1989

Heat transfer and combustion in a two-bed fluidized combustor

James Everett Foley
Iowa State University

Follow this and additional works at: http://lib.dr.iastate.edu/rtd
Part of the Mechanical Engineering Commons

Recommended Citation
INFORMATION TO USERS

The most advanced technology has been used to photograph and reproduce this manuscript from the microfilm master. UMI films the text directly from the original or copy submitted. Thus, some thesis and dissertation copies are in typewriter face, while others may be from any type of computer printer.

The quality of this reproduction is dependent upon the quality of the copy submitted. Broken or indistinct print, colored or poor quality illustrations and photographs, print bleedthrough, substandard margins, and improper alignment can adversely affect reproduction.

In the unlikely event that the author did not send UMI a complete manuscript and there are missing pages, these will be noted. Also, if unauthorized copyright material had to be removed, a note will indicate the deletion.

Oversize materials (e.g., maps, drawings, charts) are reproduced by sectioning the original, beginning at the upper left-hand corner and continuing from left to right in equal sections with small overlaps. Each original is also photographed in one exposure and is included in reduced form at the back of the book. These are also available as one exposure on a standard 35mm slide or as a 17" x 23" black and white photographic print for an additional charge.

Photographs included in the original manuscript have been reproduced xerographically in this copy. Higher quality 6" x 9" black and white photographic prints are available for any photographs or illustrations appearing in this copy for an additional charge. Contact UMI directly to order.
Heat transfer and combustion in a two-bed fluidized combustor

Foley, James Everett, Ph.D.

Iowa State University, 1989
Heat transfer and combustion in a
two-bed fluidized combustor

by

James Everett Foley

A Dissertation Submitted to the
Graduate Faculty in Partial Fulfillment of the
Requirements for the Degree of
DOCTOR OF PHILOSOPHY

Major: Mechanical Engineering

Approved:

Signature was redacted for privacy.

In Charge of Major Work

Signature was redacted for privacy.

For the Major Department

Signature was redacted for privacy.

For the Graduate College

Iowa State University
Ames, Iowa
1989
# TABLE OF CONTENTS

| ACKNOWLEDGEMENTS                           | viii |
| NOMENCLATURE                              | ix  |
| 1 INTRODUCTION                            | 1   |
| 2 BACKGROUND                              | 4   |
| 3 HEAT TRANSFER MODEL                     | 11  |
| 3.1 Energy Balance Equations              | 12  |
| 3.2 Fluidized Bed Convection Coefficients | 14  |
| 3.3 FORTRAN Program                       | 17  |
| 3.4 Heat Transfer Model Predictions       | 17  |
| 4 COMBUSTION MODEL                        | 23  |
| 4.1 Mathematical Formulation              | 25  |
| 4.2 FORTRAN Program                       | 31  |
| 4.3 Combustion Model Predictions          | 32  |
| 5 EXPERIMENTAL APPARATUS                  | 36  |
| 5.1 Two-Bed Fluidized Combustor           | 36  |
5.2 Fuel Feed System ........................................ 38
5.3 Air and Water Metering System ......................... 39
5.4 Flue-Gas Analysis System .................................. 42
5.5 Data Acquisition System .................................. 45

6  EXPERIMENTAL PROCEDURE .............................. 49
6.1 Load Turndown Tests ................................... 49
6.2 Fuel Preparation ........................................... 49
6.3 Start-Up Procedure ...................................... 53
6.4 Data Acquisition ......................................... 54
6.5 Ash Analysis ............................................... 54
6.6 Coal Feed Rate Determination ............................ 55

7  RESULTS ....................................................... 56
7.1 Load Turndown ............................................. 56
7.2 Emissions .................................................. 63
7.3 Estimate of Largest Practical Combustor ................. 66
7.4 Comparison with Heat Transfer Model .................. 67
7.5 Comparison with Combustion Model .................... 72

8  CONCLUSIONS ............................................... 78

9  BIBLIOGRAPHY ................................................ 80

10 APPENDIX A. HEAT TRANSFER COMPUTER MODEL
    LISTING ..................................................... 83
11 APPENDIX B. DERIVATION OF COMBUSTION MODEL 105

12 APPENDIX C. COMBUSTION COMPUTER MODEL LISTING 110

13 APPENDIX D. COMBUSTION MASS BALANCES 116
LIST OF FIGURES

Figure 2.1: Heat transfer rate and combustion rate vs. fluidization velocity 6
Figure 2.2: Two-bed fluidized combustor 8
Figure 3.1: Theoretical effect of particle size on load turndown 19
Figure 3.2: Theoretical effect of bed width on load turndown 21
Figure 4.1: Effect of fuel flow rate and equivalence ratio on fraction of fuel burned in-bed 34
Figure 5.1: Experimental two-bed fluidized combustor 37
Figure 5.2: Coal-water mixture feed system 40
Figure 5.3: Experimental apparatus 41
Figure 5.4: Photograph of the experimental apparatus 43
Figure 5.5: Gas sampling system 44
Figure 5.6: Data acquisition system 46
Figure 5.7: Pressure measurement system 48
Figure 7.1: Overall heat transfer coefficient vs. fluidization velocity 62
Figure 7.2: Sulfur retention vs. combustion bed temperature 64
Figure 7.3: Nitrogen oxide emission vs. combustion bed temperature 65
Figure 7.4: Energy released in-bed vs. annular air flow rate – experiment vs. two-phase theory prediction ............................... 68

Figure 7.5: Convection coefficient vs. fluidization velocity – experimental data ................................................................. 70

Figure 7.6: Energy released in-bed vs. annular air flow rate – experiment vs. experimental heat transfer data ............................... 71

Figure 7.7: Fraction of fuel burned in bed - predicted vs. observed ............................................................. 73

Figure 7.8: Predicted fraction of fuel burned in-bed and sources of incomplete combustion vs. fuel mass flow rate ..................... 75
LIST OF TABLES

Table 6.1: Illinois No. 5 coal analysis ............................................ 51
Table 7.1: Load turndown test results for crushed coal .................. 57
Table 7.2: Load turndown results for briquettes ............................ 59
Table 7.3: Load turndown results for CWLM ................................. 60
Table 7.4: Fuel flow rate - combustion model predictions vs. experi­
mental data ................................................................................. 77
ACKNOWLEDGEMENTS

This work was performed for the United States Department of Energy, through the Morgantown Energy Technology Center, Morgantown, West Virginia, under contract no. DE-AC21-86MC23248. I wish to acknowledge the financial support received from the Iowa State Mining and Minerals Resources Research Institute. I wish to thank the members of my program of study committee, Dr. Patrick Kavanagh, Dr. Thomas Wheelock, Dr. Gerald Colver, Dr. Gregory Maxwell, and Dr. George Burnet, for their helpful suggestions and criticisms in my graduate program. I thank Rod Oathout, Dean DeCock, Nathan Rosenboom, Marlin Kruse, and Roy Johansen for their assistance in performing experiments and analyzing data. I thank David Pflum for helpful discussions concerning fluidized bed modelling and William Buttermore for his expert assistance in fuel analysis and preparation.

I want to express special appreciation to my major professor, Dr. Robert Brown, who must be the hardest working and most understanding person with whom I have had the pleasure of working. I will be honored to be considered one of his peers.

Finally, I wish to give my greatest appreciation to my parents, who provide the love, patience, and encouragement that allow me to set high goals and try to attain them.
NOMENCLATURE

\( a_b \) = bubble area-to-volume ratio, \( m^2/m^3 \)

\( A_a \) = cross-sectional area of annular bed, \( m^2 \)

\( A_c \) = cross-sectional area of combustion bed, \( m^2 \)

\( A_o \) = area of distributor plate per orifice, \( m^2 \)

\( \text{Ar} \) = Archimedes number = \( \frac{\rho g (\rho_s - \rho_g) g d_p^3}{\mu_g^2} \)

\( c_g \) = specific heat of fluidizing air, \( J/kgK \)

\( c_{mf} \) = specific heat of emulsion phase, \( J/kgK \)

\( \bar{c}_{pa} \) = average specific heat of heat transfer air, \( J/kgK \)

\( \bar{c}_{pc} \) = average specific heat of combustion air, \( J/kgK \)

\( c_{pw} \) = specific heat of water, \( J/kgK \)

\( c_s \) = specific heat of solid particle, \( J/kgK \)

\( C_{CO} \) = carbon monoxide concentration, \( mole/m^3 \)

\( CO_2 \) = oxygen concentration, \( mole/m^3 \)

\( d_b \) = bubble diameter, \( m \)

\( d_p \) = average particle diameter, \( m \)

\( d \) = char particle diameter, \( m \)

\( D \) = column diameter, \( m \)

\( D_a \) = annular bed outside diameter, \( m \)
\[ D_c = \text{combustion bed diameter, } m \]
\[ D_g = \text{gas molecular diffusivity, } m^2/s \]
\[ g = \text{acceleration due to gravity, } m/s^2 \]
\[ \dot{G}_{CO} = \text{rate of carbon monoxide production, } mole/s \]
\[ \dot{G}_{O_2} = \text{rate of oxygen production, } mole/s \]
\[ h = \text{heat transfer coefficient, } W/m^2K \]
\[ h_a = \text{annular bed convection coefficient, } W/m^2K \]
\[ h_{aw} = \text{radiation heat transfer coefficient between wall and annular bed, } W/m^2K \]
\[ h_{aw} = \text{radiation heat transfer coefficient between water wall and annular bed, } W/m^2K \]
\[ h_{av} = \text{reciprocal fluidized bed resistance, } W/m^2K \]
\[ h_c = \text{combustion bed convection coefficient, } W/m^2K \]
\[ h_{ciw} = \text{radiation heat transfer coefficient between wall and combustion bed, } W/m^2K \]
\[ h_f = \text{reciprocal film resistance, } W/m^2K \]
\[ h_{fg} = \text{latent heat of vaporization, } kJ/kg \]
\[ h_{gc} = \text{gas convection coefficient, } W/m^2K \]
\[ h_p = \text{reciprocal packet resistance, } W/m^2K \]
\[ h_{pc} = \text{particle convection coefficient, } W/m^2K \]
\[ h_w = \text{water heat transfer coefficient, } W/m^2K \]
\[ h_{wall} = \text{reciprocal wall resistance, } W/m^2K \]
\[ H_c = \text{fluidized combustion bed height, } m \]
$H_f$ = fluidized bed height, m
$H_j$ = water jacket height, m
$H_{max}$ = maximum bed expansion height, m
$H_{min}$ = fluidized bed height at minimum fluidization, m
$\Delta H_{fC}$ = enthalpy of reaction for $C'(s) + O_2 \rightarrow CO_2$, J/mole
$\Delta H_{fCO}$ = enthalpy of reaction for $CO + \frac{1}{2}O_2 \rightarrow CO_2$, J/mole
$k_e^0$ = defluidized bed conductivity, W/mK
$k_g$ = gas conductivity, W/mK
$k_{mf}$ = minimally fluidized bed conductivity, W/mK
$k_q$ = interphase mass transfer coefficient, m/s
$k_s$ = solid particle conductivity, W/mK
$L$ = characteristic surface height, m
$m$ = empirical film coefficient constant
$m_c$ = bed carbon loading, kg
$\dot{m}_{dc}$ = dry coal mass flow rate, kg/hr
$\dot{m}_{f}$ = fuel mass flow rate, kg/hr
$\dot{m}_{w}$ = water mass flow rate, kg/s
$M_C'$ = molecular weight of carbon, 12.011 kg/mole
$M_{CO}$ = molecular weight of carbon monoxide, 28.011 kg/mole
$M_{H2O}$ = molecular weight of water, 18.016 kg/mole
$n$ = rate of oxygen diffusion to individual char particle, mole/s
$N$ = number of char particles in bed
$N_{C'aO}$ = moles of calcium oxide produced per lbm of fuel
\[ NCaSO_4 = \text{moles of calcium sulfate produced per lbm of fuel} \]
\[ N_{dg} = \text{moles of dry gas produced per lbm of fuel} \]
\[ N_{H_2O} = \text{moles of water produced per lbm of fuel} \]
\[ N_{O_2} = \text{moles of oxygen produced per lbm of fuel} \]
\[ Pr_w = \text{water Prandtl number} \]
\[ q = \text{heat flux, } W/m^2 \]
\[ q_b = \text{boiling water flux, } W/m^2 \]
\[ q_{max} = \text{maximum heat flux from combustor, } W/m^2 \]
\[ q_{min} = \text{minimum heat flux from combustor, } W/m^2 \]
\[ Q_{aa} = \text{energy advection rate from annular bed, } W \]
\[ Q_{ca} = \text{energy advection rate from combustion bed, } W \]
\[ Q_{comb} = \text{total energy release rate of combustor, } W \]
\[ Q_{cond} = \text{energy conduction rate across annular bed, } W \]
\[ Q_{cr} = \text{energy radiated from combustion bed to dividing wall, } W \]
\[ Q_{cw} = \text{energy convected from combustion bed to dividing wall, } W \]
\[ Q_{in} = \text{rate of energy release in bed, } W \]
\[ Q_w = \text{heat transfer rate to water jacket, } W \]
\[ Sh = \text{Sherwood number} = \frac{k_g d}{D_g} \]
\[ T_a = \text{annular bed temperature, } K \]
\[ T_c = \text{combustion bed temperature, } K \]
\[ T_g = \text{gas temperature, } K \]
\[ T_{iw} = \text{dividing wall temperature, } K \]
\[ T_o = \text{inlet air temperature, } K \]
T_{ow} = \text{water wall temperature, } K
T_{sat} = \text{saturation temperature of water, } K
T_{w} = \text{water temperature, } K
u = \text{superficial air velocity, } m/s
u_b = \text{bubble rise velocity, } m/s
u_{mf} = \text{minimum fluidization velocity, } m/s
u_{co} = \text{air inlet velocity to central bed, } m/s
u_{ao} = \text{air inlet velocity to annular bed, } m/s
U_{beds} = \text{overall heat transfer coefficient, } W/K
\dot{V}_c = \text{molar flow rate of unburned carbon, mole/s}
W_c = \text{lbm of dry coal per lbm of fuel}
W_l = \text{lbm of added limestone per lbm of fuel}
W_w = \text{lbm of water in fuel per lbm of fuel}
\Delta x = \text{annular bed width, m}
x = \text{percent excess air}
X_{CO} = \text{mole fraction of CO in exhaust, dry basis}
X_{CO_2} = \text{mole fraction of CO}_2 \text{ in exhaust, dry basis}
X_{O_2} = \text{mole fraction of O}_2 \text{ in exhaust, dry basis}
X_{N_2} = \text{mole fraction of N}_2 \text{ in exhaust, dry basis}
X_{SO_2} = \text{mole fraction of SO}_2 \text{ in exhaust, dry basis}
Y_c = \text{mass fraction of carbon in flyash}
Y_{Cf} = \text{mass fraction of fixed carbon in coal}
Y_{Ct} = \text{mass fraction of total carbon in coal}
$Y_H$ = mass fraction of hydrogen in coal
$\epsilon_c$ = combustion bed emissivity
$\epsilon_b$ = bed voidage due to bubbles
$\epsilon_{b_{\text{max}}}$ = maximum bubble voidage
$\epsilon_{i_w}$ = dividing wall emissivity
$\epsilon_{m_f}$ = bed voidage at minimum fluidization
$\epsilon_o$ = defluidized bed voidage
$\epsilon_{ow}$ = water jacket wall emissivity
$\epsilon_a$ = annular bed emissivity
$\eta$ = carbon conversion efficiency
$\mu_g$ = gas absolute viscosity, $Ns/m^2$
$\mu_{w}$ = water absolute viscosity, $Ns/m^2$
$\rho_c$ = char particle density, $kg/m^3$
$\rho_g$ = gas density, $kg/m^3$
$\rho_{m_f}$ = bed density at minimum fluidization, $kg/m^3$
$\rho_o$ = inlet air density, $kg/m^3$
$\rho_s$ = sand particle density, $kg/m^3$
$\rho_v$ = saturated water vapor density, $kg/m^3$
$\rho_w$ = water density, $kg/m^3$
$\sigma$ = Stefan-Boltzmann constant, $5.67 \times 10^{-8} W/m^2K^4$
$\sigma_b$ = bubble surface tension, $N/m$
$[i]$ = concentration of species $i$, $mole/cm^3$
1 INTRODUCTION

Fluidized bed combustion (FBC) has grown with the prospect that it can burn coal and low grade fuels in an environmentally acceptable manner. In 1985, more FBC units were sold than conventional boilers, the first time this has ever happened [1]. The reason for this growth is the ability of fluidized bed combustors to burn a wide variety of low-rank and high-sulfur coals cleanly and efficiently. The United States has a large abundance of these coals [2].

There are many unique characteristics of fluidized bed combustors. The use of a sulfur sorbent as the bed material results in large reductions in the emission of sulfur dioxide. Lower combustion temperatures compared to conventional pulverized coal boilers result in lower emissions of oxides of nitrogen (\(\text{NO}_x\)) produced from atmospheric nitrogen. Fuel nitrogen is therefore the major source of \(\text{NO}_x\) formation in FBC [3]. Reductions in \(\text{NO}_x\) emissions can be realized by feeding larger coal particles, reducing excess air levels, and utilizing staged combustion [4].

The relatively low combustion temperature also promotes optimum sulfur capture efficiency, relaxes the need for exotic boiler construction materials, and eliminates problems with ash slagging. The bed temperature is also spatially uniform; there are no hot spots that could lead to ash sintering.

The ability to burn a wide variety of fuels (high-sulfur or high-ash coals, biomass
fuels, sludge, etc.) is a result of the good gas-solids contact, long fuel residence times, and vigorous mixing found in fluidized beds. The good mixing in the bed leads to high combustion efficiencies. Fluidized beds have high rates of convection heat transfer to surfaces immersed in the bed. Convection coefficients for tubes located in the bed can be more than an order of magnitude than for tubes located in the freeboard region [5]. These high coefficients allow bed heat exchangers to be very compact.

Despite these advantages, technical problems remain that must be overcome before wider markets are developed. Prominent among these difficulties is the poor load turndown capability of fluidized bed combustors. Load turndown is the ability to vary the firing rate of a combustor to match system energy demands. Maximum load turndown ratio is defined as the ratio of maximum to minimum fuel firing rates. Inherent to conventional FBC designs is an inability to produce large variations in heat transfer rate from the fluidized bed. Generally, changes in heat transfer rate are modest and are accompanied by degradation in combustion. Innovative concepts in bed design are required to control heat transfer independent of combustion. This capability is especially important for fluidized beds targeted for coal-fired gas turbine power systems and small-scale boilers and furnaces.

The objective of this research was to investigate a new concept in fluidized bed design that improves load turndown capability. The research consisted of two major phases of work:

1. Development of computational models to predict heat transfer and combustion performance and to aid in the design of a fluidized bed combustor. The heat transfer model includes energy and mass balances on the two-bed com-
bustor and a semi-empirical theory for calculating convection coefficients. The combustion model includes mass balances for reacting species and a simple kinetic model for coal combustion.

2. Construction of the combustor and performance of coal combustion tests. The combustion tests were performed with three different coal-based fuel forms: crushed coal, coal-limestone briquettes, and coal-water-limestone mixture (CWLM).

The research goal was to determine if a load turndown ratio of 10:1 could be obtained for a fluidized bed combustor, while satisfying pollution emission criteria.
2 BACKGROUND

Heat transfer from a fluidized bed to water tubes is determined by three factors:

1. The temperature gradient between bed and water.

2. The heat transfer area.

3. The overall heat transfer coefficient between bed and water.

Boiler application usually sets the water-side temperature; attempts to control load with temperature gradients require large variations in bed temperature. However, even small variations in bed temperature from optimum design values will greatly degrade both sorbent utilization [6] and combustion efficiency [7].

Reduction of heat transfer area has been suggested as a method for reducing loads in FBC. This condition can be accomplished by either reducing fluidization velocity, which contracts bed volume, or by discharging bed material. The former approach is of little practical value because bed contraction is limited to about 30% [8]; the corresponding load turndown is modest at best. Discharging, storing, and reinjecting hot particles is fraught with many technical difficulties and has little to recommend as a method for load turndown. In addition to the above difficulties, both methods for reducing heat transfer area will expose tubes to erosion when they are in the splash zone of the bed. Another method for reducing heat transfer
area requires the air distributor to be partitioned which allows zones of the bed to be independently fluidized. Load turndown is achieved by selectively slumping part of the bed; heat transfer area in defluidized zones is effectively zero. This technique has some undesirable effects on combustion including fuel smoldering and agglomeration in the slumped regions. Although bed slumping is frequently employed in commercial FBC units, turndown capability is rather modest.

Variation of the overall heat transfer coefficient between bed and tubes can also be employed for load turndown control. Heat transfer coefficients in fluidized beds show large variation with fluidization velocity; in principle, turndown ratios exceeding ten can be achieved by reducing fluidization velocity from its maximum heat transfer value to the minimum fluidization condition. However, as Figure 2.1 illustrates, the dependence of heat transfer coefficient on fluidization velocity is strongly nonlinear; since combustion rate is proportional to fluidization velocity, a match between heat release and heat transfer rates is difficult to achieve. Horio et al., 1985 [9], have developed a baffled heat transfer tube with the goal of achieving a linear response in average heat transfer coefficient with changes in fluidization velocity. Although they were successful in obtaining a linear response in the velocity range of 0.3 to 0.5 m/s, this achievement represents only a modest turndown ratio. It is far from evident that a sufficiently linear response can be achieved over larger velocity intervals. In addition, the baffle arrangements produce a linear response at great sacrifice in heat transfer rates.

A more promising approach to improved load turndown is control of heat transfer rate independent of combustion rate [10,11]. The device as described here can be employed in fluidized beds that remove thermal energy around the perimeter of the
Figure 2.1: Heat transfer rate and combustion rate vs. fluidization velocity
bed, i.e., water jacket or water wall construction; however, the principle can also be applied to any vertical water tube design. Independent control of heat transfer rate and combustion rate is accomplished by surrounding the fluidized bed in which fuel is burned, hereafter called the combustion bed, by another fluidized bed, hereafter called the heat transfer bed, that establishes the overall heat transfer rate from the inner combustion bed. The two beds, physically divided by a wall, are fluidized independently by separate air plenums.

Figure 2.2 illustrates the device in a water jacketed, cylindrical fluidized bed. The central combustion bed is provided with fluidization air through a circular distributor plate from an air plenum which is designed to give even distribution of air through the bed. Coal or other fuel is fed into the combustion bed at a rate determined by the desired heat generation rate, while air flow into this bed is set at a rate consistent with efficient combustion. The heat transfer bed is the annular fluidized bed surrounding the combustion bed. The two beds are separated by a wall constructed of heat resistant material of reasonably high thermal conductivity such as stainless steel. The heat transfer bed is supplied with fluidization air from a plenum separate from the combustion bed plenum. The heat transfer bed is enclosed by a water jacket that removes energy from the combustor in the form of hot water or steam. Overall heat transfer rate from the combustion bed to the water jacket is determined by the heat transfer coefficients associated with the fluidized combustion bed, the conductivity of the wall separating the beds, the heat transfer coefficients at the inner and outer diameters of the heat transfer bed, the conductivity of the water jacket wall, and the convection coefficient of the water in the water jacket. However, control of the overall heat transfer rate will reside in the heat transfer bed and will
Figure 2.2: Two-bed fluidized combustor
be accomplished by changing the fluidization velocity of air entering this annular bed. The combustion bed can be operated in Region B of Figure 2.1 where heat transfer rate is only a weak function of fluidization velocity; air flow rate to this bed can be chosen consistent with good combustion and independent of heat transfer considerations. The heat transfer bed can be operated in Region A where large variations in heat transfer rate can be achieved. If no air is passed through the heat transfer bed, then it has the poor heat transfer characteristics of packed granular beds. If only sufficient air is passed through the heat transfer bed to just fluidize it, then increased heat transfer due to convection occurs. Heat transfer continues to increase with increasing air flow until enhanced heat transfer characteristic of bubbling fluidized beds is reached; the result is a continuous and large variation in heat transfer rate from the combustor that is controlled independently of combustion rate. A simple analysis provides an estimate of the load turndown capability of this device.

Let $q$ be the heat transfer per unit wall area of the combustor. The load turndown that can be achieved is approximately the ratio of $q$ for full fluidization of the heat transfer bed to $q$ for the slumped heat transfer bed. For steady state operation of the combustor, the maximum heat transfer from the combustion bed can be approximated by:

$$q_{max} = \left( \frac{1}{h_c} + \frac{2}{h_a} + \frac{1}{h_w} \right)^{-1} (T_c - T_w)$$

(2.1)

In deriving this equation it is assumed that the combustion and heat transfer beds are uniform in temperature due to vigorous mixing. In addition, the heat transfer bed is assumed to be deep compared to its radial dimension; hence, heat loss associated with energy convected out of the heat transfer bed with fluidization air
is relatively small. In the case of minimum heat transfer from the combustor, the heat transfer bed is completely defluidized:

$$q_{min} = \left( \frac{1}{h_c} + \frac{\Delta x}{k} + \frac{1}{h_w} \right)^{-1} (T_c - T_w)$$  \hspace{1cm} (2.2)

Further simplification is obtained if it is assumed that, for the maximum heat transfer condition, the beds are equally fluidized and employ identical bed material. If boiling heat transfer is employed in the water jacket, then:

$$h_a = h_c \ll h_w$$  \hspace{1cm} (2.3)

For the minimum fluidization condition:

$$\frac{k_c^0}{\Delta x} \ll h_c \ll h_w$$  \hspace{1cm} (2.4)

Hence, the turndown ratio can be approximated by the expression:

$$\frac{q_{max}}{q_{min}} = \frac{1}{3} \frac{h_c \Delta x}{k_c^0}$$  \hspace{1cm} (2.5)

For a 2.5 cm wide heat transfer bed of sand with a bulk conductivity of 0.24 W/mK and a typical fluidized bed heat transfer coefficient of 250 W/m²K, a turndown ratio of about 8 is predicted. The energy convected out of the heat transfer bed with fluidization air does not represent heat loss from the combustor; it can be recovered by heat exchange, employed in preheating fluidization air for the combustion bed, or used in staged combustion in the freeboard [12]. Staged combustion was incorporated in the combustor constructed for this study.
3 HEAT TRANSFER MODEL

The heat transfer model simulates the energy transfer processes in the two-bed fluidized combustor. The model is used to predict combustion bed temperature, combustor heat transfer rates, and load turndown for varying operating conditions. These operating conditions include annular bed width, annular bed particle sizes, annular bed flow velocity, and fuel firing rate.

Several assumptions were employed in model development:

• The central combustion bed is at a uniform temperature, $T_c$, due to vigorous mixing and large thermal mass.

• The annular bed is also isothermal, at $T_a$.

• Steady state conditions exist in the combustor.

• The energy released in the bed, $Q_{in}$, is released uniformly to the central bed.

• Boiling occurs at the water jacket wall with a temperature equal to the saturation temperature of the water.

• Both fluidized beds are opaque and the only radiation effect is from the beds to the wall surfaces. Radiation from the upper bed surface to the freeboard walls can be neglected if the freeboard is assumed to be insulated.
• Heat conduction occurs normal to wall surfaces.

• The dividing wall (between inner and annular beds) and water jacket wall are at uniform temperatures, \( T_{iw} \) and \( T_{ow} \), respectively.

• Heat transfer occurs from the annular bed to the fluidization air and to the water jacket.

• Air properties are used for exhaust gases exiting the combustion bed.

3.1 Energy Balance Equations

The mathematical formulation of the model consists of a system of nonlinear algebraic equations describing heat transfer in the combustor due to convection, radiation, and boiling. These equations, derived from energy balances on the combustion bed, dividing wall, heat transfer bed, and water jacket wall, are:

**Combustion bed energy balance**

\[
Q_{in} = \rho u c p c A_c (T_c - T_o) + \pi D_c h_j (h_c + h_{ciw})(T_c - T_{iw})
\]  \(3.1\)

**Dividing wall energy balance**

\[
(h_c + h_{ciw})(T_c - T_{iw}) - (h_a + h_{aiw})(T_a - T_{iw}) = 0
\]  \(3.2\)

**Annular bed energy balance**

\[
\frac{\rho u a c p a A_a(T_a - T_o)}{\pi H_j} + D_c (h_a + h_{aiw})(T_a - T_{iw}) - D_a (h_a + h_{aow})(T_a - T_{ow}) = 0
\]  \(3.3\)

**Water jacket wall energy balance**

\[
(h_a + h_{aow})(T_a - T_{ow}) = q_b
\]  \(3.4\)
Boiling water heat flux

Nucleate pool boiling occurs in the water jacket and is modelled using a correlation developed by Rohsenow and cited by Incropera and DeWitt [13]:

\[ q_b = \mu_w h_f g \left( \frac{g(\rho_w - \rho_v)}{\sigma_b} \right)^{\frac{1}{2}} \left( \frac{c_{pw}}{c_s f h_f g P_{rw}} \right)^{3} (T_{ow} - T_{sat})^{3} \] (3.5)

Radiation heat transfer coefficients

The radiation heat transfer coefficients have been linearized and include effects of view factors and emissivities. Here it is assumed that surface areas of wall and bed are equivalent:

\[ h_{ciw} = \frac{\sigma(T_{iw}^2 + T_c^2)(T_{iw} + T_c)}{\left( \frac{1}{\varepsilon_c} + \frac{1}{\varepsilon_{iw}} - 1 \right)} \] (3.6)

\[ h_{aiw} = \frac{\sigma(T_{iw}^2 + T_a^2)(T_{iw} + T_a)}{\left( \frac{1}{\varepsilon_a} + \frac{1}{\varepsilon_{iw}} - 1 \right)} \] (3.7)

\[ h_{aow} = \frac{\sigma(T_{ow}^2 + T_a^2)(T_{ow} + T_{a})}{\left( \frac{1}{\varepsilon_a} + \frac{1}{\varepsilon_{ow}} - 1 \right)} \] (3.8)

The preceding equations apply when the annular bed is fluidized. When the annular bed is defluidized, a second set of equations describing conduction across a static bed of particles is used:

Energy released in central bed

\[ Q_{in} = \pi D_c H_J (h_c + h_{ciw})(T_c - T_{iw}) + \rho_o u_c o_A c_\varepsilon_p c(T_c - T_0) \] (3.9)

Dividing wall energy balance

\[ (h_c + h_{ciw})(T_c - T_{iw}) + \left[ \frac{2k_o}{D_o \ln(D_c/D_a)} \right](T_{ow} - T_{iw}) = 0 \] (3.10)
Water wall energy balance

\[ Q_{\text{cond}} = q_b (\pi D_a H_j) \] (3.11)

Annular bed energy balance

\[ Q_{\text{cond}} = Q_{\text{in}} - \rho_o \left[ u_{oo} \cdot A_c \bar{c}_{pc} (T_c - T_0) + u_{oa} \cdot A_a \bar{c}_{pa} (T_a - T_0) \right] \] (3.12)

3.2 Fluidized Bed Convection Coefficients

Fluidized bed convection coefficients are found theoretically using the two-phase theory of fluidization described by Xavier and Davidson [14]. This theory treats the bed as consisting of a particle-dense phase (emulsion phase) and a particle-lean phase (bubble phase). The bubble phase is created when gas flow in excess of that required for minimum fluidization passes through the bed in the form of voids, or bubbles. The bed convection coefficient is composed of particle convection and gas convection:

\[ h_c = h_{pc} + h_{gc} \] (3.13)

Radiation heat transfer is accounted for separately in the energy balance equations given previously. The particle convection coefficient is calculated from a packet resistance in series with a film resistance:

\[ h_{pc} = \frac{1 - \epsilon_b}{\left( \frac{1}{h_p} - \frac{1}{h_f} \right)} \] (3.14)

The reciprocal film resistance is:

\[ h_f = \frac{m k_g}{d_p} \] (3.15)
The reciprocal packet resistance is:

$$h_p = 2 \left[ \frac{k_{mf} \rho_{mf} c_{mf} (u - u_{mf})}{\pi L} \right]^{0.5}$$ (3.16)

where:

$$L = 0.5 d_b$$ (3.17)

and the product of density and specific heat of the emulsion phase is:

$$\rho_{mf} c_{mf} = \rho_s (1 - \epsilon_{mf}) c_s$$ (3.18)

The height-average bubble diameter may be calculated from:

$$d_b = 0.54 (u - u_{mf})^{0.4} \left[ H_f + 4 \sqrt{A_o} \right]^{0.8} g^{-0.2}$$ (3.19)

and the bed voidage due to bubbles is:

$$\epsilon_b = 1 - \frac{u_b}{u - u_{mf}}$$ (3.20)

The bubble rise velocity is a function of relative bubble size:

$$u_b = 0.71 \sqrt{gd_b} \quad \text{if} \quad \frac{d_b}{D} < .125$$ (3.21)

$$u_b = 0.802 \sqrt{gd_b} \exp \left(-\frac{d_b}{D}\right) \quad \text{if} \quad .125 \leq \frac{d_b}{D} \leq .60$$ (3.22)

$$u_b = 0.35 \sqrt{gd_b} \quad \text{if} \quad \frac{d_b}{D} > .60$$ (3.23)

The thermal conductivity of the emulsion phase is found from:

$$k_{mf} = k_e^0 + 0.1 \rho_g c_g d_p u_{mf}$$ (3.24)

where:

$$k_e^0 = k_g \left[ \left( \frac{k_s}{k_g} \right) \left( 0.28 - 0.757 \log \epsilon_o - 0.057 \log(k_s/k_g) \right) \right]$$ (3.25)
The minimum fluidization velocity is:

$$u_{mf} = \frac{25.7 \mu g}{\rho g d_p} \left[ (1 + 5.53 \times 10^{-5} Ar)^{0.5} - 1 \right]$$

where the dimensionless Archimedes number is given by:

$$Ar = \frac{\rho g (\rho_s - \rho_g) g d_p^2}{\mu_g^2}$$

The gas convection coefficient consists of a bed resistance in series with a wall resistance:

$$h_{gc} = \left( \frac{1}{h_{av}} + \frac{1}{h_w} \right)^{-1}$$

where the reciprocal bed resistance is:

$$h_{av} = \left( \frac{4 k_{mf} \rho g c g u}{\pi H_f} \right)^{0.5}$$

and the wall resistance is approximately equal to the film resistance previously calculated:

$$\frac{1}{h_w} = \frac{1}{h_f}$$

One additional condition is used in calculating heat transfer coefficients in the annular bed. Since the annular bed is constrained on the top surface by a screen to prevent particle elutriation, there is a maximum height, $H_{max}$, to which the bed may expand. Therefore, once the bed expands to $H_{max}$, the following condition restricting bubble voidage applies:

$$\epsilon_{b \text{max}} = 1 - \frac{H_{mf}}{H_{min}}$$
3.3 FORTRAN Program

A listing of the FORTRAN program that was written to solve the preceding set of energy balance and convection coefficient equations is shown in Appendix A. The main program solves the energy balance equations, Eqs. 3.1 - 3.12, iteratively. The main program utilizes the following subroutines:

- PROP - evaluates properties of air
- MINFLU - evaluates minimum fluidization velocity, fluidized bed conductivity, and gas convection coefficient
- COEFFA - evaluates convection coefficient for annular heat transfer bed (either a semi-empirical model based on the two-phase theory of fluidization or experimental heat transfer data is used - See Chapter 7.3)
- COEFFC - evaluates convection coefficient for combustion bed using the two-phase theory model

3.4 Heat Transfer Model Predictions

The heat transfer model was used to aid in preliminary design of the two-bed fluidized combustor. Specifically, it was used to determine the effect of annular bed width and annular bed particle size on load turndown. The load turndown is obtained as the ratio of maximum to minimum firing rates in the combustor at constant combustion temperature. The minimum firing rate is obtained at the point where the combustion bed is minimally fluidized and the annular heat transfer bed is static (defluidized). At this condition, the combustion bed will be operated in
a fuel-lean manner, and it is reasonable to assume that 100% of the fuel energy is released in the central bed.

The maximum firing rate is obtained by setting the annular bed air flow rate at the point of maximum convective heat transfer coefficient. The combustion bed air flow rate is set at an expected maximum allowable level. For computational purposes, this maximum superficial velocity is assumed to be $6u_{mf}$, above which significant bed material would be elutriated. Since the combustion bed normally operates sub-stoichiometrically, it is assumed that only 50% of the heat of combustion ($Q_{comb}$) is released in the bed. The energy released in the bed ($Q_{in}$) was determined for the maximum and minimum heat transfer cases; the load turndown, then, is calculated as the ratio of maximum $Q_{comb}$ to the minimum $Q_{comb}$.

Particle size has a large effect on convection coefficients in fluidized beds; a decrease in particle size by a factor of two can double convection coefficients. It was therefore assumed that decreasing particle size would have a large positive effect on load turndown. However, simulations suggested that use of small particles would not have as dramatic an effect on load turndown as might be expected. Figure 3.1 demonstrates that reducing annular bed particle size from 1.0 mm to 0.3 mm increases load turndown from 8.3 to only 9.8 in the case of 1.0 mm particles in the combustion bed, even though convection coefficients in the annular bed increase by a factor of two. The reason for the relatively small effect is as follows. As the air flow through the annular bed becomes large, heat transfer from the combustion bed becomes rate-limiting to the overall heat transfer process. Consequently, the overall heat transfer rate between the combustion bed and the annular bed does not increase dramatically for this range of particle sizes. Nevertheless, there is an
Annular Bed Width = 25.4 mm; \( T_c = 1144 \, K \)

Combustion Bed Particle Size

- □ □ □ 1000 \( \mu m \)
- ▲ ▲ 500 \( \mu m \)
- ○ ○ ○ 250 \( \mu m \)

Figure 3.1: Theoretical effect of particle size on load turndown
advantage in using small particles in the annular bed. In addition to the slightly higher load turndown, smaller particles also require a lower air flow rate. Sand sieved to 50 x 70 mesh (0.25 \( mm \) average diameter) was selected as the smallest practical bed material that could be employed in the annular bed. Smaller particles could be too easily entrained in the gas flow by the action of gas bubbles bursting at the surface of the annular bed.

It can also be seen from Figure 3.1 that higher load turndown is achieved with small combustion bed particles. A load turndown of 15 is predicted for 0.25 \( mm \) combustion bed particles. However, particles smaller than 1.0 \( mm \) are not practical in the combustion bed because of excessively small particle terminal velocity; low air flow rates result in low firing rates. Therefore, 1.0 \( mm \) sand particles were selected for use in the combustion bed to allow reasonably high air velocities, and hence, high fuel firing rates.

Simulations also clearly indicated the advantage of a wide annular bed for producing large load turndown. Results shown in Figure 3.2 predict load turndown exceeding 15 with an annular bed width of 65 \( mm \). However, the overall diameter of the combustor to be used in experiments was constrained to 254 \( mm \) by the fixed dimensions of the water jacket prepared for experiments to be performed. Under this constraint, a large annular bed width would have produced an unreasonably small combustion bed volume. Therefore, a 25.4 \( mm \) width was selected as large enough to achieve a desired load turndown of 10.
Figure 3.2: Theoretical effect of bed width on load turndown

$T_c = 1144 \, K$

Annular Bed Particle Size $= 250 \, \mu m$
In summary, the following design parameters were chosen for the two-bed fluidized combustor:

- **Combustor bed diameter**: 203 mm
- **Annular bed width**: 25.4 mm
- **Combustion bed average particle diameter**: 1.00 mm
- **Annular bed average particle diameter**: 0.25 mm
4 COMBUSTION MODEL

One of the inputs to the heat transfer model is the fraction of fuel burned in the combustion bed. Values ranging from 50% to 100% were assumed in predictions of load turndown in the previous chapter. In order to predict fuel flow rates (and load turndown) more precisely, it is necessary to model the combustion process in the combustion bed. The combustion model consists of species conservation equations for carbon monoxide and oxygen that employ a simple kinetic mechanism for coal combustion.

When coal particles are fed into the combustion bed, they undergo rapid heating and volatile gases are released. The remaining char particles consist mainly of fixed carbon and ash. To model the combustion process, it is necessary to consider the behavior of both volatiles and char.

The mechanism of volatile combustion in fluidized beds is not fully understood [15]. Most models assume that the volatiles are released instantaneously upon fuel entrance to the bed, which is a reasonable assumption since volatiles release usually occurs within seconds upon rapid heating [16].

The rate of char combustion is usually limited by one of the following rates [15]:

1. Film diffusion - diffusion rate of oxygen from the emulsion phase to the char
particle surface

2. Pore diffusion - diffusion rate of oxygen through pores in the char particle

3. Chemical reaction - adsorption and reaction of oxygen at active sites in the char particle to produce gaseous reaction products

The char particle may burn as a constant volume sphere or as a shrinking sphere. In the case of constant diameter, the particle either completely burns up or it reaches a critical porosity, at which time the particle fragments and elutriates from the bed [17]. A particle which burns as a shrinking sphere simply shrinks in diameter as it burns until it is small enough to be elutriated from the bed.

There is evidence that char combustion in fluidized beds is film diffusion rate limited [16]. Pore diffusion and chemical reactions take place rapidly compared to the diffusion of oxygen to the char particle surface. There is also strong evidence that char particles burn with constant diameter [17], and the model that follows assumes constant diameter particles. The following assumptions have been employed in the combustion model development:

- Volatiles are released instantaneously from coal particles entering the bed.
- Volatiles are converted instantaneously to carbon monoxide and water in the emulsion phase.
- The char particle burns with constant diameter where reaction is rate limited by diffusion of oxygen to the particle surface.
- Unburned carbon losses arise only from elutriation of char particles from the bed (no loss from bed overflow).
• The char reacts to form carbon monoxide (CO); the CO does not further react in the particle boundary layer or emulsion phase [18,19].

• Carbon monoxide produced from volatile and char combustion diffuses to the bubble phase where it is converted to carbon dioxide by homogeneous reaction.

• The emulsion phase is well-mixed; gas concentrations are spatially uniform.

• The bubble phase is in plug flow; concentrations of CO and O₂ vary with height in the bubble phase.

4.1 Mathematical Formulation

4.1.1 Emulsion phase model

In the emulsion phase, char reacts with oxygen to form carbon monoxide and the volatiles react to form carbon monoxide and water. Since it is assumed that the carbon portion of the volatiles is converted to carbon monoxide instantaneously, the rate of carbon monoxide production due to volatile combustion is:

$$ \dot{G}_{CO_{vol}} = (Y_{C_t} - Y_{C_f}) \left( \frac{\dot{m}_{de}}{M_{C_t}} \right) $$

(4.1)

where \( Y_{C_t} - Y_{C_f} \) (the difference between total and fixed coal carbon content) is the carbon content of volatiles.

The char reaction is given by:

$$ C(s) + \frac{1}{2}O_2 \rightarrow CO $$

(4.2)
The reaction is rate limited by the diffusion of oxygen to the char surface [16]:

\[ n = 2\pi d Sh D g C\cdot O_{2e} \]  \hfill (4.3)

where \( n \) is the rate of oxygen diffusion to an individual char particle, \( D_g \) is the molecular diffusivity of oxygen, \( d \) is the char particle diameter, and \( C\cdot O_{2e} \) is the concentration of oxygen in the emulsion phase. The Sherwood number, \( Sh \), a measure of the mass transfer at the particle surface, was found by Avedesian and Davidson to have an average value of 1.42 for fluidized beds [16]. A constant value of molecular diffusivity, \( D_g = 2.08 \times 10^{-4} \), was assumed. The rate of oxygen consumption in the emulsion phase due to char combustion is then the rate of diffusion per particle multiplied by the number of char particles in the bed:

\[ \dot{G}_{O_2\text{char}} = -nN \]  \hfill (4.4)

The number of char particles, \( N \), is found from the bed carbon loading divided by the mass per char particle:

\[ N = \frac{m_c}{\frac{\pi}{6} \rho_c d^3} \]  \hfill (4.5)

where \( \rho_c \) is the char particle density, in kg/m\(^3\), and \( m_c \) is the bed carbon loading, in kg. In principle, carbon loading can be calculated from a population balance on burning particles [8], however, Gregory and Brown [20] found the bed carbon loading to be fairly constant at 0.080 kg for combustor geometry and operating conditions similar to those used in the present experiments. This value is used for these simulations.
Derivation of species conservation equations for both the emulsion phase and bubble phase appear in Appendix B. The resulting equations for the emulsion phase are:

\[
\frac{1}{H_c} \left[ u_m f C_{O_{2\epsilon}} - u_m f \left( \frac{T_0}{T_c} \right) C_{O_{2o}} \right] = \frac{-k_q a_b \epsilon_b \left( C_{O_{2\epsilon}} - \bar{C}_{O_{2\epsilon}} \right)}{(\epsilon_m f (1 - \epsilon_b))} \left( \frac{\dot{G}_{O_2}}{H_c A_c} \right)
\]  
(4.6)

\[
\frac{1}{H_c} \left[ u_m f C_{CO_{\epsilon}} - u_m f \left( \frac{T_0}{T_c} \right) C_{CO_o} \right] = \frac{-k_q a_b \epsilon_b \left( C_{CO_{\epsilon}} - \bar{C}_{CO_{\epsilon}} \right)}{(\epsilon_m f (1 - \epsilon_b))} \left( \frac{\dot{G}_{CO}}{H_c A_c} \right)
\]  
(4.7)

where \( C_{O_{2o}} = 0.008705 \text{ moles/m}^3 \) and \( C_{CO_o} = 0 \) correspond to concentrations of oxygen and carbon monoxide in air, respectively.

Carbon monoxide is produced in the emulsion phase by both char and volatile combustion. The rate of production due to char combustion is found from the rate of diffusion of oxygen to the char surface (Eq. 4.4). Two moles of CO are produced per mole of oxygen diffusing to the char surface, so the rate of CO production is:

\[
\dot{G}_{CO_{char}} = -2 \dot{G}_{O_{2char}} = 2n N = 4\pi d N S h D g C_{O_2\epsilon}
\]  
(4.8)

The rate of carbon monoxide production due to char has a maximum value at steady state fixed by the feed rate of coal:

\[
\dot{G}_{CO_{char max}} = Y_C f \left( \frac{\dot{m}_{dc}}{M_C} \right)
\]  
(4.9)

where \( Y_C f \) is the mass fraction of fixed carbon of the moisture-free coal, \( M_C \) is the molecular weight of carbon, and \( \dot{m}_{dc} \) is the mass flow rate of dry coal. The
difference between maximum and actual carbon monoxide production due to char represents unburned carbon that is elutriated from the combustion bed.

The rate of oxygen consumption in the emulsion phase is given by:

\[ \dot{G}_{O_2e} = \dot{G}_{O_2char} + \dot{G}_{O_2vol} \]  

(4.10)

The rate of oxygen consumption due to volatile combustion is:

\[ \dot{G}_{O_2vol} = -\frac{1}{2} \dot{G}_{C'0_vol} - \frac{1}{2} \dot{G}_{H_2O_vol} \]  

(4.11)

where the production of water due to hydrogen combustion is:

\[ \dot{G}_{H_2O_vol} = Y_H \left( \frac{\dot{m}_{de}}{M_{H_2O}} \right) \]  

(4.12)

where \( Y_H \) is the mass fraction of hydrogen in dry coal, and \( M_{H_2O} \) is the molecular weight of water.

4.1.2 Bubble phase model

In the bubble phase, carbon monoxide produced in the emulsion phase is oxidized to carbon dioxide. The elementary reaction is thought to be [21]:

\[ CO + OH \xrightarrow{k_1} CO_2 + H \]  

(4.13)

Accordingly, the irreversible rate of disappearance of CO is:

\[ \frac{d[CO]}{dt} = -k_1[CO][OH] \]  

(4.14)

The concentration of hydroxyl is assumed to be controlled by the three equilibrium reactions:

\[ O + H_2 = OH + H \]  

(4.15)
\[ \text{OH} + H_2 = H_2O + H \quad (4.16) \]
\[ O + O + M = O_2 + M \quad (4.17) \]

where the equilibrium constants are [22]:

\[ K_a = \frac{[\text{OH}][H]}{[O][H_2]} = 2.239 \exp\left(\frac{-941.1}{T_c}\right) \quad (4.18) \]
\[ K_b = \frac{H_2O[H]}{H_2[OH]} = 0.2089 \exp\left(\frac{7654.7}{T_c}\right) \quad (4.19) \]
\[ K_c = \frac{[O_2]}{[O]^2} = 0.05018 \exp\left(\frac{59561}{T_c}\right) \quad (4.20) \]

Algebraic manipulation of the three above equations leads to an expression for OH concentration:

\[ [\text{OH}] = \left[ \frac{K_a[H_2O][O_2]^{1/2}}{K_bK_c^{1/2}} \right]^{1/2} \quad (4.21) \]

which, when substituted into Eq. 4.14 gives:

\[ \frac{d[C,O]}{dt} = -\frac{k_1K_a^{1/2}}{K_b^{1/2}K_c^{1/4}} [C,O][H_2O]^{1/2}[O_2]^{1/4} \quad (4.22) \]

Howard et al. [21] found the empirical rate constant, \( k_1 \), to be approximately equal to \( 1.3 \times 10^{14} \, \text{cm}^3/\text{mole sec} \).

The species conservation equations for the bubble phase (derived in Appendix B) are:

\[ \frac{d}{dz} C_{O_2b} = \frac{-k_q a_b \epsilon_b \left(C_{O_2b} - C_{O_2e} \right) + \frac{\dot{G}_{O_2b}}{H_c A_c}}{\epsilon_b \left(u - u_{mf}\right)} \quad (4.23) \]

\[ \frac{d}{dz} C_{C'O_b} = \frac{-k_q a_b \epsilon_b \left(C_{C'O_b} - C_{C'O_e} \right) + \frac{\dot{G}_{C'O_b}}{H_c A_c}}{\epsilon_b \left(u - u_{mf}\right)} \quad (4.24) \]
where the subscript e designates emulsion phase and b designates bubble phase. The initial conditions are \( C_{O_{2b}} = C_{O_{2e}} \) and \( C_{CO_b} = 0 \) at \( z = 0 \). The interphase mass transfer coefficient, \( k_q \), is given by Grace [23] as:

\[
k_q = \frac{u_m f}{3} + \left( \frac{4 D g e_m f u_b}{\pi d_b} \right)^{1/2}
\] (4.25)

The rate of carbon monoxide consumption in the bubble phase is:

\[
\dot{G}_{CO_b} = \frac{d[C O]}{dt} (e_b H_c A_c)
\] (4.26)

Substitution of Eq. 4.22 into the above equation with proper unit conversion yields:

\[
\dot{G}_{CO_b} = -\left( \frac{3.162 \times 10^{-5} k_1 K_a^{1/2}}{K_b^{1/2} K_c^{1/4}} \right) (C_{CO_b})^{1/2} (C_{H_2O_b})^{1/4} (C_{O_{2b}})^{1/4} (e_b H_c A_c)
\] (4.27)

The rate of oxygen consumption in the bubble phase is then simply:

\[
\dot{G}_{O_{2b}} = \frac{1}{2} \dot{G}_{CO_b}
\] (4.28)

### 4.1.3 Fraction of fuel burned in-bed

With knowledge of the emulsion phase CO concentration and the bubble phase average CO concentration, the fraction of fuel energy released in the bed may be determined. Incomplete combustion of the fuel in the bed has three components:

1. Unburned carbon in char particles
2. Unreacted carbon monoxide advected from emulsion phase
3. Unreacted carbon monoxide advected from bubble phase
These components may be treated as energy losses; the loss due to unburned char in the emulsion phase is:

\[ L_1 = \dot{V}_C(\Delta H_{fC}) \]  \hspace{1cm} (4.29)

where \( \Delta H_{fC} = 393520 \text{ kJ/mole} \) is the enthalpy of reaction of \( C(s) + O_2 \rightarrow CO \).

The molar flow rate of unburned carbon, \( \dot{V}_C \), is found as the difference between maximum CO production and actual CO production due to char combustion (Eq. 4.9 - Eq. 4.8); the previous equation becomes:

\[ L_1 = \left( \frac{Y_C f \dot{m}_{dc}}{M_C} - 4\pi dNShDo_{2e} \right) \Delta H_{fC} \]  \hspace{1cm} (4.30)

The loss due to unreacted carbon monoxide in the emulsion phase is:

\[ L_2 = [C'CO_{\epsilon} u_m f_A e_b (1 - e_m f)] \Delta H_{fCO} \]  \hspace{1cm} (4.31)

where \( \Delta H_{fCO} = 110530 \text{ kJ/mole} \) is the enthalpy of reaction of \( CO + \frac{1}{2}O_2 \rightarrow CO_2 \), and the term in brackets is the molar flow rate of unreacted emulsion phase CO.

Likewise, the loss due to unreacted CO in the bubble phase is:

\[ L_3 = [\bar{C}CO_{b}(u - u_m f)A e_b] \Delta H_{fCO} \]  \hspace{1cm} (4.32)

The percent of fuel burned in-bed is then given by:

\[ \% \text{ burned in bed} = \frac{(100)Q_{in}}{\dot{m}_{dc}(LHV)} = 100 \left[ 1 - \frac{L_1 + L_2 + L_3}{\dot{m}_{dc}(LHV)} \right] \]  \hspace{1cm} (4.33)

where \( LHV \) is the lower heating value of the coal.

### 4.2 FORTRAN Program

A FORTRAN program was written to solve the preceding set of equations and a listing appears in Appendix C. The method of solution is:
1. Input bed parameters such as temperature, voidage, and velocity from the heat transfer model or from experimental data.

2. Guess values of emulsion phase oxygen and carbon monoxide concentrations and fuel mass flow rate.


4. Solve equations 4.6 and 4.7 (in closed form) for emulsion phase oxygen and carbon monoxide concentrations.

5. Solve equation 4.33 for the percent of fuel burned in-bed, and hence, the fuel mass flow rate.

6. Iterate until convergence is reached.

**4.3 Combustion Model Predictions**

Preliminary simulations were performed with the combustion model to determine the effect of fuel flow rate and equivalence ratio on the fraction of fuel burned in the combustion bed. The equivalence ratio is defined as the mass flow rate of coal divided by the mass flow rate of coal required for stoichiometric combustion. The stoichiometric requirement is based on the air flow rate entering the combustion bed. The equivalence ratio is directly proportional to coal feed rate for fixed air flow rate. Figure 4.1 is a plot of predicted fraction of fuel burned in-bed vs. coal feed rate for three different equivalence ratios. Assumptions used in these simulations were constant combustion temperature of 1144 K, combustor geometry
as selected by heat transfer modelling (See Page 22), and a coal heating value of 30000 \(kJ/kg\). The plotted equivalence ratios correspond to 73%, 100%, and 167% of stoichiometric air requirements.

At low fuel flow rates (less than 2.0 \(kg/hr\)) it is predicted that about 80% of the fuel is burned in-bed. At each plotted equivalence ratio, the fuel flow rate reaches a minimum at the point corresponding to minimum bed fluidization. For example, at an equivalence ratio of 1.37, a reduction in fuel flow rate below about 2.0 \(kg/hr\) would defluidize the bed. If the fuel flow rate is further lowered from 2.0 \(kg/hr\), the equivalence ratio will decrease (because air flow rate must be held constant at the minimum fluidization level). In Figure 4.1 if the fuel flow rate is lowered from 2.0 \(kg/hr\) (at an equivalence ratio of 1.37) to 1.0 \(kg/hr\) the equivalence ratio must decrease to 0.60.

Maximum fuel flow rates are limited by the maximum amount of air that can be admitted to the combustion bed, which is set by the bed particle entrainment velocity. For example, in Figure 4.1 at an equivalence ratio of 1.00, the maximum fuel flow rate is about 6.0 \(kg/hr\). A higher fuel flow rate would require increased equivalence ratio, as the air flow rate would be held constant to prevent excessive elutriation. So, if the fuel flow rate is to be increased to 8.0 \(kg/hr\), the equivalence ratio must be increased to 1.37.

At mid-range fuel flow rates (2 - 6 \(kg/hr\)), it is predicted that an increase in equivalence ratio (decrease in air flow rate) will result in an increase in fraction of fuel burned in-bed. One might have expected this trend to be reversed; a decrease in air flow rate decreases oxygen concentration, which decreases combustion rate. This trend suggests that gas residence time (since residence time is proportional to
Figure 4.1: Effect of fuel flow rate and equivalence ratio on fraction of fuel burned in-bed
air flow velocity) is a more important factor in in-bed energy release than is oxygen concentration.

In predicting load turndown using the heat transfer model, it was necessary to assume values for percent of fuel burned in-bed at minimum and maximum fuel flow rates (100% and 50% respectively). Combustion model predictions of 84% and 47% for percent burned in-bed, respectively, support these assumptions.
5 EXPERIMENTAL APPARATUS

The combustion system and test rig consisted of the following main components: the two-bed fluidized combustor, the fuel feed system, air and water metering equipment, the flue-gas analysis system, and a microcomputer-based data acquisition system.

5.1 Two-Bed Fluidized Combustor

A section view of the cylindrical two-bed combustor constructed for coal-fired combustion trials is shown in Figure 5.1. The combustor was constructed in four vertical sections. The bottom section consisted of separate plenum chambers serving the combustion bed and heat transfer bed. Air from these plenums flowed through a common distributor plate. The stainless steel distributor was a 12 mm thick, 406 mm diameter plate with 250 uniformly spaced, 0.238 mm drilled holes. To prevent sand backflow in the annular bed and to prevent flame flashback in the combustion bed, 80-mesh stainless steel screen was welded to the upper surface of the plate. The water jacket section bolted onto the distributor plate and plenum assembly. The third section, a smaller, ceramic-insulated section, was added immediately above the water-jacketed section. A stainless steel cylinder with an inner diameter of 203 mm, a height of 370 mm, and a wall thickness of 3.175 mm ran the length of
Figure 5.1: Experimental two-bed fluidized combustor
the water-jacketed and insulated sections and divided these combined sections into two concentric beds: the central combustion bed and the annular heat transfer bed. The insulated bed section was not an integral part of the two-bed combustor; it was included in this test combustor to simplify energy balances on the combustor and to allow the measurement of overall heat transfer coefficients in the heat transfer bed. The fourth section of the combustor was a 1.2 m long, 203 mm inside diameter, ceramic-insulated freeboard.

The central combustion bed supported combustion and the annular heat transfer bed controlled the amount of heat transfer to the water jacket. Air used to fluidize the annular bed was exhausted through nozzles to the combustor freeboard immediately above the combustion bed. The eight evenly-spaced 6 mm nozzles were installed on the bed dividing wall at a point 300 mm from the distributor plate. They were positioned in such a manner as to impart swirl to the secondary air as it mixed with gas exiting the central bed. This arrangement, in addition to the insulated freeboard, was expected to promote complete combustion of unreacted gases and coal fines released to the freeboard.

5.2 Fuel Feed System

Dry fuel (including crushed coal and coal briquettes) was fed into the combustor using an auger feeder. The above-bed feed point was through the freeboard at a point 460 mm above the distributor plate. The feeder was an Accurate Model 602 variable-speed solids-auger with a 25 mm-diameter feed screw. It was calibrated using a scale-and-stopwatch technique. The feeder was modified to allow positive hopper sealing to prevent exhaust gas backflow through the feeder. (The combustor
operated at slight positive pressure due to the pressure drop in the exhaust lines.)

Liquid fuels (e.g., coal-water mixtures) were fed into the bed using a Teel-model positive displacement pump. The coal-water mixture entered the combustor through a water-jacketed 6.4 mm stainless steel tube. When the coal-water mixture entered the combustion bed, it tended to form pea-size agglomerates. This phenomenon has been found to increase combustion efficiency since the agglomerates have long in-bed residence times. Figure 5.2 is a schematic of the pumping system. A variable speed electric motor that drives the pump was used for feed rate control. All valves were used on a full-open or full-closed basis to prevent clogging and were not used for flow rate control. The mixture was pumped from a 5-gallon reservoir placed on an Arlyn digital scale having a capacity of 50 lbs and a resolution of 0.01 lbs. Since the pump was capable of flow rates greatly in excess of desired fuel feed rates, a bypass piping system was used. A small fraction of the fuel flow was sent to the combustor while the balance was recycled back to the fuel reservoir. Since the mass of fuel in the pipes was essentially constant, the fuel feed rate could be calculated as the change in the scale weight reading divided by the elapsed time between readings.

5.3 Air and Water Metering System

Figure 5.3 shows a schematic of the overall experimental set-up. Three air flows entered the combustor: primary air entered the combustion bed, secondary air entered the heat transfer bed, and tertiary air entered the freeboard of the heat transfer bed. Secondary air and tertiary air, taken together, represented overfire air required to burn volatiles and char that escaped into the freeboard above the
Figure 5.2: Coal-water mixture feed system
Figure 5.3: Experimental apparatus
combustion bed. These air flows could be measured manually using rotameters or automatically with calibrated orifice meters. The orifice meters were assembled using 1-inch NPT pipe and a drilled stainless steel plate. The orifice meters were calibrated using a laminar flow meter. The water flow rate into the water jacket was read manually using a rotameter. Liquified petroleum gas (LPG) was used to preheat the combustor. The LPG entered the central plenum, mixed with the primary air, and was ignited above the bed with a high-voltage spark transformer. The LPG flow rate is read manually with a rotameter.

A photograph of the experimental apparatus is shown in Figure 5.4. There are two combustors in the photograph; the combustor on the left was used in previous work described in Reference 11. The combustor on the right was used in this work. The dry fuel auger and coal-water mixture pumping system are visible on a movable stand between the two combustors. The control panel is on the right side and faces away from the combustors. Not shown in the photograph are the gas analysis system and microcomputer-based data acquisition.

5.4 Flue-Gas Analysis System

Exhaust gas was sampled through a 6 mm stainless steel tube located in the top of the combustor freeboard. Figure 5.5 is a schematic of the gas sampling and analysis system. The exhaust gas flowed through a heat-tape-wrapped stainless steel tube to a particulate filter, an acid-mist filter, and a Perma-Pure dryer. The dryer removed moisture from the flue-gas stream in a the vapor phase, which prevented condensation of water (and the possible formation of sulfuric acid). The gas flowed through a filter before entering the vacuum pump. The gas was then directed to
Figure 5.4: Photograph of the experimental apparatus
Figure 5.5: Gas sampling system
the gas analyzers via a flow-rate-controlled manifold. The gas stream to the NO analyzer first passed through a heated catalytic converter to reduce NO\textsubscript{2} to NO. Five gas analyzers were used: SO\textsubscript{2}, NO, CO\textsubscript{2}, CO and O\textsubscript{2} meters. The oxygen meter was a Beckman Model 755 paramagnetic analyzer. The CO\textsubscript{2} and CO meters were Beckman Model 820 non-dispersive infra-red analyzers. Horiba Model VIA-300 and VIA-500 non-dispersive infra-red analyzers were used to measure NO\textsubscript{x} and SO\textsubscript{2} respectively.

Solids (ash and unburned carbon) sampling was performed using a 75 mm diameter cyclone located downstream of the freeboard as shown in Figure 5.3. The solids sample was collected in a 1-liter glass jar. Cyclone collection efficiency could be periodically tested by performing isokinetic sampling on the exhaust tube. In shakedown combustion trials, it was found that the cyclone collection efficiency was greater than 90\% for a variety of operating conditions.

5.5 Data Acquisition System

A Zenith Model Z-158 microcomputer equipped with a 20-megabyte hard disk drive was used for data acquisition and test control. The computer had five expansion slots for slide-in expansion boards. One expansion slot was occupied by a 24-bit, parallel, digital input/output interface board (Metrabyte Model PIO-12). Two 8-channel, high speed, analog-to-digital converter boards (Metrabyte Model DAS-8) occupied two other expansion slots. Figure 5.6 is a schematic of the data acquisition system.

Thermocouple measurements were performed using 3, 16-channel, analog input, sub-multiplexer boards (Metrabyte Model EXP-16) giving a total capacity of 48
Figure 5.6: Data acquisition system
thermocouples. The EXP-16 boards performed signal amplification, multiplexing, and cold junction compensation, and were connected in parallel to one of the DAS-8 boards. Each EXP-16 occupied one channel of the DAS-8 board.

Thermocouples were used to measure the following temperatures: air and water inlet temperatures, heat transfer and combustion bed temperatures, combustor wall temperatures, and exit temperatures of flue-gas and water. Temperature measurements were made using chromel-alumel (Type K) thermocouples. For surface temperature measurement, the thermocouple wires were individually spot-welded to the metal surfaces to form intrinsic thermocouple junctions. This method of attachment provided a more accurate measurement of the surface temperature compared to attaching a thermocouple bead to the surface. In regions of high temperature, the thermocouple leads were electrically insulated by threading them through ceramic tubes. Thermocouple probes were used for air, water, and bed temperature measurements.

Figure 5.7 is a schematic of the pressure measurement system. The orifice meter pressure drops were measured using a Schaevitz Model P3061 LVDT (linear variable displacement transformer) pressure transducer. The analog electrical output of the transducer was connected to one of the DAS-8 channels. Pressure drops of multiple orifice meters were read from the single transducer using a computer-controlled solenoid-valve manifold system. Pressure taps from each orifice meter were connected to Humphrey Model M31E1 solenoid valves. The solenoids were activated by the computer using an electromechanical relay board (Metrabyte Model ERA-01), connected to the PIO-12 interface board. A given pressure drop could then be measured by activating the corresponding set of solenoid valves.
Figure 5.7: Pressure measurement system
6 EXPERIMENTAL PROCEDURE

6.1 Load Turndown Tests

A series of load turndown tests were planned with the two-bed combustor. These tests were designed to determine the maximum load turndown while operating at constant combustion temperature. Tests were planned for three different coal-based fuel forms: double-screened crushed coal, coal-limestone briquettes, and coal-water-limestone mixture (CWLM). In addition to determining load turndown, emissions of sulfur dioxide and nitrogen oxide were monitored to evaluate the effectiveness of limestone and low bed temperature on pollution control. Mass and energy balances were then performed on each set of test data to determine carbon conversion efficiency, fraction of fuel energy released in-bed, and overall heat transfer coefficients.

6.2 Fuel Preparation

A 3200-pound bulk sample of coal was obtained from the Rapatee mine of the Midland Coal Company in Fulton County, Illinois. The sample consisted of freshly-mined, washed Illinois No. 5 seam coal, 1.5-inch in top-size. The coal was obtained directly from coal load-out facilities at the mine/preparation plant, loaded in a truck, covered with a tarpaulin, and transported to preparation facilities at Iowa.
State University.

Two representative bulk samples were obtained by cone-and-quartering from the 3200-pound bulk sample. One sample (1200 lb) was used for briquette and coal-water mixture formulation and the second sample (2000 lb) was crushed for dry coal tests.

The 1200-pound sample was ground in a Holmes Model 45 impact pulverizer fitted with a 0.0625-inch perforated screen. This pulverizer produced a coal product with a top size of approximately 40 mesh (0.125 mm). This pulverized sample was then divided by riffling into equal portions and stored in plastic bags, under argon, in lined 55-gallon steel drums. An analytical sample was also obtained from the pulverized coal by splitting and grinding to minus 60 mesh (0.250 mm). Ultimate and proximate analyses were performed on this sample and the results appear in Table 6.1.

The 2000-pound representative bulk sample was sieved to produce a particle size range of approximately 10.0 mm by 1.0 mm. This size range was selected so that the largest particles could be accurately metered into the combustor with an auger feeder and that the smallest particles were large enough not to elutriate from the combustor. After sieving, oversize material was reduced in a Brawn 3-inch jaw crusher to obtain the largest quantity of coal in the desired size range while producing a minimum of fines. Approximately 1200 pounds of 10.0 mm by 1.0 mm coal resulted from this operation. An analytical sample of this screened coal was also prepared to determine if loss of fines changed the coal analysis. The analysis was performed and the results were not significantly different than those shown in Table 6.1 for the pulverized coal sample.
Table 6.1: Illinois No. 5 coal analysis

<table>
<thead>
<tr>
<th>Proximate analysis (%)</th>
<th>As received</th>
<th>Moisture-free</th>
</tr>
</thead>
<tbody>
<tr>
<td>Moisture</td>
<td>10.8</td>
<td>0.0</td>
</tr>
<tr>
<td>Ash</td>
<td>10.1</td>
<td>11.3</td>
</tr>
<tr>
<td>Volatile matter</td>
<td>36.6</td>
<td>41.0</td>
</tr>
<tr>
<td>Fixed carbon</td>
<td>42.5</td>
<td>47.7</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Ultimate analysis (%)</th>
<th>As received</th>
<th>Moisture-free</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon</td>
<td>69.47</td>
<td>77.88</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>4.56</td>
<td>3.76</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>1.33</td>
<td>1.49</td>
</tr>
<tr>
<td>Oxygen</td>
<td>11.86</td>
<td>2.57</td>
</tr>
<tr>
<td>Sulfate sulfur</td>
<td>.07</td>
<td>.08</td>
</tr>
<tr>
<td>Pyritic sulfur</td>
<td>.96</td>
<td>1.08</td>
</tr>
<tr>
<td>Organic sulfur</td>
<td>1.65</td>
<td>1.84</td>
</tr>
<tr>
<td>Total sulfur</td>
<td>2.68</td>
<td>3.00</td>
</tr>
<tr>
<td>HHV ($kJ/kg$)</td>
<td>26,493</td>
<td>29,701</td>
</tr>
<tr>
<td>LHV ($kJ/kg$)</td>
<td>25,770</td>
<td>28,891</td>
</tr>
<tr>
<td>Free swelling index (FSI)</td>
<td>4</td>
<td>—</td>
</tr>
</tbody>
</table>
A 500-pound sample of Mississippian-age Gilmore City limestone was obtained from Ames, Iowa in Story County. It was prepared for use in briquettes and coal-water mixtures by crushing in the Holmes Model 45 pulverizer. The pulverizer was fitted with the same 0.0625-inch screen used for coal grinding and produced a limestone product with a top-size of about 20 mesh (0.850 mm).

A K.R. Komarek Model B-100 roll-briquetter was used to prepare the coal-limestone briquettes. Preliminary tests were performed with the briquetter to select the briquette forming parameters: roll pressure = 1500 psi, roll speed = 3, and screw speed = 2-3. Pre-gelatinized corn starch was selected for use in the briquette binder. The binder was produced by mixing 22% of corn starch by weight with water in an electric blender. Limestone was mixed with the pulverized coal to produce a mixture with a calcium-to-sulfur molar ratio (Ca/S) of 2.0. From the coal sulfur analysis, it can be determined that 16.7 kg of limestone must be added to every 100 kg of as-received coal. However, the inherent calcium content of the coal (CaO in ash) accounts for about 11% of this amount. Therefore, it was necessary to add only 14.9 kg of limestone to every 100 kg of coal. The binder mixture was then added to the coal-limestone mixture at an addition rate of 11% by weight. After hand-mixing the coal-limestone-binder mixture, approximately 200 kg of briquettes were prepared. The elliptical briquette had dimensions of 20 mm x 12 mm x 8 mm. The moisture-free composition of the briquettes was 83.2% moisture-free coal, 13.9% limestone, and 2.95% corn starch.

The coal-water-limestone mixture was prepared using the same limestone addition rate as for briquettes. Water was added to the coal-limestone mixture at a rate of 1.13 kg of water per kg of coal. Approximately 200 kg of CWLM were prepared.
in an emulsifier-mixer (powered by a 10-hp electric motor) for 10 minutes. The resulting mixture had the following composition: 49.7% dry coal, 8.3% limestone, and 42.0% water by weight; the solids loading was therefore 58%. This was the highest attainable solids loading that resulted in an easily-pumpable mixture. Both the CWLM and briquettes were stored in sealed 5-gallon buckets until needed for combustion tests.

6.3 Start-Up Procedure

The start-up procedure for each combustion test consisted of a series of equipment checks and combustor preheat. First, the gas sampling system was activated. This allowed the heat-taped lines to come up to operating temperature; also, the gas analyzers were used to monitor L-P gas combustion during preheat. The cooling water flow was started and the ignition spark was turned on. The computer was activated to monitor air flow rates, temperatures, and exhaust gas composition.

The combustor was preheated by burning L-P gas in the bed. Combustion bed air flow was set at about 30 scfm (standard cubic feet per minute). Air flow through the heat transfer bed was set at about 4 scfm to purge any L-P gas that might infiltrate this bed. The L-P flow rate was set at 0.75 cfm and ignition occurred within 5-10 seconds. The fuel burned at the surface of the bed initially, but began to burn within the bed as the bed temperature increased. Air and L-P flow rates were steadily decreased as the bed temperature rose. When the bed temperature reached 900 K, the coal feed was started. Then, as the temperature increased to about 1050 K, the L-P flow was turned off. The coal feed and combustion air flow rates were set at desired levels of energy release rate and bed stoichiometry.
The heat transfer bed air flow rate was set at a level to achieve a combustion bed temperature of 1140 K. To decrease combustion bed temperature the heat transfer air flow rate was lowered; to increase bed temperature the air flow rate was raised. The combustor was operated for 30-60 minutes before data acquisition began to allow steady state conditions to be reached.

6.4 Data Acquisition

Once steady state conditions were achieved for a particular test, the data acquisition program was started. The program was written in BASIC and was stored on the computer hard disk drive. The program controlled the collection of temperature, flow rate, and exhaust gas composition data and wrote these data to a data file on the hard drive once every minute. Each recorded temperature was a one-minute average of 12 5-second-interval readings. The gas concentrations and air flow rates were read once every minute, as they did not fluctuate greatly at steady state. Steady state data were recorded for 30-60 minutes for each test. Uncertainty in temperature measurement was ±20 K. Uncertainties in air flow rate measurement and in gas composition were ±5%. The coal feed rate was determined to be certain within ±5%.

6.5 Ash Analysis

An ash sample was collected for every test by opening the cyclone collection valve once test equilibrium was achieved. The sample was collected for 10-30 minutes depending on the exhaust gas flow rate. The sample was collected in a one-liter jar.
A 25 gram sample of the collected ash was dried in an oven, reweighed, and placed in an ashing oven at 800°C for 2 hours. When the sample had cooled, it was weighed and the weight loss was calculated. It was assumed that unburned carbon accounted for all of the weight loss. The carbon conversion efficiency could then be calculated by performing a carbon balance on the combustor; details on combustion mass balances are found in Appendix D.

6.6 Coal Feed Rate Determination

Coal feed rate into the combustor ranged between 1 and 10 kg/hr. These low flow rates were very difficult to measure on-line accurately. Before a test, the auger feeder was calibrated to provide an estimate of the feed rate during a test. However, for a more accurate measure of the fuel feed rate, it was determined after a test from a mass balance on the combustor. It was calculated based on accurately measured values of air flow rate, exhaust gas composition, and unburned carbon elutriation loss. See Appendix D for details on combustor mass balances.
7 RESULTS

7.1 Load Turndown

Load turndown tests were performed for the three fuel forms: crushed coal, coal-limestone briquettes, and coal-water-limestone mixture (CWLMI). Test data obtained for the crushed coal are shown in Table 7.1. A load turndown of 12.4 (obtained from the ratio of maximum fuel flow of 9.21 kg/hr to the minimum fuel flow rate of 0.74 kg/hr) at constant combustion bed temperature (1155 ± 25 K) was achieved. The temperature was held constant by admitting the proper amount of air through the heat transfer bed (secondary air flow) after the combustion fuel flow and air flow rates were set. Overall excess air is obtained as the ratio of total air flow to stoichiometric air flow rate. Percent of fuel burned in-bed is obtained from an energy balance on the combustion bed.

The combustor was designed for air-staged firing. The combustion bed was operated sub-stoichiometrically to suppress NO\textsubscript{x} formation. Overfire air was injected above the bed to complete fuel combustion at fuel lean conditions. This overfire air consisted of air exiting the heat transfer bed (secondary air) as well as air bypassing the heat transfer bed (tertiary air). The target value of excess air was 20% with a combustion bed equivalence ratio of 1.30 (77% of stoichiometric air). It was not possible to hold the excess air level at 20% for the lowest firing rates to avoid
Table 7.1: Load turndown test results for crushed coal

<table>
<thead>
<tr>
<th>Run</th>
<th>Fuel Feed Rate (kg/hr)</th>
<th>Secondary Air Flow (kg/hr)</th>
<th>Bed Air Flow Temperature (K)</th>
<th>Bed Temperature Equivalence Ratio</th>
<th>Overall Excess Air (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>A1</td>
<td>0.74</td>
<td>0.0</td>
<td>1133</td>
<td>0.44</td>
<td>126</td>
</tr>
<tr>
<td>A2</td>
<td>1.28</td>
<td>0.0</td>
<td>1172</td>
<td>0.71</td>
<td>67</td>
</tr>
<tr>
<td>A3</td>
<td>1.87</td>
<td>11.1</td>
<td>1156</td>
<td>1.30</td>
<td>42</td>
</tr>
<tr>
<td>A4</td>
<td>5.43</td>
<td>11.3</td>
<td>1144</td>
<td>1.52</td>
<td>5</td>
</tr>
<tr>
<td>A5</td>
<td>7.19</td>
<td>14.4</td>
<td>1161</td>
<td>1.37</td>
<td>10</td>
</tr>
<tr>
<td>A6</td>
<td>7.08</td>
<td>19.1</td>
<td>1172</td>
<td>1.05</td>
<td>50</td>
</tr>
<tr>
<td>A7</td>
<td>9.21</td>
<td>31.2</td>
<td>1178</td>
<td>1.36</td>
<td>18</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>( \text{O}_2 ) (%)</th>
<th>( \text{CO}_2 ) (%)</th>
<th>( \text{CO} ) (%)</th>
<th>( \text{SO}_2 ) (ppm)</th>
<th>( \text{NO}_x ) (ppm)</th>
<th>( \text{kg NO}_x ) per GJ</th>
</tr>
</thead>
<tbody>
<tr>
<td>10.6</td>
<td>7.9</td>
<td>0.051</td>
<td>---</td>
<td>300</td>
<td>0.24</td>
</tr>
<tr>
<td>7.4</td>
<td>10.9</td>
<td>0.029</td>
<td>---</td>
<td>325</td>
<td>0.19</td>
</tr>
<tr>
<td>5.6</td>
<td>12.4</td>
<td>0.034</td>
<td>1480</td>
<td>210</td>
<td>0.10</td>
</tr>
<tr>
<td>1.3</td>
<td>16.9</td>
<td>0.078</td>
<td>2385</td>
<td>435</td>
<td>0.16</td>
</tr>
<tr>
<td>3.9</td>
<td>15.9</td>
<td>0.070</td>
<td>1976</td>
<td>510</td>
<td>0.20</td>
</tr>
<tr>
<td>6.3</td>
<td>12.1</td>
<td>0.015</td>
<td>2000</td>
<td>520</td>
<td>0.27</td>
</tr>
<tr>
<td>3.8</td>
<td>15.2</td>
<td>0.037</td>
<td>2380</td>
<td>440</td>
<td>0.18</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Total Air Flow (kg/hr)</th>
<th>Carbon Conversion Efficiency (%)</th>
<th>Percent Fuel Burned in Bed (%)</th>
<th>Annular Bed Temperature (K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>15.4</td>
<td>95</td>
<td>100</td>
<td>675</td>
</tr>
<tr>
<td>19.5</td>
<td>95</td>
<td>71</td>
<td>803</td>
</tr>
<tr>
<td>24.2</td>
<td>94</td>
<td>71</td>
<td>694</td>
</tr>
<tr>
<td>52.4</td>
<td>95</td>
<td>51</td>
<td>656</td>
</tr>
<tr>
<td>72.7</td>
<td>96</td>
<td>50</td>
<td>658</td>
</tr>
<tr>
<td>97.1</td>
<td>97</td>
<td>57</td>
<td>638</td>
</tr>
<tr>
<td>100.0</td>
<td>95</td>
<td>52</td>
<td>610</td>
</tr>
</tbody>
</table>

\(^{a}\) Sum of primary, secondary, and tertiary air.
defluidization of the combustion bed. The excess air level ranged from 5% to 50% for the five highest firing rates (Runs A3-A7). This wide range was the result of an inability to accurately meter coal with the auger feeder. In subsequent tests, it was decided to use air flow rates in conjunction with exhaust gas analysis to set excess air level.

No limestone was added to the bed for these trials with dry coal. The sulfur emissions of 1480 ppm to 2385 ppm are in the range expected for the coal. Expected sulfur retention due to inherent calcium content was 11%; observed values ranged from 0% to 23%. The accuracy of sulfur balances was about 15%. Emissions of NO\textsubscript{x} ranged from 0.10 to 0.27 kg/GJ.

Load turndown test results for briquettes are shown in Table 7.2. A load turndown of 9.4 was achieved at constant combustion bed temperature (1110 ± 35 K). A test to determine conventional load turndown (the load turndown that could be obtained with a conventional one-bed design) is shown as table entry Run B6. Conventional load turndown is defined as the ratio of maximum to minimum firing rate with the annular bed defluidized, regardless of firing rate. Combustion bed temperature was allowed to rise no higher than in the two-bed turndown trials. The convention load turndown obtained from this test was only 5.3; this is compared to 9.4 obtained with the annular bed fluidized.

Test results for CWLM are shown in Table 7.3. A load turndown of 9.0 was achieved at constant combustion temperature (1110 ± 45 K). Again, the temperature was maintained constant by adjusting secondary air flow through the annular heat transfer bed for each firing rate.

Although a load turndown of 12.4 was achieved for crushed coal, turndown
Table 7.2: Load turndown results for briquettes

<table>
<thead>
<tr>
<th>Run</th>
<th>Fuel Feed Rate (kg/hr)</th>
<th>Secondary Air Flow (kg/hr)</th>
<th>Bed Temperature (K)</th>
<th>Bed Equivalence Ratio</th>
<th>Overall Excess Air (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>B1</td>
<td>1.08</td>
<td>0.0</td>
<td>1073</td>
<td>0.70</td>
<td>44</td>
</tr>
<tr>
<td>B5</td>
<td>3.46</td>
<td>5.3</td>
<td>1091</td>
<td>1.27</td>
<td>20</td>
</tr>
<tr>
<td>B4</td>
<td>4.85</td>
<td>8.6</td>
<td>1109</td>
<td>1.27</td>
<td>22</td>
</tr>
<tr>
<td>B3</td>
<td>5.77</td>
<td>11.0</td>
<td>1144</td>
<td>1.13</td>
<td>35</td>
</tr>
<tr>
<td>B2</td>
<td>10.10</td>
<td>30.6</td>
<td>1141</td>
<td>1.32</td>
<td>24</td>
</tr>
<tr>
<td>B6</td>
<td>5.69</td>
<td>0.0</td>
<td>1142</td>
<td>0.74</td>
<td>48</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>O₂ (%)</th>
<th>CO₂ (%)</th>
<th>CO (%)</th>
<th>SO₂ (ppm)</th>
<th>NOₓ (ppm)</th>
<th>kg NOₓ per GJ</th>
<th>kg S per GJ</th>
</tr>
</thead>
<tbody>
<tr>
<td>5.9</td>
<td>12.2</td>
<td>0.102</td>
<td>194</td>
<td>236</td>
<td>0.12</td>
<td>0.104</td>
</tr>
<tr>
<td>4.6</td>
<td>14.3</td>
<td>0.008</td>
<td>368</td>
<td>271</td>
<td>0.12</td>
<td>0.166</td>
</tr>
<tr>
<td>4.2</td>
<td>14.3</td>
<td>0.010</td>
<td>704</td>
<td>337</td>
<td>0.14</td>
<td>0.321</td>
</tr>
<tr>
<td>5.0</td>
<td>13.2</td>
<td>0.024</td>
<td>1170</td>
<td>342</td>
<td>0.16</td>
<td>0.586</td>
</tr>
<tr>
<td>5.0</td>
<td>14.6</td>
<td>0.024</td>
<td>936</td>
<td>313</td>
<td>0.14</td>
<td>0.437</td>
</tr>
<tr>
<td>7.2</td>
<td>11.8</td>
<td>0.022</td>
<td>706</td>
<td>353</td>
<td>0.18</td>
<td>0.394</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Total Air Flow (kg/hr)</th>
<th>Carbon Conversion Efficiency (%)</th>
<th>Percent Fuel Burned in Bed (%)</th>
<th>Sulfur Capture (%)</th>
<th>Annular Bed Temperature (K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>12.5</td>
<td>91</td>
<td>74</td>
<td>90</td>
<td>711</td>
</tr>
<tr>
<td>33.4</td>
<td>90</td>
<td>57</td>
<td>84</td>
<td>587</td>
</tr>
<tr>
<td>47.5</td>
<td>90</td>
<td>58</td>
<td>68</td>
<td>600</td>
</tr>
<tr>
<td>62.2</td>
<td>92</td>
<td>62</td>
<td>42</td>
<td>617</td>
</tr>
<tr>
<td>100.0</td>
<td>95</td>
<td>51</td>
<td>57</td>
<td>597</td>
</tr>
<tr>
<td>67.7</td>
<td>92</td>
<td>56</td>
<td>61</td>
<td>762</td>
</tr>
</tbody>
</table>

*a Sum of primary, secondary, and tertiary air.
### Table 7.3: Load turndown results for CWLM

<table>
<thead>
<tr>
<th>Run</th>
<th>Fuel Feed Rate (kg/hr)</th>
<th>Secondary Air Flow (kg/hr)</th>
<th>Bed Temperature (K)</th>
<th>Bed equivalence</th>
<th>Overall Excess Air (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>C2</td>
<td>2.49</td>
<td>0.0</td>
<td>1064</td>
<td>0.76</td>
<td>31</td>
</tr>
<tr>
<td>C1</td>
<td>5.28</td>
<td>4.1</td>
<td>1106</td>
<td>1.32</td>
<td>14</td>
</tr>
<tr>
<td>C3</td>
<td>22.29</td>
<td>30.8</td>
<td>1152</td>
<td>1.43</td>
<td>8</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th></th>
<th>Oxygen</th>
<th>Carbon Dioxide</th>
<th>Carbon monoxide</th>
<th>Sulfur Dioxide</th>
<th>Nitrogen Dioxide</th>
<th>kg NOx per GJ</th>
<th>kg S per GJ</th>
</tr>
</thead>
<tbody>
<tr>
<td>(%)</td>
<td>(%)</td>
<td>(ppm)</td>
<td>(ppm)</td>
<td>(ppm)</td>
<td>per GJ</td>
<td>per GJ</td>
<td></td>
</tr>
<tr>
<td>5.1</td>
<td>13.2</td>
<td>0.190</td>
<td>151</td>
<td>272</td>
<td>0.12</td>
<td>0.074</td>
<td></td>
</tr>
<tr>
<td>5.0</td>
<td>14.6</td>
<td>0.024</td>
<td>936</td>
<td>313</td>
<td>0.08</td>
<td>0.088</td>
<td></td>
</tr>
<tr>
<td>7.2</td>
<td>11.8</td>
<td>0.022</td>
<td>706</td>
<td>353</td>
<td>0.10</td>
<td>0.326</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Total Air Flow (kg/hr)</th>
<th>Carbon Conversion Efficiency (%)</th>
<th>Percent Fuel Burned in Bed (%)</th>
<th>Sulfur Capture (%)</th>
<th>Annular Bed Temperature (K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>16.8</td>
<td>91</td>
<td>53</td>
<td>93</td>
<td>693</td>
</tr>
<tr>
<td>30.8</td>
<td>91</td>
<td>40</td>
<td>91</td>
<td>680</td>
</tr>
<tr>
<td>123.2</td>
<td>94</td>
<td>38</td>
<td>68</td>
<td>603</td>
</tr>
</tbody>
</table>

* Sum of primary, secondary, and tertiary air.
was limited to about 9.0 for briquettes and CWLM. One possible explanation for this is evident by re-examining the minimum fuel flow rates in Tables 7.1, 7.2, and 7.3. For the crushed coal case, 100% of the fuel is burned in the bed, while for briquettes and CWLM only 74% and 53% respectively is burned in the bed. A greater percentage of fuel burned in the bed allows lower fuel firing rates to maintain the same combustion temperature. Lower minimum firing rates and correspondingly larger turndown would probably have resulted with briquettes and CWLM if leaner stoichiometries had been allowed with these fuels.

Load turndown ratio is best defined as the ratio of maximum to minimum firing rates for a combustor. In developing Equation 2.5 it was necessary to employ an alternative, but nearly equivalent, definition of load turndown; that is, load turndown was defined as the ratio of maximum to minimum heat transfer rates between the combustion bed and water jacket. The overall heat transfer coefficient between combustion bed and water jacket, $U_{beds}$, can be defined as the total heat transfer rate per unit wall area to the heat transfer bed divided by the temperature gradient between the combustion bed and water jacket. The subscript ”beds” is used here to emphasize that the overall heat transfer coefficient is limited by the convection coefficients in the fluidized beds. Figure 7.1 is a plot of overall heat transfer coefficient versus annular bed air flow rate for the coal combustion tests. The ratio of maximum to minimum overall heat transfer coefficient represents this alternative definition of load turndown ratio. The value of 7.0 obtained from Figure 7.1 is in good agreement with calculations using Equation 2.5. This alternative definition of load turndown assumes that the same fraction of fuel is burned in-bed for both maximum and minimum fuel flow rates. It was predicted by the
Figure 7.1: Overall heat transfer coefficient vs. fluidization velocity
combustion model and observed in the combustion tests that a higher fraction of fuel is burned in-bed for low fuel flow rates than for high fuel flow rates. For this reason, higher load turndowns were observed in the combustion tests than predicted by heat transfer considerations alone.

7.2 Emissions

Both the briquettes and the CWLM were formulated with a calcium-to-sulfur molar ratio of 2.0. Sulfur dioxide emission reduction of 42% to 93% was observed in the load turndown tests; emissions ranged from 0.07 to 0.59 kg S/GJ. Most tests met the EPA New Source Performance Standards of 0.52 kg/GJ [26]. The effect of combustion bed temperature on sulfur retention is shown in Figure 7.2. The data is compared to experimental results of Stantan [27] who found the optimal temperature for sulfur retention to be close to 1100 K. In contrast, the optimum bed temperature in these tests was found to be no higher than 1050 K. This discrepancy with results of other researchers possibly arises from the fact that limestone in briquettes and CWLM more closely follows the temperature history of fuel particles, in which it is incorporated, than the bulk bed material. Since fuel particles burn at temperatures 50 to 200 K higher than the bed [28], the optimal bed temperature for sulfur retention is expected to be somewhat lower for fuel forms that contain limestone. Furthermore, higher sulfur retention than that reported by Stantan was observed. This result, as well, is probably related to the intimate contact of limestone with coal briquettes and CWLM.

The effect of combustion bed temperature on emissions of NO$_x$ is illustrated in Figure 7.3. The general trend is for increased emissions of NO$_x$ with increasing bed
Figure 7.2: Sulfur retention vs. combustion bed temperature
Figure 7.3: Nitrogen oxide emission vs. combustion bed temperature
temperature, as expected. However, NO\textsubscript{x} formation also increases with combustion bed excess air. Experimental observations agree qualitatively with other researchers who concluded that, to minimize NO\textsubscript{x} emissions, the combustion bed should be operated sub-stoichiometrically at low combustion bed temperatures [12,3,29]. Emissions of NO\textsubscript{x} ranged from 0.08 to 0.27 kg/GJ. Most tests met the EPA New Source Performance Standards of 0.26 kg NO\textsubscript{x} / GJ [26].

7.3 Estimate of Largest Practical Combustor

Calculations were performed to estimate the largest practical combustor which could employ this two-bed design. The following assumptions were used:

- Annular bed width is 2.5 cm.
- The overall heat transfer coefficient between combustion bed and water jacket is 150 W/m\textsuperscript{2}K (see Figure 7.1).
- Fuel consists of dry coal with a heating value of 23258 kJ/kg that requires 11.25 kg air per kg of coal for 20% excess air.
- Water jacket wall temperature is 373 K and combustion bed temperature is 1144 K.
- Combustion bed is operated sub-stoichiometrically and 50% of fuel energy is released in the combustion bed.
- The combustor geometry is cylindrical, and bed height is limited to 1 m due to pressure drop considerations.
The heat transfer rate from the combustor wall can be equated to the energy release in the bed and the following solution is obtained:

- Combustion bed diameter = 1.43 m.
- Total combustor heat load = 1.04 MW.

Larger units could be obtained by using a modular system composed of several smaller combustors.

### 7.4 Comparison with Heat Transfer Model

Experimental results obtained for briquettes (See Table 7.2) have been compared to simulation predictions. Figure 7.4 plots firing rate in the combustor versus secondary air flow through the heat transfer bed for constant combustion temperature. Model prediction and experimental results show two major discrepancies. First, the model predicts a sudden increase in sustainable firing rate when secondary air flow is increased to the minimum fluidization velocity of the annular bed. In contrast, the experimental results show a more gradual increase in firing rate with increasing secondary air flow. Second, the model under-predicts the amount of energy that can removed from the combustor at the highest secondary air flow rates. Both of these discrepancies apparently arise from the convection coefficients calculated for the annular bed from the two-phase theory of fluidization.

Most heat transfer correlations that have been developed for fluidized beds assume that the bed is either slugging or bubbling. Slugging occurs when large bubbles with a diameter on the order of the bed width form, and travel up through the bed as one large slug. Bubbling beds have more distinct, smaller bubbles that rise up
Briquette Data

Figure 7.4: Energy released in-bed vs. annular air flow rate – experiment vs. two-phase theory prediction
through the bed. The aspect ratio of a fluidized bed is equal to its height divided by its width. The two-phase theory of fluidization that was used to predict heat transfer coefficients was expected to yield accurate results for low aspect-ratio slugging beds. However, although the heat transfer bed operates in the slugging regime, the unusually high aspect ratio may alter the hydrodynamics enough to make the two-phase theory invalid. For the annular heat transfer bed with a fluidized height of 0.25 m, the aspect ratio is 10. Pflum et al. [30] compared experimentally-obtained convection coefficients in high-aspect ratio fluidized beds to two-phase theory predictions. Their findings suggested two major differences between experiment and theory: measured convection coefficients were significantly higher than predicted by two-phase theory and increased more gradually with fluidization velocity than predicted by the two-phase theory. Figure 7.5 shows a comparison of their experimental and predicted values of convection coefficient versus fluidizing velocity. These experimental data were employed in the computer model in place of two-phase theory values.

Convection coefficients for the combustion bed were still calculated using the two-phase theory. Since this bed has a low aspect-ratio, the two-phase theory was expected to predict convection coefficients accurately. Also, the convection coefficient in the combustion bed becomes rate-limiting only at the highest annular air flow rates. Figure 7.6 shows energy released in the combustion bed versus annular air fluidizing velocity using the experimental convection coefficients. It can be seen that significant improvement has been obtained over Figure 7.4. There is still some discrepancy in the mid-range flow rates of 4-10 m³/hr. A possible explanation for this discrepancy is non-uniform air distribution in the annular bed. If one zone of
Figure 7.5: Convection coefficient vs. fluidization velocity – experimental data
Figure 7.6: Energy released in-bed vs. annular air flow rate – experiment vs. experimental heat transfer data
the annular bed fluidizes before the rest of the bed, a more linear response in firing rate with changes in fluidization velocity is expected. Then, as more air is admitted to the annular bed, the firing rate should reach a maximum as the heat transfer coefficient in the combustion bed becomes rate-limiting to the overall heat transfer process.

7.5 Comparison with Combustion Model

Experimental results obtained for crushed coal (See Table 7.1) have been compared to combustion model predictions. The combustion model was executed as follows:

1. Bed temperature and air flow rates were used as input parameters to the heat transfer model.

2. Output data from the heat transfer model (including energy released in-bed, air velocity, bubble diameter and rise velocity, and bed voidage) were used as input parameters to the combustion model.

The combustion model then predicts percent of fuel energy released in-bed, coal mass flow rate, and incomplete combustion energy loss due to unburned carbon and unreacted carbon monoxide.

Figure 7.7 shows a plot of predicted fraction of fuel burned in-bed vs. experimental values from crushed coal tests. It can be seen that the best agreement is for the four cases where the combustion bed was operated with an equivalence ratio greater than 1.05; these cases represent the higher fuel flow rates. Model predictions
Figure 7.7: Fraction of fuel burned in bed - predicted vs. observed

- Perfect agreement
- □ Equivalence ratio = 0.44
- ○ Equivalence ratio = 0.71
- △ Equivalence ratio > 1.05
are in poor agreement for the other two cases; these represent the lowest fuel firing rates.

Figure 7.8 is a bar graph showing combustion model predictions of fraction of fuel burned in-bed, fraction of incomplete combustion due to unburned carbon monoxide, and fraction due to unburned carbon, for three different coal mass flow rates. At a fuel flow rate of 0.90 \(\text{kg/hr}\), 85% of the fuel is burned in-bed and 15% leaves the bed in the form of unburned carbon monoxide. No loss due to unburned carbon is predicted. At the highest fuel flow rate of 9.5 \(\text{kg/hr}\), 46% of the fuel is burned in-bed, 1% leaves the bed in the form of unburned carbon monoxide, and 53% of the incomplete combustion loss is due to unburned carbon.

At the lowest fuel flow rate, the combustion bed is nearly minimally fluidized and is operated with high excess air. It is reasonable to expect that the char would burn completely to carbon monoxide. Also, because the model assumes that carbon monoxide burns only in the bubble phase, some unburned carbon monoxide would be expected (because near minimum fluidization, bubble phase volume is much less than emulsion phase volume). In the test case where 100% combustion in-bed was observed, some of this predicted unburned carbon monoxide must be burning in the emulsion phase.

At high fuel flow rates, it is predicted that the major source of incomplete in-bed combustion is unburned carbon. Loss due to unburned carbon monoxide is essentially zero. These predictions agree with those made by Walsh [17], who compared his fluidized bed combustion model to experimental data of others. Walsh’s model assumes that carbon monoxide formed during char oxidation is rapidly consumed and that the major source of incomplete combustion is due to elutriated
Figure 7.8: Predicted fraction of fuel burned in-bed and sources of incomplete combustion vs. fuel mass flow rate
unburned carbon. Carbon fines continue to burn in the freeboard; unburned carbon exiting the freeboard represents overall combustor carbon loss. In the present combustion experiments, it was observed that unburned carbon exiting the freeboard was usually between 3% and 10%. Since it is predicted that about 50% of the carbon elutriates from the bed, significant freeboard combustion must be taking place. Overfire air and the insulated freeboard used in the combustion tests encourage freeboard combustion.

If char particles burn as shrinking spheres, unburned carbon loss would be very small (since the carbon fines would be much less massive than the char particles). Walsh argued that char particles burn as constant diameter particles until the reach a critical porosity. Then, the particles fragment and elutriate from the bed. This would result in large amounts of elutriated carbon from the bed. The present combustion model supports the fragmentation theory, as up to 60% of unburned carbon is predicted.

Table 7.4 shows combustion model predictions of coal mass flow rate. The first column is measured coal feed rate, the second column is combustion model prediction using the heat transfer (HT) model prediction of energy released in-bed \( Q_{in} \), and the third column is combustion model prediction using measured values of \( Q_{in} \). In the second column, it is seen that, as with the heat transfer model, the best agreement is at minimum and maximum fuel flow rates. Predictions based on experimental values of energy release in-bed (column 3) offer slightly better agreement. However, this slight improvement is at the expense of the model becoming more dependent on experimental data.
Table 7.4: Fuel flow rate - combustion model predictions vs. experimental data

<table>
<thead>
<tr>
<th>Actual Coal Feed Rate (kg/hr)</th>
<th>Predicted Coal Feed Rate (with $Q_{in}$ from HT model) (kg/hr)</th>
<th>Predicted Coal Feed Rate (with measured $Q_{in}$) (kg/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.66</td>
<td>1.04</td>
<td>0.91</td>
</tr>
<tr>
<td>1.13</td>
<td>0.93</td>
<td>0.95</td>
</tr>
<tr>
<td>4.84</td>
<td>6.38</td>
<td>4.93</td>
</tr>
<tr>
<td>6.37</td>
<td>7.47</td>
<td>6.89</td>
</tr>
<tr>
<td>6.28</td>
<td>8.35</td>
<td>7.97</td>
</tr>
<tr>
<td>8.21</td>
<td>8.56</td>
<td>9.47</td>
</tr>
</tbody>
</table>
Control of heat transfer rate independent of combustion conditions in a fluidized bed combustor is possible using a two-bed design. The second bed (heat transfer bed) surrounds the combustion bed and is independently fluidized. The heat transfer bed establishes the overall heat transfer coefficient between the combustion bed and water jacket. A load turndown of 12 was achieved in the two-bed fluidized combustor when burning crushed coal, while holding combustion temperature constant. Slightly lower turndown ratios were observed for coal-limestone briquettes and coal-water-limestone mixture (CWLM), and were attributed to slightly different operating conditions. Emissions of NO\textsubscript{2} ranged from 0.08 to 0.27 kg NO/GJ. Emissions of SO\textsubscript{2} ranged from 0.074 to 0.586 kg S/GJ for briquettes and CWLM, which translates to 42% to 93% sulfur retention.

The heat transfer model predictions were in reasonable agreement with experimental observations. The two-phase theory heat transfer predictions were not sufficiently accurate for the high aspect-ratio annular fluidized bed, so experimental heat transfer data was used. Simple calculations showed that the largest practical combustor employing this two-bed, water-wall design would be 1.43 m in diameter, with 1.04 MW thermal output.

The combustion model predictions were in reasonable agreement with experi-
imental observations. The model predicted well in cases of high fuel flow rates and poorly in cases where the combustion bed was minimally fluidized. Combustion model predictions support the fragmentation theory, whereby char particles burn at constant diameter until reaching a critical porosity, when they fragment and elutriate from the bed. For most cases, the major source of incomplete in-bed combustion was due to elutriated unburned carbon.

Recommendations for future work include:

1. A more detailed analysis of the combustion process in fluidized beds. The effect of operating parameters such as excess air and fuel flow rate on percent of fuel burned in bed needs to be determined.

2. A more fundamental study of heat transfer coefficients in high aspect ratio fluidized beds. This study is needed to explain why observed heat transfer coefficients were higher than predicted by two-phase theory.
9 BIBLIOGRAPHY


This program solves the system of equations for temperature and heat transfer in a two-bed fluidized bed combustor.

The equations for calculating heat transfer coefficients in a fluidized bed are contained in the subroutine COEFFC for the center bed and COEFFA for the annular heat transfer bed.

The subroutine MINFLU evaluates minimum fluidization velocity and PROP determines air properties.

LAST UPDATE: 6/20/89 by Jim Foley

DOUBLE PRECISION AORIF, EPSB, EPSIW, EPSOW, EPSP, CPS, EMFA, EMFC
DOUBLE PRECISION EOA, EOC, RHOS, RHOL, RHOL, RHOL, RHOL, RHOL, SIG, CPW
DOUBLE PRECISION CSF, PRANW, RJUNK1, RJUNK2, DPC, DPA, BOILC, PI
DOUBLE PRECISION ABAREA, CBAREA, AHDEAD, CHDEAD, CPNOT, HJACK
DOUBLE PRECISION SAIW, SAOW, RHONOT, GPMW, UOVERUMF, CSCFM, COAL
DOUBLE PRECISION ASCFM, AFRA, DPAM, DPCM, DBC
DOUBLE PRECISION VFBIW, VFPIW, VFPW, UBIW, UPIW, UPOW, DC, DA
DOUBLE PRECISION UNOT, UHTNOT, MAIR, MHTAIR, UAIR, UHTAIR
DOUBLE PRECISION CPAIR, RHOAIR, RMUAIR, RAIR, PRAIR
DOUBLE PRECISION CPHA, RHOHT, RMUHT, RKHT, PRHT
DOUBLE PRECISION ARA, UMPHT, KE0, HGCA, ARC, UMFC, KE0C, HGCC
DOUBLE PRECISION EC, EBC, UBC, HEIGHTC, HPCC, HB
DOUBLE PRECISION HA, FLUX, QCOND
DOUBLE PRECISION TNOT, QIN, TB, TIW, TP, TOW, TW
DOUBLE PRECISION VAL(5), ERR(5), TEST, TOL
DOUBLE PRECISION QCWALL, QCAIR, QCRAD, QARADI, QARADO, QWALLI
C INPUT DATA FOR FLUIDIZED BED:

DATA AORIF, EPSB, EPSIW, EPSOW, EPS, CPA, EPS, EOA, EOC, RHDS, RMUW, HFG, G, RHOL, CPW, CSF, PRANW
* /2.0268D-04, 0.8D0, 0.40D0, 0.40D0, 0.90D0,
* 800.0D0, 0.476D0, 0.476D0, 0.40D0, 0.40D0,
* 2600.0D0, 279.0D-06, 2257.0D+03, 9.806D0, 957.9D0,
* 0.5955D0, 58.9D-03, 4217.0D0, 0.0130D0, 1.76D0 /

DATA TNOT, DA, AHDEAD, CHDEAD, CPNOT, HJACK, RHONOT, GPMW
* /294.0D0, 0.254D0, 0.229D0, 0.178D0,
* 1007.0D0, 0.2286D0, 1.19D0, 1.50D0 /

PRINT*, 'WHAT IS THE ANNULAR BED PARTICLE DIAMETER? '
PRINT*, ' IN MICRONS: '
PRINT*, ','
READ(5,*) DPAM
DPA = DPAM*1.0D-06
PRINT*, 'WHAT IS THE CENTRAL BED PARTICLE DIAMETER? ',
PRINT*, ' IN MICRONS: '
READ(5,*) DPCM
DPC = DPCM*1.0D-06
PRINT*, 'INPUT CENTRAL BED DIAMETER (IN INCHES): ',
PRINT*, ' FOR EXAMPLE: 5,7, OR 8 INCHES . . . ',
PRINT*, ','
READ(5,*) DC
DC = DC*0.02540D0
BOILC = RMUW*HFG*DSQRT( G*(RHOL-RHOU)/SIG )*
* ( (CPW/(CSF*HFG*PRANW))**3 )
PI = 4.0D0*DATAN(1.0D0)
ABAREA = ( DA*DA - DC*DC )*PI/4.0D0
85

CBAREA = PI*DC*DC/4.0D0
SAIW = PI*HJACK*DC
SAOW = PI*HJACK*DA
WRITE(6,10)
WRITE(6,20) DC,DPCM,CHDEAD
WRITE(6,30) DPAM,AHDEAD

10 FORMAT('1',X,'TWO-BED FLUIDIZED BED COMBUSTOR SIMULATION')
20 FORMAT('1',X,'COMB BED DIAMETER : ',F6.3,' m',/,X,
* 'COMB BED PARTICLE SIZE : ',F7.1,' MICRONS',/,X,
* 'COMB BED DEAD HEIGHT : ',F6.3,' m')
30 FORMAT('1',X,'ANNULAR BED PARTICLE SIZE : ',F7.1,' MICRONS'
* ,/X, 'ANN BED DEAD HEIGHT : ',F6.3,' m',/X,
* 'UNITS ARE AS FOLLOWS: ',/,6X,'U (m/s) Q (kW) V (scfm)',
* 'UA (W/K) h (W/m2K) T (K) A/F (scfm/kW)',/X)

C INPUT AIRFLOWS AND (BED TEMPERATURE OR ENERGY RELEASED IN BED)

5 PRINT*,'INPUT U/Umf FOR ANNULAR BED :
PRINT*,''
READ(5,*) UOVERUMF
PRINT*,''
PRINT*,'INPUT PRIMARY (combustion bed) AIRFLOW (scfm) :
PRINT*,''
READ(5,*) CSCFM
PRINT*,'

50 PRINT*,'DO YOU KNOW : 1) Combustion Temperature 
PRINT*,' OR : 2) QIN 
READ(5,*) II
IF (II.NE.1.AND.II.NE.2) THEN
   PRINT*,'MUST BE 1 OR 2!'
   GOTO 50
ELSE
   CONTINUE
ENDIF

IF (II.EQ.1) THEN
PRINT*, 'INPUT TBED (K)'
READ(5,*) TB

ELSE

PRINT*, 'QIN (kW)'
READ(5,*) QIN
QIN = QIN*1000.0D0
TB = 1144.0D0 ! INITIAL GUESS

ENDIF

C CALCULATE RADIATION VIEW FACTORS:

VFBIW = 5.68D-8 / ( 1.0D0/EPSB+1.0D0/EPSIW-1.0D0 )
VFPIW = 5.68D-8 / ( 1.0D0/EPSP+1.0D0/EPSIW-1.0D0 )
VFPOW = 5.68D-8 / ( 1.0D0/EPSP+1.0D0/EPSGW-1.0D0 )

C MAKE INITIAL GUESSES

TEST = 5.0D0
TOL = 0.00005D0 ! CONVERGENCE CRITERION
TW = 300.0D0 ! AVERAGE WATER TEMP.
TOW = 373.1D0 ! WATER WALL TEMP.
TP = 670.0D0 ! HT ANNULAR BED TEMP.
TIW = 900.0D0 ! INSERT WALL TEMP.

C CALCULATE INITIAL VALUES OF RADIATION COEFFICIENTS

UBIW = VFBIW*(TB*TB+TIW*TIW)*(TB+TIW) ! W/m2K
UPIW = VFPIW*(TP*TP+TIW*TIW)*(TP+TIW)
UPGW = VFPOW*(TP*TP+TOW*TOW)*(TP+TOW)

UNOT = CSCFM*4.7195D-04/CBAREA ! m/sec
MAIR = CSCFM*RHONOT*4.7195D-04 ! kg/sec
UMFHT = 0.0500D0 ! m/sec
IF (UOVERUMF.GE.1.00D0) THEN ! ANNULAR BED IS FLUIDIZED

IF (II.EQ.1) THEN ! COMBUSTION TEMP. IS KNOWN

DO 100 WHILE (TEST.GT.TOL)

CALL PROP ( TP,RHOHT,CPHT,RMUHT,RKHT,PRHT,IFLAGHT )
CALL PROP ( TB,RHOAIR,CPAIR,RMUAIR,RKAIR,PRAIR,IFLAGA )

UHTAIR = UOVERUMF*UMFHT         ! m/sec
UHTNOT = UHTAIR*TNOT/TP          ! m/sec
MHTAIR = UHTNOT*RHOAIR*ABAREA   ! kg/sec
UAIR = UNOT*TB/TNOT             ! m/sec

CALL MINFLU ( UHTAIR,RHOHT,CPHT,RKHT,RMUHT,EOA,DPA,RHOS,
               AHDEAD,PI,ARA,UMFHT,KEO,HGCA )
CALL MINFLU ( UAIR,RHOAIR,CPAIR,RKAIR,RMUAIR,EOC,DPC,RHOS,
               CHDEAD,PI,ARC,UMFC,KEGC,HGCC )

CALL COEFFC ( UAIR,CPAIR,RHAIR,RMUAIR,RKAIR,DPC,CPA,RHOS,
               EEC,EMFC,CHDEAD,DC,AORIF,UMFC,EC,EBC,UBC,DBC,
               HEIGHTC,HGCC,HPCC,HB )
CALL COEFFA ( DPA,UOVERUMF,HGCA,HA,FLAGDP )

CPAIR = 0.50D0*(CPNOT+CPAIR)
CPHT = 0.50D0*(CPNOT+CPHT)

VAL(1) = QIN
VAL(2) = TIW
VAL(3) = TP
VAL(4) = TOW

QIN = MAIR*CPAIR*(TB-TNOT) + SAIW*(HB+UBIW)*(TB-TIW)
TIW = ( (HB+UBIW)*TB + (HA+UPIW)*TP ) / ( (HB+UBIW) + (HA+UPIW) )
TP = ( MHTAIR*CPHT*TNOT + SAIW*(HA+UPIW)*TIW + 
       SAOW*(HA+UPGW)*TOW ) /
     ( MHTAIR*CPHT + SAIW*(HA+UPIW) + SAOW*(HA+UPGW) )
FLUX = (HA+UPOU)*(TP-TOW)
TOW = 373.ODO + (FLUX/BOILC)**(l.0D0/3.0D0)

ERR(1) = DABS( (QIN-VAL(1))/QIN )
ERR(2) = DABS( (TIW-VAL(2))/TIW )
ERR(3) = DABS( (TP -VAL(3))/TP )
ERR(4) = DABS( (TOW-VAL(4))/TOW )
TEST = 0.ODO

DO 99 I=1,4
   IF (ERR(I).GT.TEST) TEST=ERR(I)
99 CONTINUE

C *********************************************************

ELSE
   !QIN IS KNOWN

DO 200 WHILE (TEST.GT.TOL)

CALL PROP ( TP,RHOHT,CPTH,RMUHT,RKHT,PRHT,IFLAGHT )
CALL PROP ( TB,RHAIR,CPAIR,RRMUAIR,RKAIR,PRAIR,IFLAGA )
UHTAIR = UOVERRHOM*UMFHT !m/sec
UHTNOT = UHTAIR*TNOT/TP !m/sec
MHTAIR = UHTAIR*RHOHT*ABAREA !kg/sec
UAIR = UNOT*TB/TNOT !m/sec
CALL MINFLU (UHTAIR,RHHT,CPT,RKHT,RMUHT,EOA,DPA,RHOS,  *
   AHEAD,PI,ARA,UMFHT,KEO,HGCA)
CALL MINFLU (UAIR,RHAIR,CPAIR,RKAIR,RMUAIR,EOC,DPC,RHOS,  *
   CHDEAD,PI,ARC,UMFC,KEOC,HGCC)
CALL COEFFC ( UAIR,CPAIR,RHAIR,RRMUAIR,RKAIR,DPC,CP,S,RHOS,  *
   EOC,EMFC,CHDEAD,AC,ARIF,UMFC,EC,EBC,UBC,DBC,  *
   HEIGHTC,HGCC,HPC,HB)
CALL COEFFA ( DPA,UOVERRHOM,HGCA,HA,FLAGDP )

CPAIR = 0.50DO*(CPNOT+CPAIR)
CPTH = 0.50DO*(CPNOT+CPTH)
TP = 0.50D0*(TIW+TOW)

VAL(1) = TB
VAL(2) = TIW
VAL(3) = TP
VAL(4) = TOW

TB = ( QIN + MAIR*CPAIR*TNOT + SAIW*(HB+UBIW)*TIW ) /
* ( MAIR*CPAIR + SAIW*(HB+UBIW) )

TIW = ( (HB+UBIW)*TB + (HA+UPIW)*TP ) /
* ( (HB+UBIW) + (HA+UPIW) )

TP = ( MHTAIR*CPHT*TNOT + SAIW*(HA+UPIW)*TIW +
* SAIW*(HA+UPIW)*TOW ) /
* ( MHTAIR*CPHT + SAIW*(HA+UPIW) + SAIW*(HA+UPIW) )

FLUX = (HA+UPIW)*(TP-TOW)
TOW = 373.0D0 + (FLUX/BOILC)**(1.0D0/3.0D0)

ERR(1) = DABS( (TB -VAL(1))/TB )
ERR(2) = DABS( (TIW-VAL(2))/TIW )
ERR(3) = DABS( (TP -VAL(3))/TP )
ERR(4) = DABS( (TOW-VAL(4))/TOW )

TEST = 0.0D0

DO 199 I=1,4
   IF (ERR(I).GT.TEST) TEST=ERR(I)
199 CONTINUE

C *********************************************************
ENDIF

ELSE
   !ANNULAR BED IS DEFLUIDIZED

   IF (II.EQ.1) THEN
      !COMBUSTION TEMP. IS KNOWN

      DO 300 WHILE (TEST.GT.TOL)

      CALL PROP ( TP,RHOHT,CPHT,RMUHT,RKHT,PRHT,IFLAGHT )

300 CONTINUE
CALL PROP (TB,RHOAIR,CPAIR,RMUAIR,RKAIR,PRAIR,IFLAGA)

UHTAIR = UOVERUMF*UMFHT !m/sec
UHTNOT = UHTAIR*TNOT/TP !m/sec
MHTAIR = UHTNOT*RHONOT*AAREA !kg/sec
UAIR = UNOT*TB/TNOT !m/sec

CALL MINFLU (UHTAIR,RHOHT,CPHT,RKHT,RMUHT,EOA,DPA,RHOS,
*        AHDEAD,PI,ARA,UMFHT,KEO,HGCA)
CALL MINFLU (UAIR,RHOAIR,CPAIR,RKAIR,RMUAIR,EOC,DPC,RHOS,
*        CHDEAD,PI,ARC,UMFC,KEOC,HGCC)
CALL COEFFC (UAIR,CPAIR,RHOAIR,RMUAIR,RKAIR,DPC,CPH,
*        CHDEAD,PI,ARC,UMFC,KEOC,HGCC,HD)

CPAIR = 0.50DO*(CPN0T+CPAIR)
CPHT = 0.50DO*(CPN0T+CPHT)
TP = 0.50DO*(TIW+TOW)
VAL(1) = QIN
VAL(2) = TIW
VAL(3) = TOW

TP = 0.50DO*(TIW+TOW)
QIN = MAIR*CPAIR*(TB-TNOT) + SAIW*(HB+UBIW)*(TB-TIW)
TIW = ( (-2.0DO*KEO/(DLOG10(DC/DA)))*TOW +
*        DC*(HB+UBIW)*TB - MHTAIR*CPHT*(TP-TNOT)/(PI*HJACK) )/
*        ( (-2.0DO*KEO/(DLOG10(DC/DA)) + DC*(HB+UBIW) )
QCOND = - ( 2.0DO*PI*HJACK*KEO*(TIW-TOW))/(DLOG10(DC/DA))
TOW = ( QCOND/(BDILC*PI*DA*HJACK) )**(1.0DO/3.0DO)
*        + 373.0DO

ERR(1) = DABS( (QIN-VAL(1))/QIN )
ERR(2) = DABS( (TIW-VAL(2))/TIW )
ERR(3) = DABS( (TOW-VAL(3))/TOW )
TEST = 0.0DO

DO 299 I=1,3
   IF (ERR(I).GT.TEST) TEST=ERR(I)
CONTINUE

C

*****************************************************************************

ELSE

! QIN IS KNOWN

DO 400 WHILE (TEST.GT.TOL)

CALL PROP ( TP,RHOHT,CPHT,RMUHT,RKHT,PRHT,IFLAGHT )
CALL PROP ( TB,RHOAIR,CPAIR,RMUAIR,RKAIR,PRAIR,IFLAGA )

UHTAIR = UOVERUMF*UMFHT  ! m/sec
UHTNOT = UHTAIR*TNOT/TP   ! m/sec
MHTAIR = UHTNOT*RHONOT*ABAREA ! kg/sec
UAIR = UNOT*TB/TNOT      ! m/sec

CALL MINFLU (UHTAIR,RHOHT,CPHT,RKHT,RMUHT,EDO,DPA,RHOS,
              CHDEAD,PI,ARA,UMFHT,KEO,HGCA)
CALL MINFLU (UAIR,RHOAIR,CPAIR,RKAIR,RMUAIR,EOC,DPC,RHOS,
              CHDEAD,PI,ARC,UMFC,KEOC,HGCC)
CALL COEFFC ( UAIR,CPAIR,RHOAIR,RKAIR,CPS,RHOS,
               EOC,EMFC,CHDEAD,DC,AORIF,UMFC,EC,EBC,UBC,DBC,
               HEIGHTC,HGCC,HPCC,HB )

CPAIR = 0.50DO*(CPNOT+CPAIR)
CPHT = 0.50DO*(CPNOT+CPHT)

VAL(1) = TB
VAL(2) = TIW
VAL(3) = TP
VAL(4) = TOW

TP = 0.50DO*(TIW+TOW)
TB = ( QIN + MAIR*CPAIR*TNOT + SAIW*(HB+UBIW)*TIW ) /
     ( MAIR*CPAIR + SAIW*(HB+UBIW) )
TIW = ( (-2.0DO*KEO/(DLOG10(DC/DA)))*TGW +
       DC*(HB+UBIW)*TB - MHTAIR*CPHT*(TP-TNOT)/(PI*HJACK) )/
      ( (-2.0DO*KEO/(DLOG10(DC/DA)) + DC*(HB+UBIW) )
QCOND = - ( $2.0D0*PI*HJACK*KE0*(TIW-TOW) )/(DLG10(DC/DA))
TOW = ( QCOND/(BOILC*PI*DA*HJACK) )**($1.0D0/3.0D0)
* + 373.0DO

ERR(1) = DABS( (TB -VAL(1))/TB )
ERR(2) = DABS( (TIW-VAL(2))/TIW )
ERR(3) = DABS( (TP -VAL(3))/TP )
ERR(4) = DABS( (TOW-VAL(4))/TOW )
TEST = 0.0DO

DO 399 I=1,4
  IF (ERR(I).GT.TEST) TEST=ERR(I)
399 CONTINUE
400 CONTINUE

C ***************************************************************
ENDIF
C ***************************************************************
ENDIF

*************** IF ANNULAR BED IS FLUIDIZED ***************
IF (UOVERUMF.GE.1.00DO) THEN

ASCFM = MHTAIR*1763.67D0
RATIOA = UOVERUMF
RATIOC = UAIR/UMFC
QIN = QIN*1.0D-03
QCWALL = HB*HJACK*PI*DC*(TB-TIW)*1.0D-03
QCAIR = MAIR*CPAIR*(TB-TNDT)*1.0D-03
QCRAD = UBIW*(TB-TIW)*HJACK*PI*DC*1.0D-03
QARADI = UPIW*(TIW-TP)*HJACK*PI*DC*1.0D-03
QWALLI = HA*HJACK*PI*DC*(TIW-TP)*1.0D-03
QWALLO = HA*HJACK*PI*DA*(TP-TGW)*1.0D-03
QARADO = UPDOW*(TP-TGW)*HJACK*PI*DA*1.0D-03
QHTAIR = MHTAIR*CPHT*(TP-TNOT)*1.0D-03
QH2O = FLUX*HJACK*PI*DA*1.0D-03
UABEDS = 1000.0DO*(QH2O+QHTAIR)/(TB-TP)
AFRAT = CSCFM/QIN

ELSE

*************** IF ANNULAR BED IS DEFLUIDIZED

QCAIR = MAIR*CPAIR*(TB-TNCT)*1.0D-03
QCRAD = UBIW*HJACK*PI*DC*(TB-TIW)*1.0D-03
QHTAIR = MHTAIR*CPHT*(TP-TN0T)*1.0D-03
QH20 = HJACK*PI*DA*BOILC*( (TOW-373.0)**3 )*1.0D-03
QIN = QIN*1.0D-03
QCWALL = HB*(TB-TIW)*HJACK*PI*DC*1.0D-03
QWALLI = HA*(TIW-TP)*HJACK*PI*DC*1.0D-03
RATIOC = UAIR/UMFC
RATIOA = UQVERUMF
UABEDS = 1000.0D*(QH20+QHTAIR)/(TB-TP)
AFRAT = CSCFM/QIN

ENDIF

999 CONTINUE

WRITE(6,501) TB,QIN,RATIOC,HB
WRITE(6,502) TIW,QH20,CSCFM,AFRAT
WRITE(6,503) TP, QCAIR, RATIOA, HA
WRITE(6,504) TOW, QHTAIR, ASCFM, UABEDS

WRITE(7,721) QIN,TB,RATIOA,QCRAD,QCAIR,QCWALL
WRITE(7,722) QWALLI, QARADI, QHTAIR, QWALLO, QARADO
WRITE(7,723) QH20
WRITE(7,724)

WRITE(8,*) UAIR
WRITE(8,*) UMFC
WRITE(8,*) UBC
WRITE(8,*) DBC
WRITE(8,*) TN0T
WRITE(8,*) TB
WRITE(8,*) EBC
WRITE(8,*) EMFC
WRITE(8,*) HEIGHTC
WRITE(8,*) QIN

PRINT*, 'WANT TO RUN AGAIN? 1 FOR YES, 2 FOR NO'
READ(5,*) IQUES
IF(IQUES.EQ.1) GOTO 5

504 FORMAT(' ',2X,'TOW = ',F7.1, ' QHTAIR = ',F7.2, ' ASCFM = ',F7.2, ' UABEDS = ',F8.3)

721 FORMAT('O',X,' QIN TB RATIO QCRAD QCAIR QCWALL', /,X,F6.2,F9.1,F6.2,3(2X,F6.2) )
722 FORMAT('O',X,' QHTCONVI QRADI QHTAIR QHTCONVO QRADO ', /,3X,5(F6.2,3X) )
723 FORMAT('O',X,'QH2O = ',F6.2)
724 FORMAT(' ',X,'**********************************',/)

STOP
END

DOUBLE PRECISION T,RHO,CP,MU,K,PR
INTEGER IFLAG

DOUBLE PRECISION TEMP(25),DENS(25),HEAT(25),VISC(25),COND(25)
DOUBLE PRECISION PRAN(25),X

INTEGER I
IFLAG = 0

DATA (TEMP(I),I=1,22) / 200.0D0, 250.0D0, 300.0D0, 350.0D0,
* 400.0D0, 450.0D0, 500.0D0, 550.0D0, 600.0D0, 650.0D0,
* 700.0D0, 750.0D0, 800.0D0, 850.0D0, 900.0D0, 950.0D0,
* 1000.0D0, 1100.0D0, 1200.0D0, 1300.0D0, 1400.0D0, 1500.0D0
/

DATA (DENS(I),I=1,22) / 1.7458D0, 1.3947D0, 1.1614D0, .9950D0,
* .8711D0, .7740D0, .6964D0, .6329D0, .5804D0, .5356D0,
* .4975D0, .4643D0, .4354D0, .4097D0, .3868D0, .3666D0,
* .3482D0, .3166D0, .2902D0, .2679D0, .2488D0, .2322D0 /

DATA (HEAT(I),I=1,22) / 1007.0D0, 1006.0D0, 1007.0D0, 1009.0D0,
* 1014.0D0, 1021.0D0, 1030.0D0, 1040.0D0, 1051.0D0, 1063.0D0,
* 1075.0D0, 1087.0D0, 1099.0D0, 1110.0D0, 1121.0D0, 1131.0D0,
* 1141.0D0, 1159.0D0, 1175.0D0, 1189.0D0, 1207.0D0, 1230.0D0 /

DATA (VISC(I),I=1,22) / 132.5D-7, 159.6D-7, 184.6D-7, 208.2D-7,
* 230.1D-7, 250.7D-7, 270.1D-7, 288.4D-7, 305.4D-7, 322.5D-7,
* 338.8D-7, 354.6D-7, 369.8D-7, 384.3D-7, 398.1D-7, 411.3D-7,
* 424.4D-7, 449.0D-7, 473.0D-7, 496.0D-7, 530.0D-7, 557.0D-7 /
DATA (COND(I), I=1,22)/ 18.1D-3, 22.3D-3, 26.3D-3, 30.0D-3, 33.8D-3, 37.3D-3, 40.7D-3, 43.9D-3, 46.9D-3, 49.7D-3, 52.4D-3, 54.9D-3, 57.3D-3, 59.6D-3, 62.0D-3, 64.3D-3, 66.7D-3, 71.5D-3, 76.3D-3, 82.0D-3, 92.0D-3, 100.D-3 /

DATA (PRAN(I), I=1,22)/ .737DO, .720DO, .707DO, .700DO, .690DO, .686DO, .684DO, .683DO, .685DO, .690DO, .695DO, .702DO, .709DO, .716DO, .720DO, .723DO, .726DO, .728DO, .728DO, .719DO, .703DO, .685DO /

IF (T.LE.TEMP(1)) THEN

    IFLAG = 1
    RHO = DENS(1)
    CP = HEAT(1)
    MU = VISC(1)
    K = COND(1)
    PR = PRAN(1)
    RETURN

ELSE

    IF (T.GE.TEMP(22)) THEN

        IFLAG = 2
        RHO = DENS(22)
        CP = HEAT(22)
        MU = VISC(22)
        K = COND(22)
        PR = PRAN(22)
        RETURN

    ELSE

        CONTINUE

    ENDIF

ENDIF

ENDIF

DO 100 I=1,21
IF(T.GE.TEMP(I).AND.T.LT.TEMP(I+1)) J=I

100 CONTINUE

X = ( T-TEMP(J) ) / ( TEMP(J+1)-TEMP(J) )

CP = ( HEAT(J+1)-HEAT(J) )*X + HEAT(J)
RHO = ( DENS(J+1)-DENS(J) )*X + DENS(J)
MU = ( VISC(J+1)-VISC(J) )*X + VISC(J)
K = ( COND(J+1)-COND(J) )*X + COND(J)
PR = ( PRAN(J+1)-PRAN(J) )*X + PRAN(J)

RETURN
END
SUBROUTINE MINFLU ( U, RHOAIR, CPAIR, KAIR, MUAIR, EO, DP, 
* 
    RHOS, HEIGHT, PI, AR, UMF, KEO, HGC )

C This routine evaluates the minimum fluidization velocity of a 
C bed of particles given the particle and gas stream properties. 
C It also calculates bed height, Archimedes number, conductivity 
C of the fluidized bed, and the convection coefficient from gas 
C to container wall. 

DOUBLE PRECISION U, RHOAIR, CPAIR, KAIR, MUAIR, EO, DP 
DOUBLE PRECISION RHOS, HEIGHT, PI, AR, UMF, KEO, HGC 
DOUBLE PRECISION G, KS, M, KMF, HF, HAV

G = 9.806D0
KS = 1.90D0
M = 6.0D0
AR = RHOAIR*(RHOS-RHOAIR)*G*(DP**3)/(MUAIR*MUAIR)
UMF = ( 25.7D0*MUAIR/(RHOAIR*DP) )* (DSQRT(1.0D0+ 
*      5.53D-05*AR)) - 1.0D0 )
HF = M*KAIR/DP
KEO = KAIR*( (KS/KAIR)**( 0.28D0-0.757D0*DLOG10(EO)- 
*          0.057D0*DLOG10(KS/KAIR) ) )
KMF = KEO + 0.10D0*RHOAIR*CPAIR*DP*UMF
HAV = DSQRT( 4.0D0*KMF*RHOAIR*CPAIR*U/(PI*HEIGHT) )
IF (U.EQ.0.0D0) THEN
    HGC = HF
ELSE
    HGC = 1.0D0/( (1.0D0/HAV) + (1.0D0/HF) )
ENDIF

RETURN
END
This subroutine evaluates the annular bed convective heat transfer coefficient using data from Pflum and Brown. Particle diameter and the velocity ratio \( u/umf \) is input and the convection coefficient is output. FLAGDP is set to 99 if DP is improperly specified. Last update: June 23, 1989 by Jim Foley.

```
SUBROUTINE COEFFA ( DP,UOVERUMF,HGC,HTOT,FLAGDP )

DOUBLE PRECISION DP,UOVERUMF,HGC,HTOT,X
INTEGER FLAGDP,J

FLAGDP=0
J=0
X = UOVERUMF

IF (DP.GT.240.0D-06.AND.DP.LT.260.0D-06) J=1
IF (DP.GT.490.0D-06.AND.DP.LT.520.0D-06) J=2

IF (J.EQ.1) THEN ! 250 MICRON PARTICLES

   IF (X.GE.1.00D0.AND.X.LT.1.34D0) THEN
       HPC = 90.30D0
   ELSE

       IF (X.GE.1.34D0.AND.X.LT.2.24D0) THEN
           HPC = 94.11D0*(X-1.34D0) + 90.3D0
       ELSE

           IF (X.GE.2.24D0.AND.X.LT.4.90D0) THEN
               HPC = -479.6D0 + 365.85D0*X - 32.864D0*X*X
           ELSE

```
IF (X.GE.4.90D0.AND.X.LT.7.58D0) THEN

HPC = -43.25D0 + 179.51D0*X - 13.009D0*X*X

ELSE

IF (X.GE.7.58DO.AND.X.LT.9.32) THEN

HPC = -9.195D0*(X-7.58) + 570.0D0

ELSE ! X > 9.32

HPC = 554.0D0

ENDIF

ENDIF

ENDIF

ENDIF

ELSE

IF (J.EQ.2) THEN ! 500 MICRON PARTICLES

IF(X.GE.1.00D0.AND.X.LT.1.36D0) THEN

HPC = 290.22D0 - 442.1D0*X + 292.517D0*X*X

ELSE

IF (X.GE.1.36D0.AND.X.LT.2.11D0) THEN

HPC = - 867.47D0 + 1173.25D0*X - 269.33D0*X*X

ELSE

ENDIF

ENDIF

ENDIF

ENDIF
IF (X .GE. 2.11D0 .AND. X .LT. 3.18D0) THEN

    HPC = 102.0D0 + 263.125D0*X - 55.747D0*X*X

ELSE  ! X > 3.18

    HPC = 375.0D0

ENDIF

ENDIF
ENDIF
ELSE  ! DP IMPROPERLY SPECIFIED

    FLAGDP = 99
    HPC = -999.0D0

ENDIF

ENDIF

HTOT = HGC + HPC

RETURN
END
SUBROUTINE COEFFC ( UAIR, CPAIR, RHOAIR, RMUAIR, RKAIR,
*   DP, CPS, RHOS, EO, EMF, ZDEAD, WIDTH, AORIF,
*   UMF, E, EB, UB, DB, HEIGHT, HGC, HPC, HTOT )

C--
C
C THIS ROUTINE WILL EVALUATE THE HEAT TRANSFER COEFFICIENTS FOR
C A FLUIDIZED BED FOLLOWING THE MODEL OF XAVIER AND DAVIDSON.
C
C CALL PARAMETERS:
C
C   UAIR     FLUIDIZING AIR SUPERFICIAL VELOCITY ( m/s )
C   CPAIR    AIR SPECIFIC HEAT ( J/kgK )
C   RHOAIR   AIR DENSITY ( kg/m3 )
C   RMUAIR   AIR DYNAMIC VISCOSITY ( Ns/m2 )
C   RKAIR    AIR THERMAL CONDUCTIVITY ( W/mK )
C   DP       PARTICLE DIAMETER ( m )
C   CPS      PARTICLE SPECIFIC HEIGHT ( m )
C   RHOS     PARTICLE DENSITY ( kg/m3 )
C   EO       BED VOIDAGE OF PACKED BED
C   EMF      BED VOIDAGE AT MINIMUM FLUIDIZATION
C   ZDEAD    PACKED BED HEIGHT ( m )
C   WIDTH    CHARACTERISTIC BED WIDTH ( m )
C   AORIF    DISTRIBUTOR AREA PER ORIFICE ( m2 )
C
C OUTPUT PARAMETERS:
C
C   UMF      SUPERFICIAL AIR VELOCITY AT MIN. FLUIDIZATION
( m/s )
C   E        BED VOIDAGE OF FLUIDIZED BED
C   EB       VOIDAGE CONTAINED IN BUBBLES
C   UB       BUBBLE RISE VELOCITY ( m/s )
C DB AVERAGE BUBBLE DIAMETER (m)
C HEIGHT FLUIDIZED BED HEIGHT (m)
C HGC GAS-CONVECTION HEAT TRANS. COEFF. (W/m2K)
C HPC PARTICLE-CONVECTION HEAT TRANS. COEFF. (W/m2K)
C HTOT TOTAL HEAT TRANSFER COEFF. (W/m2K)

FIRST VERSION BY TIM PEDERSEN,
REVISED BY JIM FOLEY, 6/20/89.

IMPLICIT REAL*8 (A-H,O-Z)
REAL*8 KMF,KEO,KS

PI = 4.0D0*DATAN(1.0D0)
G = 9.806D0
RM = 6.0D0
KS = 1.90D0

C PARTICLE REYNOLDS NUMBER
REP = UAIR*RHOAIR*DP/RMUAIR

ZMF = ZDEAD*(1.0D0-EO)/(1.0D0-EMF)
RHOCMF = RHOS*(1.0D0-EMF)*CPS

AR = RHOAIR*(RHOAIR*RHOAIR)*G*(DP**3)/(RMUAIR*RMUAIR)
UMF = (RMUAIR/(RHOAIR*DP))*(((1.0D0+5.53D-05*AR)**(0.5)-1.0D0)*
*25.7D0)

KEO = RKAIR*(KS/RKAIR)**(0.28D0-0.757D0*DLOG10(EO)-0.057D0*
* DLOG10(KS/RKAIR))
KMF = KEO+0.1DO*RHOAIR*CPAIR*DP*UMF

DB = (0.54D0*(UAIR-UMF)**(0.4DO)*(ZDEAD+4.0DO*DSQRT(AORIF))
* **(0.8DO))/G**(0.2DO)
PROD = DB/WIDTH
BUBBLE RISE VELOCITY:

IF PROD > 0.6 SLUG FLOW

IF .125 < PROD < 0.6 BUBBLE FLOW WITH WALL EFFECTS

IF .125 > PROD BUBBLE FLOW

UB = 0.71DO*DSQRT(G*DB)

IF( PROD .GE. 0.125DO .AND. PROD .LE. 0.6DO )
* UB = 1.13DO*0.71DO*DSQRT(G*DB)*DEXP(-PROD)

IF( PROD .GT. 0.6DO ) UB = 0.35*DSQRT(G*WIDTH)

QUANT = UB/(UAIR-UMF+UB)

EB = 1.0DO - QUANT

E = 1.0DO - (1.0DO-EB)*(1.0DO-EO)*ZDEAD/ZMF

HEIGHT = ZMF/QUANT

HP = 2.0DO*DSQRT(2.0DO*KMF*RHOCMF*(UAIR-UMF)/(PI*DB) )

HF = RM*RKAIR/DP

HPC = (1.0DO/(1.0DO/HP+1.0DO/HF))*QUANT

U=UMF

HAV=DSQRT(4.0DO*KMF*RHOAIR*CPAIR*U/(PI*HEIGHT))

HGC = 1.0DO/(1.0DO/HAV+1.0DO/HF)

HTOT = HPC + HGC

RETURN

END
The two-phase theory of fluidization treats the combustion bed as being composed of a particle-dense phase (emulsion phase) and a particle-lean phase (bubble phase). It is assumed that all air flow in excess of minimum fluidization passes through the bubble phase.

The volume of bubble phase is the product of bed volume and bed voidage due to bubbles: $V_{bgas} = \epsilon_b H_c A_c$. The volume of emulsion phase is then $V_e = (1 - \epsilon_b) H_c A_c$. The volume of gas in the emulsion phase is $V_{egas} = \epsilon_m (1 - \epsilon_b) H_c A_c$. Bubble phase carbon monoxide and oxygen concentrations are based on bubble volume, and emulsion phase concentrations are based on emulsion air volume.

**Emulsion phase**

Conservation of species $A$ (either oxygen or carbon monoxide) in the emulsion phase, using the entire emulsion phase as the control volume is:
$$\frac{d}{dt} \left[ \epsilon_m f (1 - \epsilon_b) H_e A_e C_{A_e} \right]$$

where $C_{A_e}$ is the concentration of species A in the emulsion phase ($\text{moles/m}^3$).

$$u_m f \left( \frac{T_o}{T_c} \right) A_e C_{A_e} \epsilon_m f (1 - \epsilon_b)$$

where $u_m f \left( \frac{T_o}{T_c} \right)$ is the velocity of minimum fluidization air referenced to the air inlet temperature, and $C_{A_o}$ is the inlet gas flow concentration of species A.
Diffusion of $A$ from emulsion to bubble phase

$\frac{1}{H_c} \left[ u_m f C_{Ae} - u_m f \left( \frac{T_o}{T_c} \right) C_{Ao} \right] = -\frac{k_q \alpha_b \varepsilon_b (C_{Ae} - \bar{C}_{Ab})}{\varepsilon_m f (1 - \varepsilon_b)} \left( \frac{-k_q \alpha_b \varepsilon_b (C_{Ae} - \bar{C}_{Ab})}{\varepsilon_m f (1 - \varepsilon_b)} \right)$

where $k_q$ is the interphase mass transfer coefficient ($m/s$), $\alpha_b$ is the bubble area-to-volume ratio ($m^2/m^3$), and $C_{Ab}$ is the height-average concentration of species $A$ in the bubble phase.
Bubble phase

\[
\text{Rate of change of species A in bubble phase} = \begin{bmatrix}
\text{Flow of A into bubble phase} \\
\text{Flow of A out of bubble phase}
\end{bmatrix} - \begin{bmatrix}
\text{Diffusion of A from bubble to emulsion phase} \\
\text{Production of A within bubble phase}
\end{bmatrix}
\]

The concentration of a species of gas in the bubble phase varies with height. Therefore, a semi-infinite control volume with cross-sectional area \( A_c \), and height \( dz \), is used for the bubble phase conservation equation.

\[
\text{Rate of change of species A in bubble phase} = \frac{d}{dt} \left[ (e_b A_c C_{A_b}) dz \right]
\]

\[
\text{Flow of A into bubble phase} = (u - u_m f) e_b C_{A_b} A_c
\]
Flow of A out of bubble phase

\[ \text{Flow of A out of bubble phase} = (u - u_{mf}) \left( C'_{A_b} + \frac{\partial C'_{A_{eb}}}{\partial z} dz \right) \varepsilon_{eb} A_c \]

Diffusion of A from bubble to emulsion phase

\[ \text{Diffusion of A from bubble to emulsion phase} = k q a_b \varepsilon_b \left( C' A_b - C' A_e \right) A_c dz \]

Production of A within bubble phase

\[ \text{Production of A within bubble phase} = \frac{\dot{G}_{A_b}}{H_c} dz \]

The steady state bubble species differential equation then becomes:

\[ \frac{d}{dz} C'_{A_b} = \frac{-k q a_b \varepsilon_b (C A_b - C A_e) + \frac{\dot{G}_{A_b}}{H_c A_c}}{\varepsilon_b (u - u_{mf})} \]

(11.2)

The initial condition is the concentration of species A at \( z = 0 \):

\[ C'_{A_b} \bigg|_{z=0} = C' A_0 \]

(11.3)
This program solves the oxygen and carbon monoxide conservation equations for the two-bed fluidized combustor. It utilizes the IMSL routine DIVPAG to solve the two first-order differential equations for bubble phase oxygen and carbon monoxide concentration. The subroutine FCN evaluates the two derivatives. The subroutine FCNJ is never called (it is needed by DIVPAG in case the Jacobian needed to be evaluated).

Last update: June 27, 1989 by Jim Foley.

EXTERNAL DIVPAG, FCN, FCNJ
INTEGER NEQ, NPARAM, IMETH, MXSTEP, IDO, PRFLAG, IFLAG
PARAMETER (NEQ=2, NPARAM=50)
DOUBLE PRECISION U, UMF, UB, DB, TIN, TBED, AB, EB, EMF, KQ, H
DOUBLE PRECISION COB, COE, O2B, O2E, QIN, MDOTDC, PI, DIFFUS
DOUBLE PRECISION H2OB, AREA, HINIT, ZSTEP, TRAT
DOUBLE PRECISION TOL, O2E0LD, COE0LD, PERCENT, KA, KB, KC, K1
DOUBLE PRECISION RATE, LOSS1, LOSS2, LOSS3, LHV, D, SH, NUM
DOUBLE PRECISION YB(2), YPRIME(2), Z, MDOT0LD, TEST1, TEST2
DOUBLE PRECISION TEST3, ZEND, Y10LD, Y20LD, SUM1, SUM2, QUEST
DOUBLE PRECISION A(1,1), DYDPY(1,1), PARAM(NPARAM)
DOUBLE PRECISION A1, A2, A3, A4, A5, A6, A7, A8, A9, A10, O2IN
DOUBLE PRECISION B1, B2, B3, B4, B5
CHARACTER*20 CASE

COMMON/BLK1/B1, B4, A3, A4, O2E, COE

READ(8,*) U
READ(8,*) UMF
READ(8,*) UB
READ(8,*) DB
READ(8,*) TIN
READ(8,*) TBED
READ(8,*) EB
READ(8,*) EMF
READ(8,*) H
READ(8,*) QJUNK

WRITE(9,*)'ECHO PRINT: U,UMF,UB,DB,TIN,TBED,EB,EMF,H,QIN'
WRITE(9,*)U,UMF,UB,DB,TIN,TBED,EB,EMF,H,QIN

PRINT*, 'WHICH COMBUSTION CASE? '
READ(5,1) CASE
WRITE(9,*) 'FOR COMBUSTION CASE: ',CASE
1 FORMAT(A20)

PRINT*, 'WHAT IS QIN (kW)?'
READ(5,*) QIN
WRITE(9,*)'QIN = ',QIN
QIN = QIN*1.0D+03
D = 0.0040D0

C ***** INITIALIZE VARIABLES **************

SH = 1.420D0
PI = 4.00D0*DATAN(1.0D0)
NUM = 0.080D0*6.00D0/(PI*D*D*720.0D0)
H20B = 0.0010D0
DIFFUS = 2.08D-04
AREA = 0.0324D0
AB = 6.00D0/DB
LHV = 2.8891D+07

KQ = UMF/3.0D0 + DSQRT( 4.00D0*DIFFUS*EMF*UB/(PI*DB) )
K1 = 1.30D+14
KA = 2.23900D0*DEXP( -941.10D0/TBED )
KB = 0.20890D0*DEXP( 7654.70D0/TBED )
KC = 0.05018DO*DEXP( 59561.0DO/TBED )
RATE = K1*(KA**0.50D0) / ( (KB**0.50D0) *(KC**0.25D0) )

C Make initial guesses for fuel feed rate and concentrations
MDOTDC = 2.0DO*QIN*3600.0DO/LHV
YB(1) = 0.00DO
YB(2) = 0.008705D0
O2IN = YB(2)
C0E = 0.0010D0
O2E = 0.0003280D0

A1 = UMF/H
A2 = TIN*O2IN/TBED
A3 = KQ*AB*EB
A4 = EMF*(1.0DO-EB)
A5 = 1.0DO/(H*AREA)
A6 = 2.0DO*PI*NUM*SH*DIFFUS*D
write(9,*),'A1,A2,A3,A4,A5,A6'
write(9,*),A1,A2,A3,A4,A5,A6

B1 = U - UMF
B2 = -3.162D-05*RATE
B3 = DSQRT(H20B)
B4 = 0.50D0*B2*EB*B3
write(9,*),'B1,B2,B3,B4'
write(9,*),B1,B2,B3,B4

TOL = 1.0D-06
HINIT = 1.0D-06
IMETH = 1
MXSTEP = 10000
PARAM(1) = HINIT
PARAM(12) = IMETH
PARAM(4) = MXSTEP
ZSTEP = H/50.0D0
TRAT = TIN/TBED
IDO = 1
Z = 0.0DO
IT = 0
IFLAG = 99
PRFLAG = 0

DO 222 WHILE (IFLAG.EQ.99)

C Perform integration with 50 calls to DIVPAG
Z = 0.0D0
IDO = 1
SUM1 = 0.0D0
SUM2 = 0.0D0
DO 30 I = 1,50
   Y10LD = YB(1)
   Y20LD = YB(2)
   ZEND = DFLGAT(I)*ZSTEP
   CALL DIVPAG (IDO,NEQ,FCN,FCNJ,A,Z,ZEND,TOL,PARAM,YB)
   SUM1 = SUM1 + 0.5D0*(YB(1)+Y10LD)*H/50.0D0
   SUM2 = SUM2 + 0.5D0*(YB(2)+Y20LD)*H/50.0D0
C IF(PRFLAG.EQ.1) write(6,* ) Z,YB(1),YB(2)
30 CONTINUE

C Last call to DIVPAG, release workspace
IDO = 3
CALL DIVPAG (IDO,NEQ,FCN,FCNJ,A,Z,ZEND,TOL,PARAM,YB)

C Calculate average bubble concentrations
CDB = SUM1/H
O2B = SUM2/H

C Calculate emulsion phase concentrations
A8 = 5.5196D-06*MDOTDC
A10 = 18.016D-06*MDOTDC
QUEST = A6*C0E + A7
C0E0LD = C0E
O2E0LD = O2E
IF (QUEST.LT.A8) THEN
   C0E = ( ( A3*C0B + A5*A9 + 2.0D0*A5*A6*O2E ) /
\!
\begin{align*}
\text{LOSS}_1 &= (11.032 \times 10^{-6} \times \text{MDQTDC} - A_6 \times \text{COE}) \times 3.9352 \times 10^8 \\
\text{ELSE} & \\
0_2E &= \left( \frac{A_1 \times A_2 \times A_4 + A_3 \times 0_2B - A_5 \times A_8}{(A_1 \times A_4 + A_3)} \right) \\
\text{CQE} &= \left( \frac{A_3 \times CQE + A_5 \times A_10}{(A_1 \times A_4 + A_3)} \right) \\
\text{LOSS}_1 &= 0.0D0 \\
\end{align*}

\text{ENDIF}

\text{C Calculate percent burned in bed and coal feed rate}
\text{LOSS}_2 = \text{CQE} \times \text{UMF} \times \text{AREA} \times \text{EMF} \times (1-\text{EB}) \times 1.1053 \times 10^8 \\
\text{LOSS}_3 = \text{CQE} \times (\text{QIN} - \text{UMF}) \times \text{AREA} \times \text{EB} \times 1.1053 \times 10^8 \\
\text{MDQTOLD} = \text{MDQTDC} \\
\text{MDQTDC} = 3600.0D0 \times (\text{QIN} + \text{LOSS}_1 + \text{LOSS}_2 + \text{LOSS}_3) / \text{LHV} \\
\text{PERCENT} = \frac{\text{QIN}}{\text{QIN} + \text{LOSS}_1 + \text{LOSS}_2 + \text{LOSS}_3}

\text{C Test for convergence}
\text{TEST}_1 = \text{DABS}(\text{CQEOLD} - \text{CQE}) / \text{CQE} \\
\text{TEST}_2 = \text{DABS}(\text{CQEOOLD} - \text{CQE}) / \text{CQE} \\
\text{TEST}_3 = \text{DABS}(\text{MDQTOLD} - \text{MDQTDC}) / \text{MDQTDC} \\
\text{IF (PRFLAG.EQ.1) IFLAG = 0} \\
\text{PRFLAG} = 1 \\
\text{IF (TEST}_1 > 1.0D-04) \text{PRFLAG} = 0 \\
\text{IF (TEST}_2 > 1.0D-04) \text{PRFLAG} = 0 \\
\text{IF (TEST}_3 > 1.0D-04) \text{PRFLAG} = 0 \\
\text{IT} = IT + 1

\text{C Don’t exceed 150 iterations}
\text{IF (IT.GT.150) IFLAG = 0}

\text{222 CONTINUE}

\text{WRITE}(9,*) 'COE,02E,COB,02B' \\
\text{WRITE}(9,*) COE,02E,COB,02B \\
\text{WRITE}(9,*) 'ITERATIONS = ',IT \\
\text{WRITE}(9,*) '' \\
\text{WRITE}(9,*) 'LOSS1,LOSS2,LOSS3'
WRITE(9,*) 'LOSS1,LOSS2,LOSS3
WRITE(9,*) ' ,
WRITE(9,*) 'MDOTDC,PERCENT'
WRITE(9,*) MDOTDC,PERCENT

END

C ********************************************************
SUBROUTINE FCN (NEQ,Z,YB,YPRIME)
C ********************************************************

This subroutine evaluates the derivatives of YB

INTEGER NEQ
DOUBLE PRECISION Z,YB(2),YPRIME(2)
DOUBLE PRECISION A3,A4,B1,B4,02E,COE

COMMON/BLK1/B1,B4,A3,A4,02E,COE

YPRIME(1) = ( - A3*(YB(1)-(COE)) + 2.0D0*B4*YB(1)*
              ( DABS(YB(2))**0.25D0 ) ) / (B1*A4)
YPRIME(2) = ( - A3*(YB(2)-(02E)) + B4*YB(1)*
              ( DABS(YB(2))**0.25D0 ) ) / (B1*A4)

RETURN
END

C ********************************************************
SUBROUTINE FCNJ (NEQ,X,Y,DYDPY)
C ********************************************************

This subroutine is never called

INTEGER NEQ
DOUBLE PRECISION X,Y(2),DYDPY(*)
RETURN
END
13 APPENDIX D. COMBUSTION MASS BALANCES

Fuel composition

\[
\begin{align*}
C &= 0.7788 \\
H &= 0.0376 \\
N &= 0.0149 \\
S &= 0.0300 \\
CaO &= 0.01149 \\
O &= 0.0257 \\
Ash &= 0.1015
\end{align*}
\]

\[
\begin{align*}
W_c &= \text{lbm of dry coal per lbm of fuel} \\
w_l &= \text{lbm of added limestone per lbm of fuel} \\
w_w &= \text{lbm of water (inherent + added) per lbm of fuel}
\end{align*}
\]

Gas analysis

\[
\begin{align*}
X_{CO} &= \text{mole fraction of CO in dry exhaust gas} \\
X_{CO_2} &= \text{mole fraction of CO}_2 \text{ in dry exhaust gas} \\
X_{O_2} &= \text{mole fraction of O}_2 \text{ in dry exhaust gas}
\end{align*}
\]
\[ X_{N_2} = \text{mole fraction of } N_2 \text{ in dry exhaust gas} \]
\[ = 1 - X_{CO} - X_{CO_2} - X_{O_2} - X_{SO_2} \]

\[ X_{SO_2} = \text{mole fraction of } SO_2 \text{ in dry exhaust gas} \]

**Carbon in fly-ash**

\[ Y_c = \text{lbm of carbon per lbm of dry fly-ash} \]

**Air flow rate**

\[ \text{SCFM} = \text{standard cubic feet per minute of air flow} \]

\[ = \text{primary + secondary - tertiary air flow} \]

**Calculations**

\[ A_1 = \frac{(W_c)(CaO)}{56.08} \quad (13.1) \]

\[ A_2 = \frac{(W_l)}{100.091} \quad (13.2) \]

\[ A_3 = \frac{(W_c)(S)}{32.06} \quad (13.3) \]

\[ A_4 = \frac{(W_c)}{32.0} + \frac{CaO}{112.16} + \frac{1.5W_l}{100.091} + \frac{0.5W_w}{18.016} \quad (13.4) \]

\[ A_5 = \frac{X_{CO}}{2.0} + X_{CO_2} - X_{O_2} - X_{SO_2} \quad (13.5) \]

\[ A_6 = \frac{N}{28.016} \quad (13.6) \]

\[ N_{H_2O} = \frac{(H)(W_c)}{2.016} + \frac{W_w}{18.016} \quad (13.7) \]

**Initial guess of carbon conversion**

\[ \eta = 1.0 - \frac{(Y_c)(Ash)}{C(1.0 - Y_c)} \quad (13.8) \]
The following equations are solved iteratively:

\[ \frac{N_{dg}}{12.011 + \frac{W_l}{100.991}} \left( \frac{1}{X_{CO} - X_{CO_2}} \right) \]  
\[ N_{CaSO_4} = A_1 - X_{SO_2} V_{dg} \]  
\[ N_{CaO} = A_1 + A_2 - N_{CaSO_4} \]  
\[ \frac{V_{O_2}}{3.76W_c} = \frac{N_{dg} x N_2 - W_c A_6}{N_{CaSO_4} - 56.08 N_{CaO} + (W_c)(A_{sh}) Y_c} \]  
\[ \eta = 1.0 - \frac{[136.14 N_{CaSO_4} - 56.08 N_{CaO} + (W_c)(A_{sh}) Y_c]}{C W_c (1.0 - Y_c)} \]

Then the excess air is:

\[ x = \frac{(V_{O_2} - 0.07477)(100)}{0.07477} \]

The mass flow rate of fuel is:

\[ \dot{m}_{dc} = \frac{(0.03276)SFIM}{V_{O_2} W_c} \]

\[ \frac{kg\text{NO}}{GJ} = \frac{(0.0010099)Y(\text{ppm NO})}{W_c} \]

\[ \frac{kg\text{S}}{GJ} = \frac{(0.0010792)Y(\text{ppm SO}_2)}{W_c} \]
EXAMPLE - for Run B-3 with briquettes

\[
\begin{align*}
W_c &= 0.7763 \\
W_l &= 0.1297 \\
W_w &= 0.0940 \\
Y_c &= 0.200 \\
X_{CO} &= 0.00024 \\
X_{CO_2} &= 0.1324 \\
X_{O_2} &= 0.0497 \\
X_{N_2} &= 0.8165 \\
X_{SO_2} &= 0.00117 \\
SCFM &= 19.92 + 5.38 - 5.0 = 30.30 \\
A_1 &= 0.0001591 \\
A_2 &= 0.001296 \\
A_3 &= 0.0007264 \\
A_4 &= 0.005256 \\
A_5 &= 0.19392 \\
A_6 &= 0.0005318 \\
Z &= 0.019696
\end{align*}
\]
Solve equations iteratively:

\[ N_{dg} = 0.360 \text{ moles dry gas per lbm fuel} \]
\[ N_{CaSO_4} = 0.000305 \text{ moles CaSO}_4 \text{ per lbm fuel} \]
\[ N_{CaO} = 0.00115 \text{ moles CaO per lbm fuel} \]
\[ \eta = 0.924 = \text{carbon conversion efficiency} \]
\[ N_{O_2} = 0.1006 \text{ moles O}_2 \text{ per lbm fuel} \]

Fuel mass flow rate:

\[ \dot{m}_f = 12.71 \text{ lbm fuel per hr} \]
\[ \dot{m}_{dc} = 9.87 \text{ lbm dry coal per hr} \]

Excess air:

\[ x = 34.59\% \]

Emissions:

\[ \frac{kg \ NO}{GJ} = 0.160 \]
\[ \frac{kg \ S}{GJ} = 0.586 \]