376.77	650.00
93300	Printing/Duplicating	2,000.00	2,000.00	3,372.82	2,000.00
93400	Telecommunications	3,000.00	3,000.00	518.16	1,766.00
93500	Membership/Subscriptions	1,000.00	1,000.00	226.50	400.00
93600	Travel			-	
93601	Travel - Domestic	14,000.00	10,000.00	4,899.30	10,000.00
93602	Travel - Foreign	10,000.00	10,000.00	7,807.80	10,946.46
93606	Visa			241.35	241.35
93700	Entertainment (Advisory Board Meetings)	10,000.00	10,000.00	18,359.05	10,000.00
94803	Consultants		16,000.00	18,500.04	18,500.00
94813	Outside Services	20,000.00	20,000.00	81,989.03	68,278.29
95103	Equipment Rental			5,483.72	
95200	F&A (55.6%)	119,456.33	121,324.00	106,956.20	102,344.00
98901	Employee Recruiting	3,000.00	3,000.00	285.25	147.00
99001	Equipment	600,000.00	200,000.00	262,316.76	263,500.00
99002	Computers	8,000.00	8,000.00	768.70	4,000.00
99300	Bank Charges	40.00	40.00	-	-
81801	Tuition/Fees	30,306.00	53,103.00	45,893.00	55,342.00
81806	Graduate Fellowship			1,811.75	1,809.22
	Total Expenditures	**1,182,933.16**	**831,284.00**	**880,830.49**	**875,035.16**

Anticipated Reserve Fund Balance as of 12/31/08 **500,403.82**

2008 MMS Account Summary

2008 TUFFP MMS Budget Summary
(Prepared February 11, 2009)

Reserve Balance as of 12/31/07	**$5,322**
2008 Budget	**40,000**
Total Budget	**45,322**

Projected Budget/Expenditures for 2008

		Budget	2008 Expenditures
91000	Students - Monthly	25,600.00	25,600.00
95200	F&A	14,233.60	13,952.00
81801	Tuition/Fees		
Total Anticipated Expenditures as of 12/31/08		**39,833.60**	**39,552.00**

Total Anticipated Reserve Fund Balance as of 12/31/08 **5,769.94**

2009 TUFFP Industrial Account Budget Summary
(Prepared February 25, 2009)

Anticipated Reserve Fund Balance on January 1, 2009				**500,403.82**
Income for 2009				
	2009 Membership Fees (15 @ $48,000 - excludes MMS)			**$720,000**
Total Budget				**1,220,403.82**
Projected Budget/Expenditures for 2009				
		Budget 09/08/08	**Revised Budget 2/11/09**	
90101	Principal Investigator - Sarica	29,251.82	29,074.14	
90103	Co-Principal Investigator - Zhang			
90600	Professional Salary - Jones	11,596.24	11,728.00	
90601	Professional Salary - Li	13,157.00	12,875.00	
90602	Professional Salary - Graham	19,923.00	20,149.00	
90603	Professional Salary - Polat	62,000.00	65,000.00	
90701	Technician - Miller	9,943.45	10,016.00	
90702	Technician - Waldron	16,826.00	16,517.00	
90703	Technician - Kelsey	19,097.00	18,746.00	
90800	Salaries - Part-time	-		
91000	Graduate Students - Monthly	58,100.00	62,850.00	
91100	Students - Hourly	15,000.00	15,000.00	
91800	Fringe Benefits (33%)	59,992.19	62,850.00	
93100	General Supplies	3,000.00	3,000.00	
93101	Research Supplies	100,000.00	100,000.00	
93102	Copier/Printer Supplies	500.00	500.00	
93104	Computer Software	4,000.00	4,000.00	
93106	Office Supplies	2,000.00	2,000.00	
93200	Postage/Shipping	500.00	500.00	
93300	Printing/Duplicating	2,000.00	2,000.00	
93400	Telecommunications	3,000.00	3,000.00	
93500	Membership/Subscriptions	1,000.00	1,000.00	
93600	Travel			
93601	Travel - Domestic	10,000.00	10,000.00	
93602	Travel - Foreign	10,000.00	10,000.00	
93606	Visa			
93700	Entertainment (Advisory Board Meetings)	10,000.00	10,000.00	
94803	Consultants	16,000.00	18,500.00	
94813	Outside Services	20,000.00	20,000.00	
95200	F&A (55.6%)	141,721.35	141,867.93	
98901	Employee Recruiting	3,000.00	3,000.00	
99001	Equipment	600,000.00	400,000.00	
99002	Computers	8,000.00	8,000.00	
99300	Bank Charges	40.00	40.00	
81801	Tuition/Fees	30,665.00	30,067.00	
81806	Graduate Fellowship			
	Total Expenditures	**1,280,313.05**	**1,092,280.07**	
Anticipated Reserve Fund Balance as of 12/31/09				**128,123.75**

2009 MMS Account Summary

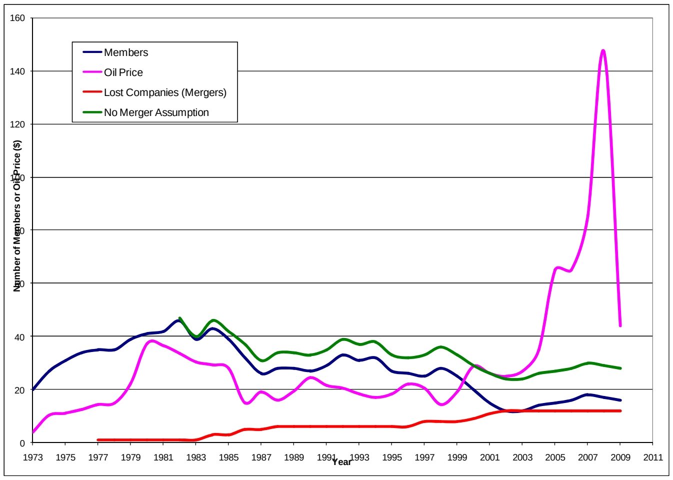
2009 TUFFP MMS Budget Summary
(Prepared February 11, 2009)

Reserve Balance as of 12/31/08				**5,769.94**
2009 Budget				**48,000.00**
Total Budget				**53,769.94**
Projected Budget/Expenditures for 2009				
		Budget	**2009 Anticipated Expenditures**	
91000	Students - Monthly	27,900.00	23,200.00	
95200	F&A	15,512.40	12,899.20	
81801	Tuition/Fees			
Total Anticipated Expenditures as of 12/31/09		**43,412.40**	**36,099.20**	
Total Anticipated Reserve Fund Balance as of 12/31/09				**17,670.74**

Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

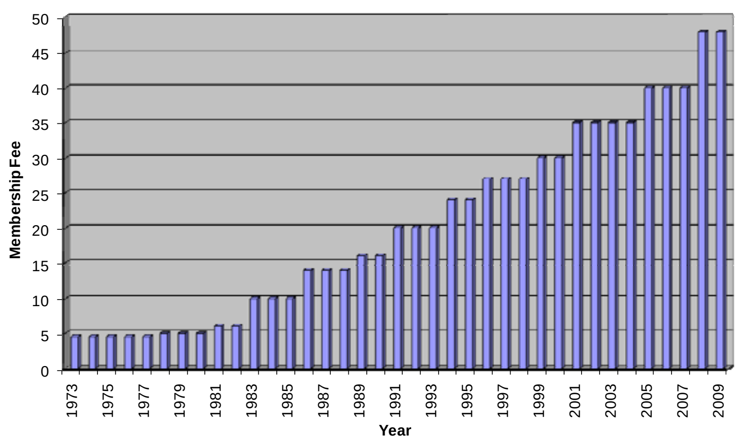
History – Membership



Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

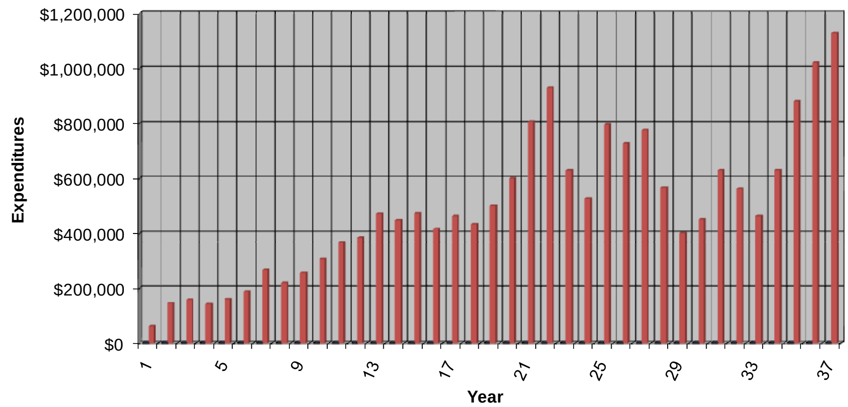
History – Membership Fees



Fluid Flow Projects

Advisory Board Meeting, March 25, 2009

History - Expenditures



Fluid Flow Projects

Membership Fees

- 2008 Membership Dues
  - 1 Unpaid
- 2009 Membership Dues
  - 10 of 16 Paid as of March 9
  - Need Your Prompt Payments

Fluid Flow Projects

Introduction

This semi-annual report is submitted to Tulsa University Fluid Flow Projects (TUFFP) members to summarize activities since the September 17, 2008 Advisory Board meeting and to assist in planning for the next six months. It also serves as a basis for reporting progress and generating discussion at the 72nd semi-annual Advisory Board meeting to be held in the Gallery Room of Allen Chapman Activity Center (ACAC) at the University of Tulsa Main Campus, 440 South Gary, Tulsa, Oklahoma on Wednesday, March 25, 2009.

Activities will start with the Tulsa University High-Viscosity Projects (TUHOP) Advisory Board Meeting on March 24, 2009 between 8:00 a.m. and noon in the Gallery Room in ACAC. Between 1:00 and 3:00 p.m. on March 24, 2009, there will be a TUFFP workshop in the same room. There will be presentations made by TUFFP member companies. A facility tour will be held on March 24, 2009 between 3:30 and 5:30 p.m. on the North Campus. Following the tour, there will be a TUHOP/TUFFP reception between 6:00 and 9:00 p.m. in McFarlin Library, Faculty Study at the University of Tulsa Main Campus.

The TUFFP Advisory Board meeting will convene at 8:00 a.m. on March 25th and will adjourn at approximately 4:30 p.m. Following the meeting, there will be a joint TUFFP and TUPDP reception between 5:30 and 9:00 p.m. in McFarlin Library, Faculty Study at the University of Tulsa Main Campus.

The Tulsa University Paraffin Deposition Projects (TUPDP) Advisory Board meeting will be held on March 26th in Gallery Room of Allen Chapman Activity Center (ACAC) at the University of Tulsa Main Campus, between 8:30 a.m. and 1:00 p.m.

The receptions on March 24 and 25 will provide an opportunity for informal discussions among members, guests, and TU staff and students.

Several TUFFP facilities will be operating during the tour. An opportunity will also be available to view the paraffin deposition test facilities and the hydrate flow loop.

The following dates have tentatively been established for Fall 2009 Advisory Board meetings. The venue for Fall 2009 Advisory Board Meetings is tentatively set to be held at the University of Tulsa Main Campus.

2009 Fall Meetings

September 29, 2009	Tulsa University High Viscosity Oil Projects (TUHOP) JIP Meeting Tulsa University Hydrate Flow Performance JIP (TUHFP) Tulsa University Fluid Flow Projects (TUFFP) Workshop Facility Tour TUHOP/TUHFP/TUFFP Reception
September 30, 2009	Tulsa University Fluid Flow Projects (TUFFP) Advisory Board Meeting TUFFP/TUPDP Reception
October 1, 2009	Tulsa University Paraffin Deposition Projects (TUPDP) Advisory Board Meeting

Personnel

Dr. Cem Sarica, Professor of Petroleum Engineering, continues as Director of TUFFP and TUPDP, and as Co-Principal Investigator of TUHFP and TUHOP.

Dr. Holden Zhang, Assistant Professor of Petroleum Engineering, serves as Principal Investigator of TUHOP and Associate Director of TUFFP.

Dr. Brill serves as a Research Professor of Petroleum Engineering on a consultancy basis.

We are very pleased to introduce Dr. Polat Abduvayt, a Chinese National, as our newest TUFFP Post Doctoral Research Associate. He has a BS degree in Petroleum Geology from the University of Petroleum in China, and MS and Ph.D. degrees in Petroleum Engineering from Waseda University in Japan. Before joining TUFFP, Dr. Abduvayt was a Senior Petroleum Engineer with Japan Oil Engineering Company and conducted extensive studies on fluid flow characteristics in wellbores and pipelines and in development of pressure drop and heat transfer prediction models as part of his duties. Before joining Japan Oil Engineering Company, he was a research associate in Waseda University in Japan. During his tenure with Waseda University, he worked as a Research Engineer conducting several experimental and theoretical multiphase flow studies for JOGMEC (formerly JNOC)

Dr. Mingxiu (Michelle) Li continues to serve as a Research Associate for TUHOP, TUFFP, and related projects.

Mr. Scott Graham continues to serve as Project Engineer. Scott oversees all of the facility operations and continues to be the senior electronics technician for TUFFP, TUPDP, and TUHOP.

Mr. Craig Waldron continues as Research Technician, addressing our needs in mechanical areas. He also serves as a flow loop operator for TUPDP and Health, Safety, and Environment (HSE) officer for TUFFP, TUPDP and TUHOP.

Mr. Brandon Kelsey serves as an electro-mechanical technician serving TUFFP, TUPDP and TUHOP projects. Brandon is a graduate of OSU Okmulgee with a BS degree in instrumentation and automation.

Ms. Linda Jones continues as Project Coordinator of TUFFP, TUPDP and TUHOP projects. She keeps the project accounts in addition to other responsibilities such as external communications, providing computer support for graduate students, publishing and distributing all research reports and deliverables, managing the computer network and web sites, and supervision of part-time office help.

Mr. James Miller, Computer Manager, and TUFFP TUPDP and TUHOP Web Administrator, resumed his duties in November 2008 after over a year of military leave.

Table 1 updates the current status of all graduate students conducting research on TUFFP projects for the last six months.

Dr. Bahadir Gokcal successfully completed his Ph.D. degree requirements during the fall 2008 semester. Dr. Gokcal is currently working for Technip in their Houston office.

Mrs. Gizem Ersoy Gokcal, from Turkey, started her Ph.D. degree studies. She is studying slug flow evolution in three-phase, gas-oil-water flow in hilly-terrain pipelines. Gizem received a BS degree in Petroleum and Natural Gas Engineering from Middle East Technical University and an MS degree in Petroleum Engineering from The University of Tulsa.

Mr. Kyle Magrini, a US National, received a BS degree in Electrical Engineering from The University of Tulsa. Kyle is studying “Liquid Entrainment in Annular Two-phase in Inclined Pipes” as part of his MS degree requirements.

Mr. Anoop Sharma, from India, has a BS degree in Chemical Engineering from the National Institute of Technology, Karnataka, India. He has also been involved in research at other universities such as Indian Institute of Science, Bangalore, India. He is studying to improve the two-phase oil-water flow modeling and closure relationship development as part of his MS degree requirements.

Ms. Tingting Yu graduated in 2007 from China University of Petroleum (East China), majoring in Oil and Gas Storage and Transportation. Tingting is a Teaching Assistant for the Petroleum Engineering Department. She is studying multiphase flow in an annulus as part of her MS degree requirements. Ms. Yu has accepted a position with SPT in their Houston Office, following completion of her MS studies.

Ms. Ceyda Kora, from Turkey, is pursuing her MS degree in Petroleum Engineering. Ceyda has received a BS degree in Petroleum and Natural Gas Engineering from Middle East Technical University in 2008. She is studying the effects of high viscosity oil on slug liquid holdup.

Mr. Kiran Gawas has recently joined the TUFFP team as a Research Assistant to pursue his Ph.D. degree in Petroleum Engineering. Mr. Gawas has a BS degree in Chemical Engineering from University of Mumbai, Institute of Chemical Technology and a Master of Technology degree from Indian Institute of Technology (IITB). Kiran is assigned to the Low Liquid Loading Three-phase Flow project.

A list of all telephone numbers and e-mail addresses for TUFFP personnel is given in Appendix D.

Table 1

2009 Spring Research Assistant Status

Name	*Origin*	*Stipend*	*Tuition*	*Degree Pursued*	*TUFFP Project*	*Completion Date*
Gizem Ersoy Gokcal	Turkey	Yes – TUFFP	Yes – TUFFP	Ph.D. – PE	Slug Flow Evolution in Three-Phase Gas-Oil-Water Flow in Hilly Terrain Pipelines	Summer 2009
Kiran Gawas	India	Yes – TUFFP	Waived (TU)	Ph.D. – PE	High Viscosity Oil Multiphase Flow Behavior	Fall 2012
Ceyda Kora	Turkey	Yes – TUFFP	Waived (MMS)	MS. – PE	Effects of High Viscosity Oil on Slug Liquid Holdup	Fall 2010
Kyle Magrini	USA	Yes – TUFFP	Yes – TUFFP	MS – PE	Entrainment Fraction in Annular Two-phase Flow in Inclined Pipes	Summer 2009
Anoop Sharma	India	Yes – TUFFP	Yes – TUFFP	MS – PE	Modeling of Hydrodynamics and Dispersions in Oil-water Pipe Flow	Summer 2009
Tingting Yu	PRC	Partial – TUFFP	No – PE Depart.	MS – PE	Multiphase Flow in a Vertical Annulus	Spring 2009

Membership

The current membership of TUFFP stands at 15 industrial members and Department of Interior's Mineral Management Services (MMS).

Our efforts continue to increase the TUFFP membership level. SPT joined TUFFP this year. A new five-year membership contract with MMS has been finalized. Drs. Cem Sarica and Holden Zhang visited PetroChina and Chinese National Offshore Oil Company (CNOOC) in June to establish relationships and solicit membership in our research programs. They have shown significant interest. It is expected CNOOC will soon join TUFFP.

Petronas informed us in November of 2008 that they wanted to terminate their TUFFP membership for 2008. We are currently trying to collect past due TUFFP membership fees from Petronas.

Table 2 lists all the current 2009 TUFFP members. A list of all Advisory Board representatives for these members with pertinent contact information appears in Appendix B. A detailed history of TUFFP membership is given in Appendix C.

Table 2

2009 Fluid Flow Projects Membership

Baker Atlas	Minerals Management Service
BP Exploration	PEMEX
Chevron	Petrobras
ConocoPhillips	Rosneft
Exxon Mobil	Schlumberger
JOGMEG	Shell Global Solutions
KOC	SPT
Marathon Oil Company	Total

Equipment and Facilities Status

Test Facilities

The design of a high pressure (500 psi) large diameter (6 in. ID) facility was completed and presented at the last Advisory Board meeting. The facility P&ID was prepared by an independent engineering company. The final stage before construction will be a HAZOP exercise with the help of one of the member company's HAZOP engineers. In Mid-December, we applied for all the applicable building permits that we need from the City of Tulsa and are awaiting approval at this time. We are currently in the process of securing bids for both the concrete work and the structural steel fabrication. Our goal is to have all the concrete work done by the end of March and begin assembling the structural steel in April. The flow-loop itself will be constructed during the summer of 2009. The Sundyne Gas compressor is on location as well as the 500KVA diesel generator that will be providing the electricity for the compressor and liquid pumps. Equipment that still needs to be purchased is as follows: Separation System, Pumps, Liquid Tanks, Surge tank for gas, Flow-Meters, and Instrumentation.

Detailed descriptions of these modification efforts appear in a progress presentation given in this brochure. A site plan showing the location of the various TUFFP and TUPDP test facilities on the North Campus is given in Fig. 1.

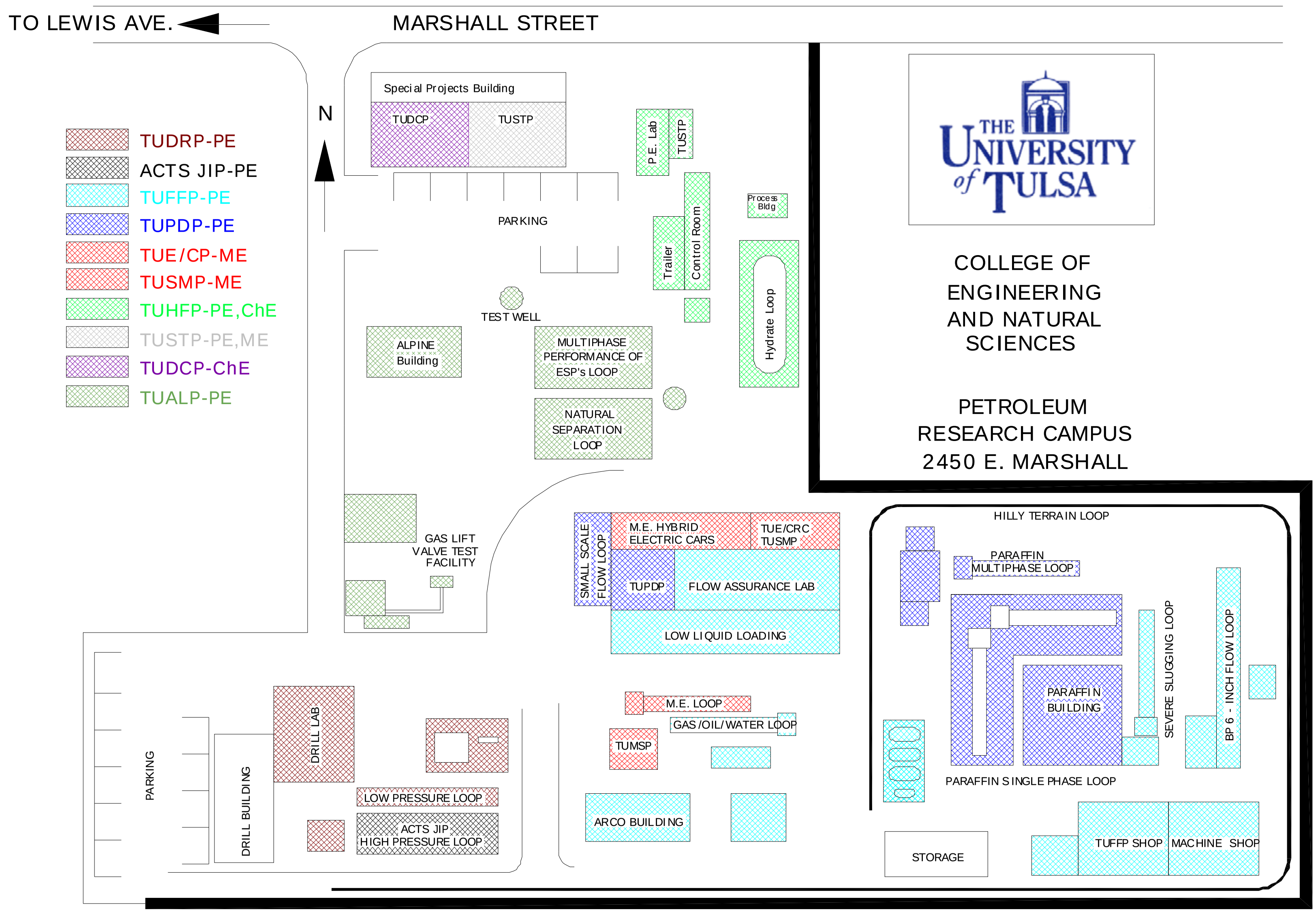


Figure 1 – Site Plan for the North Campus Research Facilities

Financial Status

TUFFP maintains separate accounts for industrial and U.S. government members. Thus, separate accounts are maintained for the MMS funds.

As of February 28, 2009, 16 of the 17 TUFFP members had paid their 2008 membership fees. We are pursuing Petronas to have them honor their contractual obligations. Moreover, 10 of the 16 TUFFP members have paid their 2009 membership fees. We would appreciate your prompt payment of the membership dues if your company has not yet paid the membership fee.

Table 3 presents a financial analysis of income and expenditures for the 2008 Industrial member account as of December 31, 2008. This serves as the unofficial closing budget for 2008. Also shown are previous 2008 budgets that have been reported to members. The total industry expenditures for 2008 were $875,035. This resulted in a balance on December 31, 2008 of $500,404 in the industry reserve account, assuming the successful collection of the 2008 membership fee from Petronas.

Table 4 presents a financial analysis of expenditures and income for the MMS Account for 2008. This account is used primarily for graduate student stipends. A balance of $5,770 is carried over to 2009.

The University of Tulsa waives up to 19 hours of tuition for each graduate student that is paid a stipend from the United States government, including both MMS and DOE funds. A total of 45 hours of tuition (equivalent of $32,850) was waived for 2008.

Tables 5-6 present the projected budgets and income for the Industrial, and MMS accounts for 2009. The 2009 TUFFP industrial membership is assumed to be 15 in this analysis. This will provide $720,000 of industrial membership income for 2009. The sum of the 2009 income and the reserve account is projected to be $1,220,404. Expenses for the industrial member account are estimated to be $1,092,280 leaving a positive balance of $128,124 at the end of 2009. The MMS account is expected to have a carryover of $17,671 at the end of 2009.

Table 3: TUFFP 2008 Industrial Budget

2008 TUFFP Industrial Account Budget Summary
(Prepared February 25, 2009)

Anticipated Reserve Fund Balance on January 1, 2008	**575,234**
Income for 2008	
2008 Membership Fees (16 @ $48,000 - excludes MMS)	**768,000**
2008 Membership Fees (1 @ 38,000)	**38,000**
Total Budget	**1,381,234**

Projected Budget/Expenditures for 2008

		Budget	Revised Budget 4/08	Expenses 2/11/09	Anticipated 2008 Expenses
90101	Principal Investigator - Sarica	24,392	25,907	21,494	27,930
90103	Co-Principal Investigator - Zhang	19,665		3,623	3,623
90600	Professional Salary - Jones	5,141	7,330	8,472	8,472
90601	Professional Salary - Li	13,505	28,854	22,568	24,651
90602	Professional Salary - Graham	5,237	15,785	15,506	15,506
90603	Professional Salary - Al-Sarkhi	32,500	38,750	16,646	6,458
90701	Technician - Miller	15,065		1,854	
90702	Technician - Waldron	6,575	11,428	11,218	9,869
90703	Technician - Kelsey	9,750	10,154	14,637	12,754
90800	Salaries - Part-time	4,290		519	
91000	Graduate Students - Monthly	50,100	65,000	60,700	60,745
91100	Students - Hourly	15,000	15,000	14,798	14,065
91800	Fringe Benefits (33%)	50,911	45,609	38,286	36,057
93100	General Supplies	3,000	3,000	284	500
93101	Research Supplies	100,000	100,000	87,679	100,000
93102	Copier/Printer Supplies	500	500	169	280
93104	Computer Software	4,000	4,000	502	1,200
93106	Office Supplies	2,000	2,000	2,071	3,000
93200	Postage/Shipping	500	500	377	650
93300	Printing/Duplicating	2,000	2,000	3,373	2,000
93400	Telecommunications	3,000	3,000	518	1,766
93500	Membership/Subscriptions	1,000	1,000	227	400
93600	Travel			-	
93601	Travel - Domestic	14,000	10,000	4,899	10,000
93602	Travel - Foreign	10,000	10,000	7,808	10,946
93606	Visa			241	241
93700	Entertainment (Advisory Board Meetings)	10,000	10,000	18,359	10,000
94803	Consultants		16,000	18,500	18,500
94813	Outside Services	20,000	20,000	81,989	68,278
95103	Equipment Rental			5,484	
95200	F&A (55.6%)	119,456	121,324	106,956	102,344
98901	Employee Recruiting	3,000	3,000	285	147
99001	Equipment	600,000	200,000	262,317	263,500
99002	Computers	8,000	8,000	769	4,000
99300	Bank Charges	40	40	-	-
81801	Tuition/Fees	30,306	53,103	45,893	55,342
81806	Graduate Fellowship			1,812	1,809
	Total Expenditures	**1,182,933**	**831,284**	**880,830**	**875,035**

Anticipated Reserve Fund Balance as of 12/31/08	**500,404**

Table 4: TUFFP 2008 MMS Budget

2008 TUFFP MMS Budget Summary
(Prepared February 11, 2009)

Reserve Balance as of 12/31/07	**$5,322**
2008 Budget	**40,000**
Total Budget	**45,322**

Projected Budget/Expenditures for 2008

		Budget	**2008 Expenditures**	
91000	Students - Monthly	25,600	25,600	
95200	F&A	14,234	13,952	
81801	Tuition/Fees			
Total Anticipated Expenditures as of 12/31/08		**39,834**	**39,552**	
Total Anticipated Reserve Fund Balance as of 12/31/08				**5,770**

Table 6: 2009 Projected TUFFP Industrial Budget

2009 TUFFP Industrial Account Budget Summary

(Prepared February 25, 2009)

Anticipated Reserve Fund Balance on January 1, 2009			**500,404**
			500,404
Income for 2009			
2009 Membership Fees (15 @ $48,000 - excludes MMS)			**720,000**
Total Budget			**1,220,404**

Projected Budget/Expenditures for 2009

		Budget	Revised Budget 2/11/09
90101	Principal Investigator - Sarica	29,252	29,074
90103	Co-Principal Investigator - Zhang		
90600	Professional Salary - Jones	11,596	11,728
90601	Professional Salary - Li	13,157	12,875
90602	Professional Salary - Graham	19,923	20,149
90603	Professional Salary - Polat	62,000	65,000
90701	Technician - Miller	9,943	10,016
90702	Technician - Waldron	16,826	16,517
90703	Technician - Kelsey	19,097	18,746
90800	Salaries - Part-time	-	
91000	Graduate Students - Monthly	58,100	62,850
91100	Students - Hourly	15,000	15,000
91800	Fringe Benefits (33%)	59,992	62,850
93100	General Supplies	3,000	3,000
93101	Research Supplies	100,000	100,000
93102	Copier/Printer Supplies	500	500
93104	Computer Software	4,000	4,000
93106	Office Supplies	2,000	2,000
93200	Postage/Shipping	500	500
93300	Printing/Duplicating	2,000	2,000
93400	Telecommunications	3,000	3,000
93500	Membership/Subscriptions	1,000	1,000
93600	Travel		
93601	Travel - Domestic	10,000	10,000
93602	Travel - Foreign	10,000	10,000
93606	Visa		
93700	Entertainment (Advisory Board Meetings)	10,000	10,000
94803	Consultants	16,000	18,500
94813	Outside Services	20,000	20,000
95200	F&A (55.6%)	141,721	141,868
98901	Employee Recruiting	3,000	3,000
99001	Equipment	600,000	400,000
99002	Computers	8,000	8,000
99300	Bank Charges	40	40
81801	Tuition/Fees	30,665	30,067
81806	Graduate Fellowship		
	Total Expenditures	**1,280,313**	**1,092,280**

Anticipated Reserve Fund Balance as of 12/31/09	**128,124**

Table 7: TUFFP Projected 2009 MMS Budget

2009 TUFFP MMS Budget Summary
(Prepared February 11, 2009)

Reserve Balance as of 12/31/08				**5,770**
2009 Budget				**48,000**
Total Budget				**53,770**
Projected Budget/Expenditures for 2009				
		Budget	**2009 Anticipated Expenditures**	
91000	Students - Monthly	27,900	23,200	
95200	F&A	15,512	12,899	
81801	Tuition/Fees			
Total Anticipated Expenditures as of 12/31/09		**43,412**	**36,099**	
Total Anticipated Reserve Fund Balance as of 12/31/09				**17,671**

Miscellaneous Information

Fluid Flow Projects Short Course

The 34th TUFFP "Two-Phase Flow in Pipes" short course offering is scheduled for May 18-22, 2009 in Tulsa, Oklahoma. For this short course to be self sustaining, at least 10 enrollees are needed. We urge our members to let us know soon if they plan to enroll people in the short course. The detailed registration information can be found at http://www.cese.utulsa.edu/programdetail.php?ID=134.

Holden Zhang Promoted to Associate Professor Rank with Tenure

We are proud and happy to announce that, recently, after a lengthy rigorous academic evaluation process, Dr. Hong-Quan (Holden) Zhang has been promoted to the rank of Associate Professor with tenure with unanimous recommendations from both the Petroleum Engineering faculty and the college tenure and promotion committees. Please join us in congratulating Holden for this milestone. Holden has been with TU for 10 years, 5 years as a Research Associate for TUFFP and TUPDP and 5 years as a tenure track faculty member in the Department of Petroleum Engineering. Currently, Dr. Zhang serves as Associate Director for TUFFP and Principal Investigator for the Tulsa University High-Viscosity Oil Projects (TUHOP). He has proven himself as an effective teacher, an excellent researcher and an invaluable asset to the department.

BHR Group Conference on Multiphase Technology

Since 1991, TUFFP has participated as a co-sponsor of BHR Group Conferences on Multiphase Production. TUFFP personnel participate in reviewing papers, serving as session chairs, and advertising the conference to our members. This conference is one of the premier international events providing delegates with opportunities to discuss new research and developments, and consider innovative solutions in the multiphase production area.

The 14th International Conference on Multiphase Technology, supported by IFP, Technology Initiatives and TUFFP, will be held 17-19 June 2009 in Cannes, France. The conference will benefit anyone engaged in the application, development and research of multiphase technology for the oil and gas industry. Applications in the oil and gas industry will also be of interest to engineers from other industries for which multiphase technology offers a novel solution to their problems. The conference will also be of particular value to designers, facility and operations engineers, consultants and researchers from operating, contracting, consultancy and technology companies. The scope of the conference includes a variety of subjects pertinent to Multiphase Production in both technology development and applications of existing technologies. The theme of the conference is *"Bigger, Deeper, Longer"*. Detailed information about the conference can be found in BHRg's website (www.brhgroup.com).

Publications & Presentations

Since the last Advisory Board meeting, the following publications and presentations have been made.

1) Atmaca, S., Alsarkhi, A., Zhang, H. Q. and Sarica, C.: "Characterization of Oil Water Flows in Inclined Pipes," SPE 115485, Accepted for Publication in *SPE Projects, Facilities & Construction Journal*, 2009.
2) Gokcal, B., Alsarkhi, A. and Sarica, C.: "Effects of High Viscosity on Drift Velocity for Inclined Pipes," Accepted for Publication in *SPE Projects, Facilities & Construction Journal*, 2009.
3) Al-Safran, E. Kappos, L. and Sarica, C. "Experimental and Numerical Investigation of Separator Pressure Fluctuation Effect on Terrain Slugging in a Hilly-Terrain Two-phase Flow Pipeline," *Journal of Energy Resources Technology* September 2008.
4) Vielma, M., Atmaca, S., Zhang, H. Q., and Sarica, C.: "Characterization of Oil/Water Flows in Horizontal Pipes," SPE 109591, *SPE Projects, Facilities & Construction Journal*, December 2008.

Paraffin Deposition Projects Activities

The third three-year phase of TUPDP continues. The studies concentrate on paraffin deposition characterization of single-phase turbulent flow, oil-water paraffin deposition, and gas-oil-water paraffin deposition.

TUHOP Activities

The Center of Research Excellence (TUCoRE) initiated by Chevron at The University of Tulsa funds several research projects on flow assurance topics. TUFFP researchers are involved in various TUCoRE activities. One such activity is on High-Viscosity Multiphase Flow (TUHOP). Chevron has also provided TU with $680,000 for improvement of an existing high pressure multiphase flow facility. Moreover, this research was leveraged by forming a new Joint Industry Project. Current members of the TUHOP JIP are BP, Chevron and Petrobras.

Two-Phase Flow Calendar

Several technical meetings, seminars, and short courses involving two-phase flow in pipes are scheduled for 2009. Table 9 lists meetings that would be of interest to TUFFP members.

Table 9

Meeting and Conference Calendar

2009

March 24	TUHOP Advisory Board Meeting, Tulsa, Oklahoma
	TUFFP Spring Workshop, Tulsa, Oklahoma
March 25	TUFFP Spring Advisory Board Meeting, Tulsa, Oklahoma
March 26	TUPDP Spring Advisory Board Meeting, Tulsa, Oklahoma
April 4 - 8	SPE Production and Operations Symposium, Oklahoma City, Oklahoma
May 4 - 7	Offshore Technology Conference, Houston, Texas
May 18 - 22	TUFFP Short Course, Tulsa, Oklahoma
May 31- June 5	2009 OMAE Meeting, Honolulu, Hawaii
June 17 - 19	BHRg's Multiphase Technology 2009, Cannes, France
September 29	TUHOP Advisory Board Meeting, Tulsa, Oklahoma
	TUFFP Fall Workshop, Tulsa, Oklahoma
September 30	TUFFP Advisory Board Meeting, Tulsa, Oklahoma
October 1	TUPDP Advisory Board Meeting, Tulsa, Oklahoma
October 4 - 7	SPE Annual Technical Conference and Exhibition, New Orleans, Louisiana
December 7 - 9	International Petroleum Technology Conference, Doha, Qatar

Appendix A

Fluid Flow Projects Deliverables[1]

1. "An Experimental Study of Oil-Water Flowing Mixtures in Horizontal Pipes," by M. S. Malinowsky (1975).

2. "Evaluation of Inclined Pipe Two-Phase Liquid Holdup Correlations Using Experimental Data," by C. M. Palmer (1975).

3. "Experimental Evaluation of Two-Phase Pressure Loss Correlations for Inclined Pipe," by G. A. Payne (1975).

4. "Experimental Study of Gas-Liquid Flow in a Pipeline-Riser Pipe System," by Z. Schmidt (1976).

5. "Two-Phase Flow in an Inclined Pipeline-Riser Pipe System," by S. Juprasert (1976).

6. "Orifice Coefficients for Two-Phase Flow Through Velocity Controlled Subsurface Safety Valves," by J. P. Brill, H. D. Beggs, and N. D. Sylvester (Final Report to American Petroleum Institute Offshore Safety and Anti-Pollution Research Committee, OASPR Project No. 1; September, 1976).

7. "Correlations for Fluid Physical Property Prediction," by M. E. Vasquez A. (1976).

8. "An Empirical Method of Predicting Temperatures in Flowing Wells," by K. J. Shiu (1976).

9. "An Experimental Study on the Effects of Flow Rate, Water Fraction and Gas-Liquid Ratio on Air-Oil-Water Flow in Horizontal Pipes," by G. C. Laflin and K. D. Oglesby (1976).

10. "Study of Pressure Drop and Closure Forces in Velocity- Type Subsurface Safety Valves," by H. D. Beggs and J. P. Brill (Final Report to American Petroleum Institute Offshore Safety and Anti-Pollution Research Committee, OSAPR Project No. 5; July, 1977).

11. "An Experimental Study of Two-Phase Oil-Water Flow in Inclined Pipes," by H. Mukhopadhyay (September 1, 1977).

12. "A Numerical Simulation Model for Transient Two-Phase Flow in a Pipeline," by M. W. Scoggins, Jr. (October 3, 1977).

13. "Experimental Study of Two-Phase Slug Flow in a Pipeline-Riser Pipe System," by Z. Schmidt (1977).

14. "Drag Reduction in Two-Phase Gas-Liquid Flow," (Final Report to American Gas Association Pipeline Research Committee; 1977).

15. "Comparison and Evaluation of Instrumentation for Measuring Multiphase Flow Variables in Pipelines," Final Report to Atlantic Richfield Co. by J. P. Brill and Z. Schmidt (January, 1978).

16. "An Experimental Study of Inclined Two-Phase Flow," by H. Mukherjee (December 30, 1979).

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[1] Completed TUFFP Projects – each project consists of three deliverables – report, data and software. Please see the TUFFP website

17. "An Experimental Study on the Effects of Oil Viscosity, Mixture Velocity and Water Fraction on Horizontal Oil-Water Flow," by K. D. Oglesby (1979).

18. "Experimental Study of Gas-Liquid Flow in a Pipe Tee," by S. E. Johansen (1979).

19. "Two Phase Flow in Piping Components," by P. Sookprasong (1980).

20. "Evaluation of Orifice Meter Recorder Measurement Errors in Lower and Upper Capacity Ranges," by J. Fujita (1980).

21. "Two-Phase Metering," by I. B. Akpan (1980).

22. "Development of Methods to Predict Pressure Drop and Closure Conditions for Velocity-Type Subsurface Safety Valves," by H. D. Beggs and J. P. Brill (Final Report to American Petroleum Institute Offshore Safety and Anti-Pollution Research Committee, OSAPR Project No. 10; February, 1980).

23. "Experimental Study of Subcritical Two-Phase Flow Through Wellhead Chokes," by A. A. Pilehvari (April 20, 1981).

24. "Investigation of the Performance of Pressure Loss Correlations for High Capacity Wells," by L. Rossland (1981).

25. "Design Manual: Mukherjee and Brill Inclined Two-Phase Flow Correlations," (April, 1981).

26. "Experimental Study of Critical Two-Phase Flow through Wellhead Chokes," by A. A. Pilehvari (June, 1981).

27. "Experimental Study of Pressure Wave Propagation in Two-Phase Mixtures," by S. Vongvuthipornchai (March 16, 1982).

28. "Determination of Optimum Combination of Pressure Loss and PVT Property Correlations for Predicting Pressure Gradients in Upward Two-Phase Flow," by L. G. Thompson (April 16, 1982).

29. "Hydrodynamic Model for Intermittent Gas Lifting of Viscous Oils," by O. E. Fernandez (April 16, 1982).

30. "A Study of Compositional Two-Phase Flow in Pipelines," by H. Furukawa (May 26, 1982).

31. "Supplementary Data, Calculated Results, and Calculation Programs for TUFFP Well Data Bank," by L. G. Thompson (May 25, 1982).

32. "Measurement of Local Void Fraction and Velocity Profiles for Horizontal Slug Flow," by P. B. Lukong (May 26, 1982).

33. "An Experimental Verification and Modification of the McDonald-Baker Pigging Model for Horizontal Flow," by S. Barua (June 2, 1982).

34. "An Investigation of Transient Phenomena in Two-Phase Flow," by K. Dutta-Roy (October 29, 1982).

35. "A Study of the Heading Phenomenon in Flowing Oil Wells," by A. J. Torre (March 18, 1983).

36. "Liquid Holdup in Wet-Gas Pipelines," by K. Minami (March 15, 1983).

37. "An Experimental Study of Two-Phase Oil-Water Flow in Horizontal Pipes," by S. Arirachakaran (March 31, 1983).

38. "Simulation of Gas-Oil Separator Behavior Under Slug Flow Conditions," by W. F. Giozza (March 31, 1983).

39. "Modeling Transient Two-Phase Flow in Stratified Flow Pattern," by Y. Sharma (July, 1983).

40. "Performance and Calibration of a Constant Temperature Anemometer," by F. Sadeghzadeh (August 25, 1983).

41. "A Study of Plunger Lift Dynamics," by L. Rosina (October 7, 1983).

42. "Evaluation of Two-Phase Flow Pressure Gradient Correlations Using the A.G.A. Gas-Liquid Pipeline Data Bank," by E. Caetano F. (February 1, 1984).

43. "Two-Phase Flow Splitting in a Horizontal Pipe Tee," by O. Shoham (May 2, 1984).

44. "Transient Phenomena in Two-Phase Horizontal Flowlines for the Homogeneous, Stratified and Annular Flow Patterns," by K. Dutta-Roy (May 31, 1984).

45. "Two-Phase Flow in a Vertical Annulus," by E. Caetano F. (July 31, 1984).

46. "Two-Phase Flow in Chokes," by R. Sachdeva (March 15, 1985).

47. "Analysis of Computational Procedures for Multi-Component Flow in Pipelines," by J. Goyon (June 18, 1985).

48. "An Investigation of Two-Phase Flow Through Willis MOV Wellhead Chokes," by D. W. Surbey (August 6, 1985).

49. "Dynamic Simulation of Slug Catcher Behavior," by H. Genceli (November 6, 1985).

50. "Modeling Transient Two-Phase Slug Flow," by Y. Sharma (December 10, 1985).

51. "The Flow of Oil-Water Mixtures in Horizontal Pipes," by A. E. Martinez (April 11, 1986).

52. "Upward Vertical Two-Phase Flow Through An Annulus," by E. Caetano F. (April 28, 1986).

53. "Two-Phase Flow Splitting in a Horizontal Reduced Pipe Tee," by O. Shoham (July 17, 1986).

54. "Horizontal Slug Flow Modeling and Metering," by G. E. Kouba (September 11, 1986).

55. "Modeling Slug Growth in Pipelines," by S. L. Scott (October 30, 1987).

56. "RECENT PUBLICATIONS" - A collection of articles based on previous TUFFP research reports that have been published or are under review for various technical journals (October 31, 1986).

57. "TUFFP CORE Software Users Manual, Version 2.0," by Lorri Jefferson, Florence Kung and Arthur L. Corcoran III (March 1989)

58. "Simplified Modeling and Simulation of Transient Two Phase Flow in Pipelines," by Y. Taitel (April 29, 1988).

59. "RECENT PUBLICATIONS" - A collection of articles based on previous TUFFP research reports that have been published or are under review for various technical journals (April 19, 1988).

60. "Severe Slugging in a Pipeline-Riser System, Experiments and Modeling," by S. J. Vierkandt (November 1988).

61. "A Comprehensive Mechanistic Model for Upward Two-Phase Flow," by A. Ansari (December 1988).

62. "Modeling Slug Growth in Pipelines" Software Users Manual, by S. L. Scott (June 1989).

63. "Prudhoe Bay Large Diameter Slug Flow Experiments and Data Base System" Users Manual, by S. L. Scott (July 1989).

64. "Two-Phase Slug Flow in Upward Inclined Pipes", by G. Zheng (Dec. 1989).

65. "Elimination of Severe Slugging in a Pipeline-Riser System," by F. E. Jansen (May 1990).

66. "A Mechanistic Model for Predicting Annulus Bottomhole Pressures for Zero Net Liquid Flow in Pumping Wells," by D. Papadimitriou (May 1990).

67. "Evaluation of Slug Flow Models in Horizontal Pipes," by C. A. Daza (May 1990).

68. "A Comprehensive Mechanistic Model for Two-Phase Flow in Pipelines," by J. J. Xiao (Aug. 1990).

69. "Two-Phase Flow in Low Velocity Hilly Terrain Pipelines," by C. Sarica (Aug. 1990).

70. “Two-Phase Slug Flow Splitting Phenomenon at a Regular Horizontal Side-Arm Tee,” by S. Arirachakaran (Dec. 1990)

71. "RECENT PUBLICATIONS" - A collection of articles based on previous TUFFP research reports that have been published or are under review for various technical journals (May 1991).

72. "Two-Phase Flow in Horizontal Wells," by M. Ihara (October 1991).

73. "Two-Phase Slug Flow in Hilly Terrain Pipelines," by G. Zheng (October 1991).

74. "Slug Flow Phenomena in Inclined Pipes," by I. Alves (October 1991).

75. "Transient Flow and Pigging Dynamics in Two-Phase Pipelines," by K. Minami (October 1991).

76. "Transient Drift Flux Model for Wellbores," by O. Metin Gokdemir (November 1992).

77. "Slug Flow in Extended Reach Directional Wells," by Héctor Felizola (November 1992).

78. "Two-Phase Flow Splitting at a Tee Junction with an Upward Inclined Side Arm," by Peter Ashton (November 1992).

79. "Two-Phase Flow Splitting at a Tee Junction with a Downward Inclined Branch Arm," by Viswanatha Raju Penmatcha (November 1992).

80. "Annular Flow in Extended Reach Directional Wells," by Rafael Jose Paz Gonzalez (May 1994).

81. "An Experimental Study of Downward Slug Flow in Inclined Pipes," by Philippe Roumazeilles (November 1994).

82. "An Analysis of Imposed Two-Phase Flow Transients in Horizontal Pipelines Part-1 Experimental Results," by Fabrice Vigneron (March 1995).

83. "Investigation of Single Phase Liquid Flow Behavior in a Single Perforation Horizontal Well," by Hong Yuan (March 1995).

84. "1995 Data Documentation User's Manual", (October 1995).

85. "Recent Publications" A collection of articles based on previous TUFFP research reports that have been published or are under review for various technical journals (February 1996).

86. "1995 Final Report - Transportation of Liquids in Multiphase Pipelines Under Low Liquid Loading Conditions", Final report submitted to Penn State University for subcontract on GRI Project.

87. "A Unified Model for Stratified-Wavy Two-Phase Flow Splitting at a Reduced Tee Junction with an Inclined Branch Arm", by Srinagesh K. Marti (February 1996).

88. "Oil-Water Flow Patterns in Horizontal Pipes", by José Luis Trallero (February 1996).

89. "A Study of Intermittent Flow in Downward Inclined Pipes" by Jiede Yang (June 1996).

90. "Slug Characteristics for Two-Phase Horizontal Flow", by Robert Marcano (November 1996).

91. "Oil-Water Flow in Vertical and Deviated Wells", by José Gonzalo Flores (October 1997).

92. "1997 Data Documentation and Software User's Manual", by Avni S. Kaya, Gerad Gibson and Cem Sarica (November 1997).

93. "Investigation of Single Phase Liquid Flow Behavior in Horizontal Wells", by Hong Yuan (March 1998).

94. "Comprehensive Mechanistic Modeling of Two-Phase Flow in Deviated Wells" by Avni Serdar Kaya (December 1998).

95. "Low Liquid Loading Gas-Liquid Two-Phase Flow in Near-Horizontal Pipes" by Weihong Meng (August 1999).

96. "An Experimental Study of Two-Phase Flow in a Hilly-Terrain Pipeline" by Eissa Mohammed Al-Safran (August 1999).

97. "Oil-Water Flow Patterns and Pressure Gradients in Slightly Inclined Pipes" by Banu Alkaya (May 2000).

98. "Slug Dissipation in Downward Flow – Final Report" by Hong-Quan Zhang, Jasmine Yuan and James P. Brill (October 2000).

99. "Unified Model for Gas-Liquid Pipe Flow – Model Development and Validation" by Hong-Quan Zhang (January 2002).

100. "A Comprehensive Mechanistic Heat Transfer Model for Two-Phase Flow with High-Pressure Flow Pattern Validation" Ph.D. Dissertation by Ryo Manabe (December 2001).

101. "Revised Heat Transfer Model for Two-Phase Flow" Final Report by Qian Wang (March 2003).

102. "An Experimental and Theoretical Investigation of Slug Flow Characteristics in the Valley of a Hilly-Terrain Pipeline" Ph.D. Dissertation by Eissa Mohammed Al-safran (May 2003).

103. "An Investigation of Low Liquid Loading Gas-Liquid Stratified Flow in Near-Horizontal Pipes" Ph.D. Dissertation by Yongqian Fan.

104. “Severe Slugging Prediction for Gas-Oil-Water Flow in Pipeline-Riser Systems,” M.S. Thesis by Carlos Andrés Beltrán Romero (2005)

105. “Droplet-Homophase Interaction Study (Development of an Entrainment Fraction Model) – Final Report,” Xianghui Chen (2005)

106. “Effects of High Oil Viscosity on Two-Phase Oil-Gas Flow Behavior in Horizontal Pipes” M.S. Thesis by Bahadir Gokcal (2005)

107. “Characterization of Oil-Water Flows in Horizontal Pipes” M.S. Thesis by Maria Andreina Vielma Paredes (2006)

108. “Characterization of Oil-Water Flows in Inclined Pipes” M.S. Thesis by Serdar Atmaca (2007).

109. “An Experimental Study of Low Liquid Loading Gas-Oil-Water Flow in Horizontal Pipes” M.S. Thesis by Hongkun Dong (2007).

110. “An Experimental and theoretical Investigation of Slug Flow for High Oil Viscosity in Horizontal Pipes” Ph.D. Dissertation by Bahadir Gokcal (2008).

Appendix B

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Appendix C

History of Fluid Flow Projects Membership

		1973	
1.	TRW Reda Pump	12 Jun. '72	T: 21 Oct. '77
2.	Pemex	15 Jun. '72	T: 30 Sept. '96 R: Dec '97 **Current**
3.	Getty Oil Co.	19 Jun. '72	T: 11 Oct. '84 with sale to Texaco
4.	Union Oil Co. of California	7 Jul. '72	T: for 2001
5.	Intevep	3 Aug. '72	TR: from CVP in '77; T: 21 Jan '05 for 2006
6.	Marathon Oil Co.	3 Aug. '72	T: 17 May '85 R: 25 June '90 T: 14 Sept. '94 R: 3 June '97 **Current**
7.	Arco Oil and Gas Co.	7 Aug. '72	T: 08 Dec. '97
8.	AGIP	6 Sep. '72	T: 18 Dec. '74
9.	Otis Engineering Corp.	4 Oct. '72	T: 15 Oct. '82
10.	ConocoPhillips, Inc.	5 Oct. '72	T: Aug. '85 R: 5 Dec. '86 **Current**
11.	Mobil Research and Development Corp.	13 Oct. '72	T: 27 Sep. 2000
12.	Camco, Inc.	23 Oct. '72	T: 15 Jan. '76 R: 14 Mar. '79 T: 5 Jan. '84
13.	Crest Engineering, Inc.	27 Oct. '72	T: 14 Nov. '78 R: 19 Nov. '79 T: 1 Jun. '84
14.	Chevron	3 Nov. '72	**Current**
15.	Aminoil	9 Nov. '72	T: 1 Feb. '77

16.	Compagnie Francaise des Petroles (TOTAL)	6 Dec. '72	T: 22 Mar. '85 R: 23 Oct. '90 T: 18 Sep. '01 for 2002 R: 18 Nov. '02 **Current**
17.	Oil Service Co. of Iran	19 Dec. '72	T: 20 Dec. '79
18.	Sun Exploration and Production Co.	4 Jan. '73	T: 25 Oct. '79 R: 13 Apr. '82 T: 6 Sep. '85
19.	Amoco Production Co. (now as BP Amoco)	18 May '73	
20.	Williams Brothers Engrg. Co.	25 May '73	T: 24 Jan. '83
	1974		
21.	Gulf Research and Development Co.	20 Nov. '73	T: Nov. '84 with sale to Chevron
22.	El Paso Natural Gas Co.	17 Dec. '73	T: 28 Oct. '77
23.	Arabian Gulf Exploration Co.	27 Mar. '74	T: 24 Oct. '82
24.	ExxonMobil Upstream Research	27 Mar. '74	T: 16 Sep. '86 R: 1 Jan. '88 T: 27 Sep. 2000 R: 2007 **Current**
25.	Bechtel, Inc.	29 May '74	T: 14 Dec. '76 R: 7 Dec. '78 T: 17 Dec. '84
26.	Saudi Arabian Oil Co.	11 Jun. '74	T: for 1999
27.	Petrobras	6 Aug. '74	T: for 2000 R: for 2005 **Current**
	1975		
28.	ELF Exploration Production (now as TotalFina Elf)	24 Jul. '74	T: 24 Feb. '76 Tr. from Aquitaine Co. of Canada 19 Mar. '81 T: 29 Jan. '87 R: 17 Dec. '91
29.	Cities Service Oil and Gas Corp.	21 Oct. '74	T: 25 Oct. '82 R: 27 Jun. '84 T: 22 Sep. '86

30.	Texas Eastern Transmission Corp.	19 Nov. '74	T: 23 Aug. '82
31.	Aquitaine Co. of Canada, Ltd.	12 Dec. '74	T: 6 Nov. '80
32.	Texas Gas Transmission Corp.	4 Mar. '75	T: 7 Dec. '89
	1976		
33.	Panhandle Eastern Pipe Line Co.	15 Oct. '75	T: 7 Aug. '85
34.	Phillips Petroleum Co.	10 May '76	T: Aug. 94 R: Mar 98 T: 2002
	1977		
35.	N. V. Nederlandse Gasunie	11 Aug. '76	T: 26 Aug. '85
36.	Columbia Gas System Service Corp.	6 Oct. '76	T: 15 Oct. '85
37.	Consumers Power Co.	11 Apr. '77	T: 14 Dec. '83
38.	ANR Pipeline Co.	13 Apr. '77	TR: from Michigan- Wisconsin Pipeline Co. in 1984 T: 26 Sep. '84
39.	Scientific Software-Intercomp	28 Apr. '77	TR: to Kaneb from Intercomp 16 Nov. '77 TR: to SSI in June '83 T: 23 Sep. '86
40.	Flopetrol/Johnston-Schlumberger	5 May '77	T: 8 Aug. '86
	1978		
41.	Norsk Hydro a.s	13 Dec. '77	T: 5 Nov. '82 R: 1 Aug. '84 T: 8 May ‘96
42.	Dresser Industries Inc.	7 Jun. '78	T: 5 Nov. '82
	1979		
43.	Sohio Petroleum Co.	17 Nov. '78	T: 1 Oct. '86
44.	Esso Standard Libya	27 Nov. '78	T: 2 Jun. '82
45.	Shell Internationale Petroleum MIJ B.V. (SIPM)	30 Jan. '79	T: Sept. 98 for 1999
	1980		
46.	Fluor Ocean Services, Inc.	23 Oct. '79	T: 16 Sep. '82
47.	Texaco	30 Apr. '80	T: 20 Sep. ’01 for 2002
48.	BG Technology (Advantica)	15 Sep. '80	T: 2003

	1981		
49.	Det Norske Veritas	15 Aug. '80	T: 16 Nov. '82
	1982		
50.	Arabian Oil Co. Ltd.	11 May '82	T: Oct.'01 for 2002
51.	Petro Canada	25 May '82	T:28 Oct. '86
52.	Chiyoda	3 Jun. '82	T: 4 Apr '94
53.	BP	7 Oct. '81	**Current**
	1983		
54.	Pertamina	10 Jan. '83	T: for 2000 R: March 2006
	1984		
55.	Nippon Kokan K. K.	28 Jun. '83	T: 5 Sept. '94
56.	Britoil	20 Sep. '83	T: 1 Oct. '88
57.	TransCanada Pipelines	17 Nov. '83	T:30 Sep. '85
58.	Natural Gas Pipeline Co. of America (Midcon Corp.)	13 Feb. '84	T:16 Sep. '87
59.	JGC Corp.	12 Mar. '84	T: 22 Aug. '94
	1985		
60.	STATOIL	23 Oct. '85	T:16 Mar. '89
	1986		
61.	JOGMEC (formerly Japan National Oil Corp.)	3 Oct. '86	T: 2003 R: 2007 **Current**
	1988		
62.	China National Oil and Gas Exploration and Development Corporation	29 Aug. '87	T:17 Jul. '89
63.	Kerr McGee Corp.	8 Jul. '88	T:17 Sept. '92
	1989		
64.	Simulation Sciences, Inc.	19 Dec. '88	T: for 2001
	1991		
65.	Advanced Multiphase Technology	7 Nov. '90	T:28 Dec. '92

66.	Petronas	1 Apr. '91	T: 02 Mar. 98 R: 1 Jan 2001 T: Nov. 2008 for 2009
	1992		
67.	Instituto Colombiano Del Petroleo	19 July '91	T: 3 Sep. '01 for 2002
68.	Institut Francais Du Petrole	16 July. '91	T: 8 June 2000
69.	Oil & Natural Gas Commission of India	27 Feb. '92	T: Sept. 97 for 1998
	1994		
70.	Baker Jardine & Associates	Dec. '93	T: 22 Sept. '95 for 1996
	1998		
71.	Baker Atlas	Dec. 97	**Current**
72.	Minerals Management Service (Department of Interior's)	May. 98	**Current**
	2002		
73.	Schlumberger Overseas S.A.	Aug. 02	**Current**
74.	Saudi Aramco	Mar. 03	T: for 2007
	2004		
75.	YUKOS	Dec. '03	T: 2005
76.	Landmark Graphics	Oct. '04	T: 2008
	2005		
77.	Rosneft	July '05	**Current**
	2006		
78.	Tenaris		T: Sept 2008 – for 2009
79.	Shell Global		**Current**
80.	Kuwait Oil Company		**Current**
	2009		
81.	SPT		**Current**

Note: T = Terminated; R = Rejoined; and TR = Transferred

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