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FLUIDIZED BED GASIFICATION OF SOME WESTERN CANADIAN COALS

by

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ABSTRACT

Three different Western Canadian coals were gasified with air and steam in a fluidized bed of 0.73 mm Ottawa sand and coal, at atmospheric pressure, and temperatures of 1023 - 1175 K to produce a low Btu gas. The coals tested were of two types: one non-caking and two caking coals. The results were compared with those previously obtained for the same three coals when gasified in essentially the same equipment, but operated as a spouted bed.

The effects of temperature, coal feed rate, air to coal ratio, steam to coal ratio, coal quality, coal particle size, and bed depth on the gas composition, gas calorific value and the operating stability of the gasifier, were established by running gasification tests over a wide range of operating conditions.

Typical calorific value of the gas obtained for all three coals was in the range of 2.0 - 2.6 MJ/m^3, which is lower than reported for the spouted bed and commercially available fluidized bed gasifiers. Analysis of the results suggested that in the present low temperature gasifier, the combustion and pyrolysis reactions predominate over the gasification reactions.

The ability to treat caking coals in fluidized bed and spouted bed reactors is discussed. It is concluded that the dispersion of coal in a bed of inert silica and ash, rather than hydrodynamic characteristics is the key factor in their success in handling caking coals.

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CHAPTER I - INTRODUCTION

MAJOR OBJECTIVE OF THE RESEARCH PROGRAM

The major objective of this research program was to study the gasification of some Western Canadian coals in a fluidized bed under conditions as close as possible to those under which spouted bed gasification of these coals had been previously investigated in this Department.(1)

A secondary objective was to elucidate whether the relative success in gasifying caking coals in a spouted bed was due to the high gas velocities in the spout or to a coal dilution effect.

BACKGROUND

The impending shortage of oil and natural gas compounded with the unreliability of continuous supply and the escalating prices of these premium fuels have prompted a massive research effort in coal conversion technology as a means of providing clean burning fuel and petrochemical feedstocks. The principal scene of this research development has been Europe and the United States, although it has been suggested (2,4) that Canada should also be involved in developing this technology. A very large number of processes have been investigated or proposed for the production of low, medium or high Btu gases. Inasmuch as several comprehensive reviews on the subject have been published (5-9), only a few salient facts will be pointed out here.

Low Btu gas is produced by the gasification of coal, char or even wood, with steam and air, generally at atmospheric pressure, and temperatures of about 1073-1400 K. The resulting gas is therefore diluted with nitrogen, which comprises 45 - 60% by volume of the dry gas, the rest being carbon monoxide, hydrogen, carbon dioxide and small amounts (up to 3%) of methane. Low Btu gases have a gross calorific value in the range of 3.73 - 9.31 MJ/std m^3(100-250 Btu/scf).

Medium Btu gas is produced by gasification of coal or coke with steam and oxygen at atmospheric or higher pressure, and temperatures of 755-1755K. The produced gas has little nitrogen (2%), carbon monoxide and hydrogen concentrations of 30-40%, up to 4% methane, and a gross calorific value of 9.31-20.49 MJ/std m^3 (250-550 Btu/scf). The chemistry involved in the production of medium and low Btu gas is essentially the same and will be analysed later in this chapter.

High Btu gas or synthetic natural gas is essentially methane and has a gross heating value of 35.40÷37.26 MJ/std m^3 (950-1000 Btu/scf). High Btu gases can be produced by two routes. The first starts from a medium Btu gas and involves shift conversion of water to hydrogen by carbon monoxide followed by methanation. The shift conversion (eq. (1))

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad \Delta H = -32.55 \text{ KJ/at.g C} \qquad (1)$$

is generally done at 588 - 700 K and 2750 kPa in the presence of a catalyst, in such a way that the $CO:H_2$ ratio is adjusted to 1:3 in preparation for the catalytic methanation step (eq.(2)). Carbon dioxide is removed by chemical absorption.

$$CO + 3H_2 \overset{Ni}{\rightleftharpoons} H_2O + CH_4 \qquad \Delta H = - 232.50 \text{ KJ/at.g C} \qquad (2)$$

The second route is the direct hydrogenation of coal or hydrogasification (eq. (3)).

$$C + 2H_2 \rightleftharpoons CH_4 \qquad \Delta H = - 91.46 \text{ KJ/at.g C} \qquad (3)$$

The reaction is carried out at pressures in excess of 3435 kPa.

From the coal gasification literature it is evident that although there are many processes that have been proposed, or are under development, only a few have been commercially proven. These are, with some variations, essentially four: The fixed bed Lurgi: (10-14), the entrained bed Koppers-Totzek(10,11,15) the fluidized bed Winkler (10,11,14), and the moving bed Wellman-Galusha gasifier (16). All of these reactors will yield a low Btu gas when operated with air, and a medium Btu gas when operated with oxygen. The Koppers-Totzek is only operated with oxygen. Comparison of the commercially available technology is provided in references 7, 8, 14, & 18.

From an inspection of the coal gasification literature, it also appears that the processes which are either available now or likely to become available in the near future are coal gasifiers producing low or medium Btu gas. Of these, many use a fluidized bed reactor e.g. the Winkler gasifier, the CO_2 acceptor process (Conoco Coal Development Co.)(10-12), the Synthane Process (U.S. Bureau of Mines, ERDA)(10-12,18), the U-gas process (Institute of Gas Technology)(10,11), the Agglomerating Ash Process (Batelle/Union Carbide)(12), the Westinghouse Process (10,19), the Hydrane Process (U.S. Bureau of Mines, ERDA)(10,20), and the Hygas Process (Institute of Gas Technology)(12,13). All of these processes except for the Winkler gasifier, are at the development stage, either on a pilot or demonstration scale.

In spite of a wealth of descriptive literature on fluidized bed gasifiers, very little data on actual operating conditions or quality of the gas produced is available. Table 1 summarizes such data for fluidized bed reactors producing low Btu gas. The available data show that for air blown fluidized bed reactors operating at atmospheric pressure, a gas quality of 3.71-4.40

TABLE 1: TYPICAL DATA FOR LOW BTU GAS FLUIDIZED BED COAL GASIFIERS

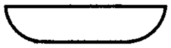
Variable		Bamag Winkler (21)	Davy-Power Gas Winkler (22)	Modified Synthane (18)	U Gas (5)	Westinghouse (19)
Type of Coal		Subbituminous A	Montana Subbituminous	Illinois No.6	Pittsburg coal	Indiana & Pittsburg bituminous
Coal size (mm)		15.9 - 0	9.53 - 0	0.84 - 0	-	0.8 - 1.2 (av.)
Coal feed rate (g/s) or (g/s x m^2)		135.6 (g/s m^2)		1.89 - 2.58 (g/s)	-	75.6 - 163.8 (g/s)
Pressure (kPa)		101	101	2020	2405	1546
Average temperature (K)			1366	1253 - 1292	811	1200 - 1294
Air flow std (m^3/s)				1.18-2.2 $\times 10^{-3}$		
Steam flow (g/s)		Nil		1.26-2.39		
Gas velocity at average temp (m/s)				0.07-0.16		0.37 - 0.61
Gas composition % by volume (dry basis)	H_2	11.7	13.1	21.5	13.3	
	CO	21.7	22.10	10.1	19.4	
	CH_4	0.7	0.84	5.6	4.7	
	CO_2	9.8	7.12	17.9	10.1	
	N_2	55.3	56.82	43.5	51.9	
	C_2H_6		-	0.7	-	
	H_2S	0.8	0.1	0.7	0.7	
Gas Gross heating value (MJ/std m^3)		3.91	4.40	6.15	5.71	3.73 - 4.47

MJ/std m^3 is expected, while operating at higher pressures yields a gas of a higher heating value, in the range of 4.47-6.15 MJ/std m^3.

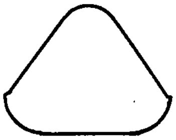
Advantages of fluid bed operation are excellent gas-solid contact, high heat transfer rates, relatively low residence times, high turndown capability, and uniform temperature and bed composition which provides good control. However, one of the main disadvantages of fluid bed gasifiers is their inability to treat caking coals without pretreatment.

When heated, caking coals first soften and flow like thermoplastics and then resolidify (cake) into a swollen porous solid (coke). The resolidification process is accompanied by emission of gas and condensable vapours. Two types of models for the caking process have been proposed: physical and physiochemical (23). In the physical models, caking is assumed to be a superposition of the physical phenomenom of melting and the chemical phenomenon of pyrolysis. In the physiochemical models, the softening and resolidification of coal is considered a consequence of pyrolysis which alters chemical bonds crosslinking polynuclear structures. Caking properties of coal have been related to their petrography and rank (24).

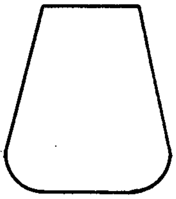
The caking properties of coal are customarily represented by their free swelling index (FSI), which is a measure of the increase in volume of a coal when it is heated without restriction under specified conditions. The standard method of test for FSI involves (25) heating several 1 g samples of coal to 1090 K within a specific time to prepare buttons of coke. The shape or profile of the coke buttons is standardized and numbered from 1 - 9. The sketch below illustrates these profiles for coal samples with FSI at 1,4 and 7.



FSI = 1



FSI = 4



FSI = 7

For a coal with a FSI of 7 the area of the profile is roughly 5 times that of a coal with a FSI = 1

When a caking coal is fed into a fluid bed gasifier, agglomeration of the coal occurs, particularly near the feed point. As well, sintering at the fluidization grid may occur. These two problems lead to defluidization and render smooth, continuous operation impossible. The caking problem is aggravated in a hydrogen atmosphere and with an increase in operating pressure (26). Since most of the coal in the Eastern States and a sizeable portion of that in Western Canada is of the caking type, efforts have been directed towards developing processes to accommodate these coals. Proposed solutions for avoiding defluidization in gasifiers have been reviewed in the literature (26). These include oxidative pretreatment of the coal in a separate fluidized vessel, alkaline pretreatment (27), entrained pretreatment (28), char dilution, good solid mixing in a turbulent or fast fluidized bed coupled with coal dispersion in dry ash, and mechanical stirring of the bed. However, all of these propositions imply the use of yet unproven technology, added hardware or added chemical costs. Another recent approach that was reported to have permitted the successful gasification of various caking coals (19) consisted of pneumatically introducing the coal into the fluidized bed reactor through a single orifice at the bottom.

Foong et al. at the University of British Columbia (1) noted that this and other proposed modifications tended to make the fluidized bed reactor similar to a spouted bed which they had been investigating for the gasification of two Western Canadian caking coals. Their results showed that these coals could successfully be gasified in a spouted bed of silica particles of the same size as the coal feed, albeit with some problems in feeding the coal into the reactor. However, a question arose from this work. Was this success due to the dilution effect of the silica and ash or to the ability of the high velocity gases in the spout to break-up any agglomeration? Or was it a combination of these two phenomena? Furthermore, since gas quality depends on coal properties, it is diffucult to assess a novel gasifier based on tests on a few coals which have not been gasified in more standard equipment.

The objective of this research was to elucidate these questions by processing the same coals in a fluidized bed, under operating conditions as similar as possible to the ones used in the spouted bed experiments. This work would also provide basic data on the gasification of Western Canadian coals in a fluidized bed.

THEORY OF COAL GASIFICATION

Coal gasification is the reaction of coal with air or oxygen and steam, or mixtures of these, to yield a gaseous product suitable as a fuel or as a petrochemical feedstock. This gaseous product is a mixture of carbon monoxide, carbon dioxide, hydrogen, methane, nitrogen, light hydrocarbons and sulfur bearing compounds, principally hydrogen sulfide.

The chemical reactions occurring in a coal gasifier can be divided in three main groups: combustion reactions, gasification reactions and pyrolysis reactions.

The combustion reactions are homogeneous or heterogeneous reactions with oxygen as shown below (Eq. (4) to (7)

$$C + \tfrac{1}{2} O_2 \rightleftharpoons CO \qquad (4)$$

$$CO + \tfrac{1}{2} O_2 \rightleftharpoons CO_2 \qquad (5)$$

$$C + O_2 \rightleftharpoons CO_2 \qquad (6)$$

$$H_2 + \tfrac{1}{2} O_2 \rightleftharpoons H_2O \qquad (7)$$

All of these reactions are exothermic (Table 2) and provide the heat necessary for some of the endothermic gasification reactions, as well as the heat necessary to bring the reactants to the reaction temperature. Overall the gasification system is autothermic.

The gasification reactions produce combustible gases from heterogeneous reactions between carbon and steam or gaseous products, or from homogeneous reactions among the gaseous products. The principal gasification reactions are as follows:

$$C + 2H_2 \rightleftharpoons CH_4 \qquad (3)$$

$$C + H_2O \rightleftharpoons CO + H_2 \qquad (8)$$

$$C + 2H_2O \rightleftharpoons CO_2 + 2H_2 \qquad (9)$$

$$C + CO_2 \rightleftharpoons 2\,CO \qquad (10)$$

These reactions are endothermic (Table 2) except reactions (9) and (3). In addition the homogeneous water-gas shift reaction will take place in a gasifier

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad (11)$$

At higher pressures the following reactions can also occur:

$$CO + 3H_2 \rightleftharpoons CH_4 + H_2O \qquad (12)$$

$$CO_2 + 4H_2 \rightleftharpoons CH_4 + 2H_2O \qquad (13)$$

Of the gasification reactions, the carbon steam reaction (eq. 8) is considered to be the most important and typical of coal gasification. Although reactions (3),(10),(12)&(13) produce a gas of a much greater heating value,

TABLE 2: HEATS OF REACTION AND EQUILIBRIUM CONSTANTS FOR MAIN GASIFICATION REACTIONS (Adapted from reference 8)

Reaction	Δ H (KJ/mol)		Kp	
	1100 K	1300 K	1100 K	1300 K
$C + \frac{1}{2} O_2 \rightleftharpoons CO$	-112.614	-113.882	8.8×10^9	1.31×10^9
$CO + \frac{1}{2} O_2 \rightleftharpoons CO_2$	-282.295	-281.416	7.21×10^8	6.29×10^6
$C + O_2 \rightleftharpoons CO_2$	-394.913	-395.298	6.35×10^{18}	8.25×10^{15}
$H_2 + \frac{1}{2} O_2 \rightleftharpoons H_2O$	-248.422	-249.685	7.60×10^8	1.15×10^7
$C + H_2O \rightleftharpoons CO + H_2$	135.807	135.636	2.62	1.14×10^2
$C + 2H_2O \rightleftharpoons CO_2 + 2H_2$	-146.492	-145.780	1.11×10^1	6.24×10^1
$C + CO_2 \rightleftharpoons 2\ CO$	169.685	167.534	1.22×10^1	2.08×10^2
$CO + H_2O \rightleftharpoons CO_2 + H_2$	-33.878	-31.898	1.10	0.55
$C + 2H_2 \rightleftharpoons CH_4$	-90.605	-91.735	3.68×10^{-2}	7.93×10^{-3}

their importance is limited for gasification at atmospheric pressure.

Pyrolysis refers to the endothermic thermal decomposition of coal into char and volatile compounds. The latter are represented by tar, light oil, water, hydrogen, carbon monoxide, carbon dioxide and light hydrocarbons (Eq. (14).

$$\text{Coal} \xrightarrow[\text{decomposition}]{\text{thermal}} CO + CO_2 + H_2 + H_2O + C_mH_n + \text{tar} + \text{light oil} + \text{char} \quad (14)$$

The volatile components can further react with water or hydrogen.

From the number of reactions involved it can be appreciated that a gasifier is a very complex chemical system with all reactions occurring simultaneously throughout the reactor or in localized areas (typically in a fixed bed). Sometimes it may even be of advantage to carry out some of these reactions in separate vessels. For instance, many proposed gasification processes carry out devolatilization of coal in one reactor followed by gasification of the char in a second reactor.

Table 2 lists the equilibrium constants for the important reactions at two typical gasification temperatures and atmospheric pressure. This data shows that from the thermodynamic point of view the combustion of carbon to carbon dioxide is by far the most favorable reaction. The rest of the combustion reactions follow, while the direct hydrogenation of carbon (reaction (3)) is the least favorable reaction. A temperature increase favors the carbon-steam reaction (Eq. (8)) and the Boudouard reaction (Eq. (10)) the most, while the importance of the rest of the combustion reactions decreases. Therefore, an increase in gasification temperature should doubly increase the concentration of combustible gases.

Although thermodynamics is a very useful tool in predicting the maximum concentrations obtainable, the over-riding consideration in a practical system is the kinetic of the reactions involved. Unfortunately, the kinetic of a coal gasification system are extremely difficult to analyse due to the number of reactions superimposed, effects of the coal and ash structure and complex fluid circulation patterns. However kinetic studies of the individual reactions involved can at least provide a comparison of the rates of the different reactions. These studies indicate that the fastest reaction in the gasification system is the oxidation of hydrogen to water.

$$H_2 + \tfrac{1}{2} O_2 \rightleftharpoons H_2O \qquad (7)$$

The rest of the oxidation reactions are also fast when compared with the gasification reactions (8) and (10).

$$C + H_2O \rightleftharpoons CO + H_2 \qquad (8)$$

$$C + CO_2 \rightleftharpoons 2\ CO \qquad (10)$$

while the slowest reaction is the direct hydrogeneration of carbon (Eq.(3), Table 3).

$$C + 2H_2 \rightleftharpoons CH_4 \qquad (3)$$

In summary, both thermodynamic and kinetic considerations indicate that in a coal gasification system at atmospheric pressure as long as any oxygen is present the combustion reactions are favoured over the gasification reactions of C-H_2O and C-CO_2 while the direct hydrogenation of carbon is the least favorable reaction.

EXPERIMENTAL PLAN

Given the objectives of this research, the experimental apparatus used for spouted bed gasification(1) was essentially conserved with only a few modifications and improvements. The spouted bed was transformed into a

TABLE 3: APPROXIMATE RELATIVE RATES OF THE GAS-CARBON REACTIONS AT 800°C AND 10.1 kPa PRESSURE (29)

Reaction	Relative Rates
$C - O_2$	1×10^5
$C - H_2O$	3
$C - CO_2$	1
$C - H_2$	3×10^{-3}

fluid bed by simply introducing a fluidization grid above the spouting orifice. Other modifications were related to solving the previously encountered problems in feeding caking coal into the bed, improving the method of obtaining mass balances by metering the total gas flow out of the reactor, and making a rough measurement of the amount of tars produced.

Since coal was to be fed into a fluidized bed of inerts, some room temperature experiments were carried out to assess the extent of segregation under fluidization conditions of different mixtures of coal and inert particles.

The gasification experiments in the fluidized bed were to utilize the same three Western Canadian coals used in the spouted bed experiments: One non-caking coal (Forestburg) and two caking coals (Sukunka and Coleman). The non-caking coal was used in a series of experiments to establish the effect of the various operating parameters on both the quality of operation and of the gas obtained. These parameters included:

- Bed temperature
- Coal feed rate
- Air to coal ratio
- Fluidizing velocity
- Steam to coal ratio
- Bed depth
- Particle size
- Presence of fines in feed.

Having established the effect of the different operating parameters, and the conditions under which best results could be obtained with non-caking coal, the gasification of caking coals in the fluidized bed was attempted under selected operating conditions.

During all gasification runs the following raw data were obtained in addition to the values of the operating parameters:

- Gas composition
- Total gas flow
- Ash production and characteristics
- Carbon bed content
- Tar production (selected runs).

CHAPTER II - EXPERIMENTAL APPARATUS

SEGREGATION STUDIES AND MINIMUM FLUIDIZATION VELOCITY

Prior to starting any gasification experiments, the minimum fluidization velocity and mixing patterns of coal/silica mixtures were studied at room temperature in the simple experimental set-up shown in Figure 1. This consisted of a 0.15 m (6") I.D. by 0.79 m (31") long glass column with a 60° brass conical base. The column was fitted with a water manometer for measuring the pressure drop across the bed and a calibrated rotameter for measuring the air flow. The fluidization grid was a perforated acrylic plate designed according to the procedure outlined by Kunii and Levenspiel (30). It consisted of a circular flange with an inner 0.15 m (6") diameter circle perforated with 61 holes of 2 mm diameter arranged in a triangular pitch (Appendix I).

GASIFICATION EXPERIMENTS

The coal gasification work was conducted in the small pilot plant schematically shown in Figure 2. Particulars of the main items of equipment are listed in Table 4. The fluid bed reactor and coal feeding arrangement are shown with more detail in Figure 3. The principal components of the pilot plant are described below. Numbers in brackets refer to legend in Figure 2.

The coal, crushed and screened to the desired particle size was loaded into the storage bin (1) maintained under a small nitrogen pressure. Coal flow was aided by a small nitrogen current. When working with caking coals a small scraper was fit flush with the feeder's rotating disc. Coal flowed then into a slanted 0.13 m (5") I.D. pipe and then into the fluidized bed (3), at a point near its top. The slanted feeding pipe was cooled by water flowing through a copper coil wrapped around it (Figure 3). When working

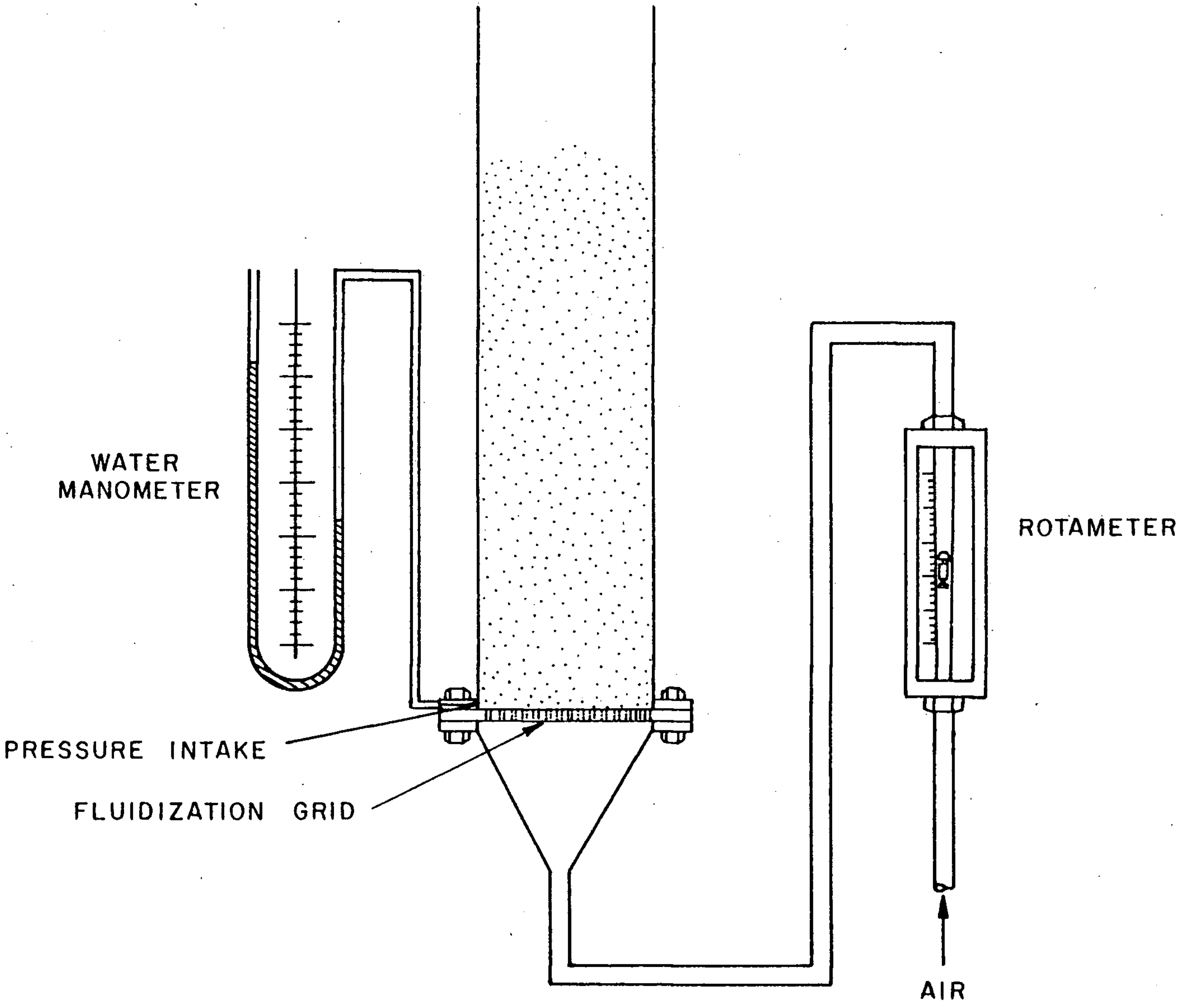

WATER MANOMETER
ROTAMETER
PRESSURE INTAKE
FLUIDIZATION GRID
AIR

FIGURE 1: FLUIDIZED BED FOR SEGREGATION STUDIES

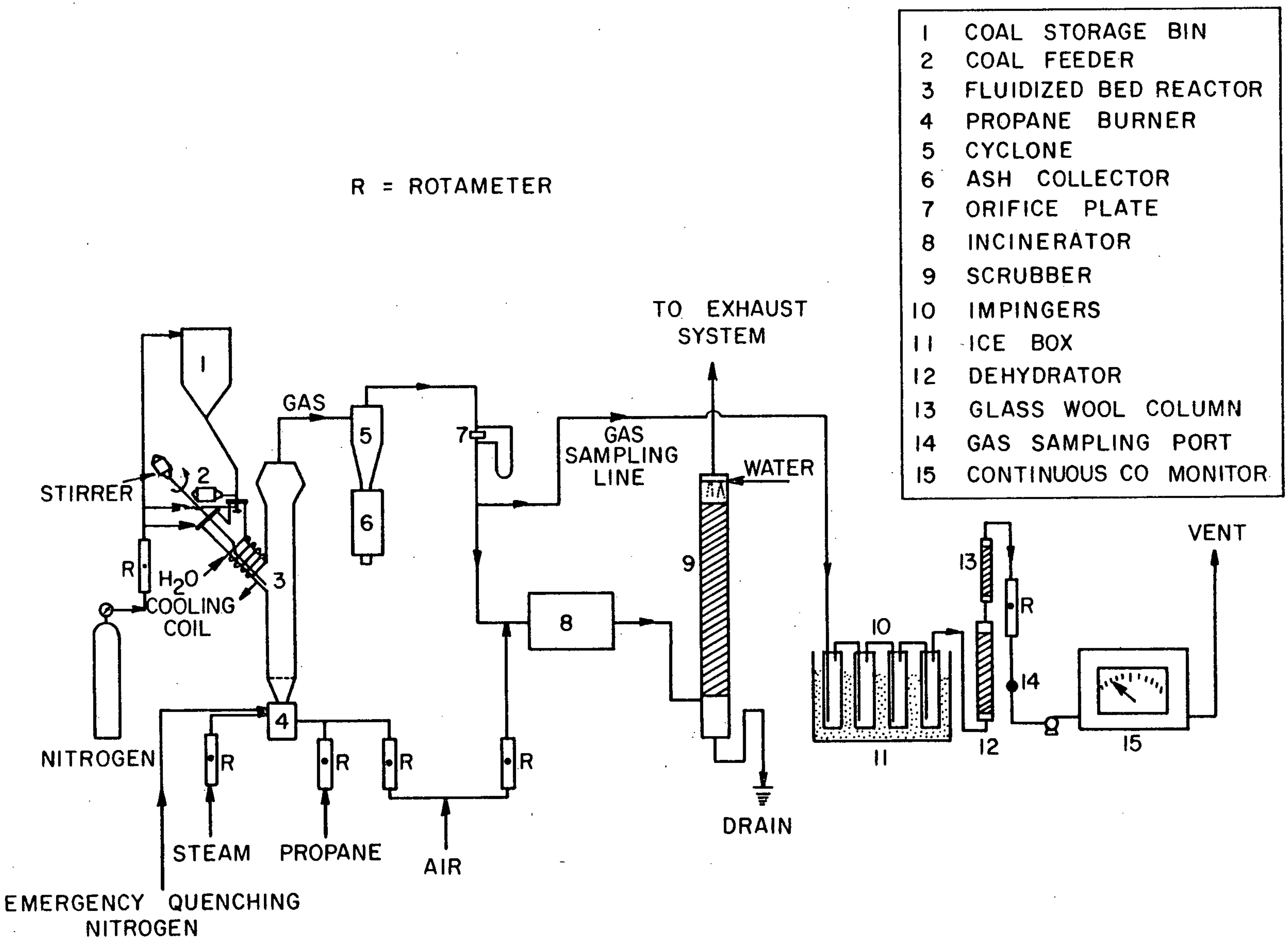

R = ROTAMETER
1 COAL STORAGE BIN
2 COAL FEEDER
3 FLUIDIZED BED REACTOR
4 PROPANE BURNER
5 CYCLONE
6 ASH COLLECTOR
7 ORIFICE PLATE
8 INCINERATOR
9 SCRUBBER
10 IMPINGERS
11 ICE BOX
12 DEHYDRATOR
13 GLASS WOOL COLUMN
14 GAS SAMPLING PORT
15 CONTINUOUS CO MONITOR
STIRRER
GAS
TO EXHAUST SYSTEM
GAS SAMPLING LINE
WATER
VENT
H2O COOLING COIL
R
NITROGEN
STEAM
PROPANE
AIR
DRAIN
EMERGENCY QUENCHING NITROGEN
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15

FIGURE 2: SCHEMATIC DIAGRAM OF FLUIDIZED BED PILOT PLANT

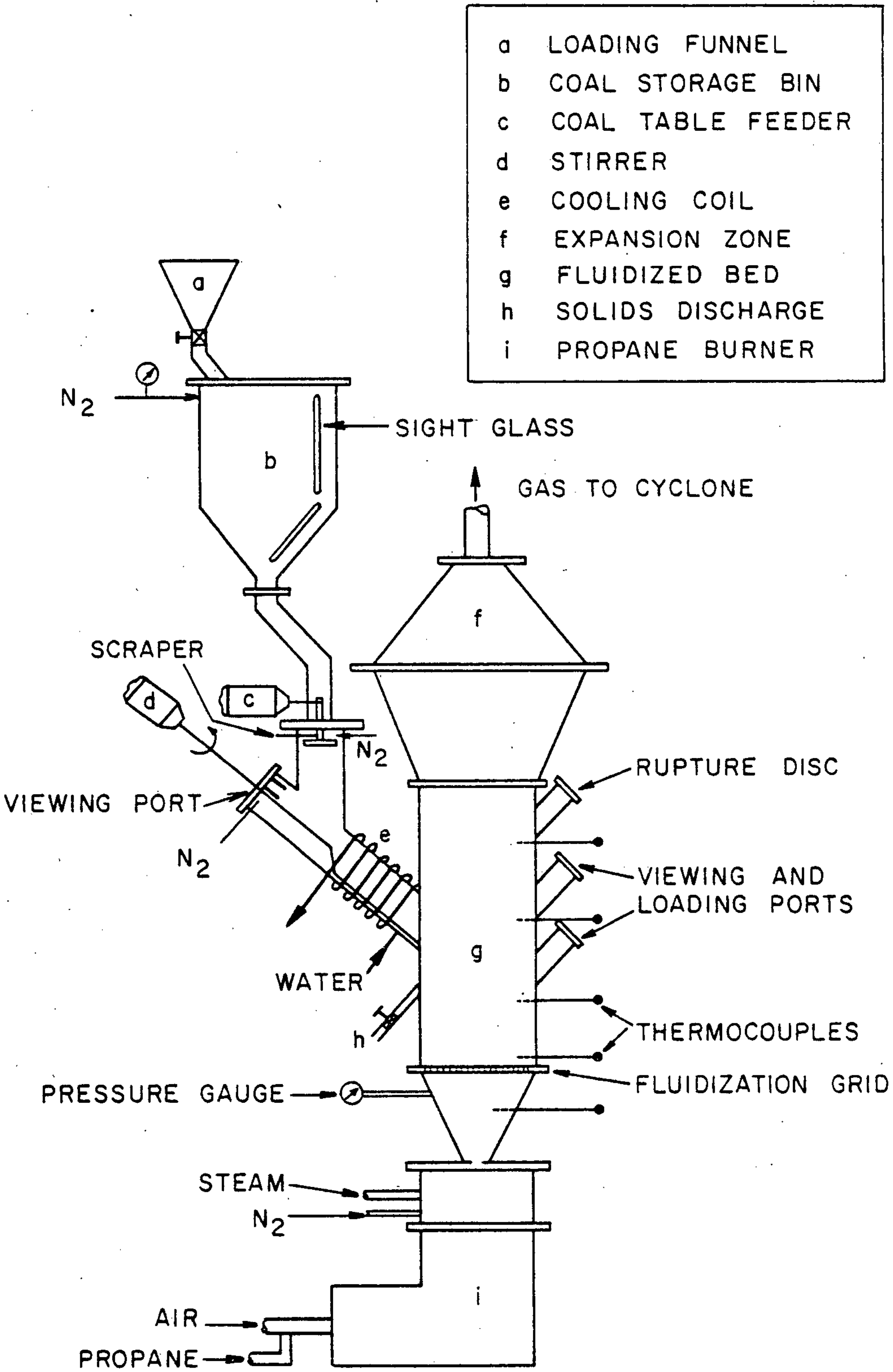

a LOADING FUNNEL
b COAL STORAGE BIN
c COAL TABLE FEEDER
d STIRRER
e COOLING COIL
f EXPANSION ZONE
g FLUIDIZED BED
h SOLIDS DISCHARGE
i PROPANE BURNER
a
N_2
b
SIGHT GLASS
GAS TO CYCLONE
f
SCRAPER
d
c
N_2
VIEWING PORT
N_2
e
RUPTURE DISC
VIEWING AND LOADING PORTS
g
WATER
h
THERMOCOUPLES
FLUIDIZATION GRID
PRESSURE GAUGE
STEAM
N_2
i
AIR
PROPANE

FIGURE 3: FLUIDIZED BED GASIFIER

TABLE 4: PARTICULARS OF MAJOR EQUIPMENT

Fluid Bed:

Diameter: .15 m
Height: .86 m
Cone angle: 60^{o}
Fluidization grid: perforated plate (Appendix I)
Material: column type 316L S.S.

Coal Feeder:

Type: Table feeder, variable speed
Capacity: 25 kg/h

Gas Burner for start-up:

Propane fired, Premix
Lining: refractory cement
Manufacturer: Eclipse Fuel Engineering Co.

Gas/ash Separation cyclone:

Diameter: 150 mm
Cylinder height: 500 mm
Cone height: 300 mm

Off-gas combustor:

Type: Electric furnace

Scrubber:

Type: water spray, counter current to gas flow
Diameter: 150 mm
Height: 1.8 m
Packing: 13 mm porcelain saddle or metallic lathe shavings

with caking coal a small variable speed stirrer was used to break up any agglomerations forming in the feeding pipe. The feeding pipe was also fitted with a viewing port kept clean by a small nitrogen flow. All nitrogen fed into the reactor was metered by a rotameter.

The fluidized bed (Figure 3) consisted of a main cylindrical section of 0.15 m (6") I.D. by 0.61 m (24") long type 316 L stainless steel with a wall thickness of 6.4 mm (¼"). A similar section of 0.25 m (10") long could be added to lengthen the reactor to a total fluid bed depth of 0.86 m (34"). The main section was equipped with two viewing ports which could also be used as loading ports, a valved 25 mm (1") solid discharge pipe, a rupture disc for safety, and chromel-alumel thermocouples spaced at 0.15 m. The bottom of the reactor was fitted with a fluidization grid. This consisted of a 6.4 mm (¼") thick stainless steel perforated plate of identical design to the acrylic fluidization grid used during the segregation studies (see Section 1 this Chapter). Even air distribution into the fluidization grid was achieved by a 60° conical section just below it.

The gasification reactor was also equipped with a premix propane burner for start-up (4). Air and propane flows into the refactory lined combustion chamber were metered by calibrated rotameters. Just above the burner and below the conical section a refractory lined mixing chamber was fitted. This mixing chamber was a short cylindrical section fitted with entrances for reacting steam and nitrogen for emergency quenching of the reactor.

Gas produced in the fluidized bed reactor flowed into an expansion zone for solid disengagement consisting of two cones joined at their base. The gas then flowed into a cyclone (5) (see Table 4) equipped with an ash collection chamber (6). Cleaned hot gas was metered by a calibrated orifice plate (7) (see Appendix II) before being incinerated in an electric furnace (8).

A small gas stream of approximately 80 cm^3/s by-passed the incinerator and was pumped by a vacuum pump through an ice-cooled train of four impingers (10) similar to the ones used for high volume stack sampling (31) (Appendix III). The impingers were used to cool the gas sample, to capture remaining solid particles and to condense any tars. The cleaned gas sample was then passed through a drierite desiccating column (12), a glass wool filled column (13), and pumped into a continuous CO analyser (15) which monitored the gasifier performance. A calibrated rotameter metered the gas. A sampling port (14) just before the entrance of the continuous CO monitor enabled a gas sample to be taken for complete analysis in a gas chromatograph (see Chapter III).

CHAPTER III - EXPERIMENTAL METHODS

MINIMUM FLUIDIZATION VELOCITY AND SEGREGATION STUDIES

The minimum fluidization velocity and solids mixing pattern of mixtures containing 0, 5, 15 and 20% coal in silica were studied at room temperature using the experimental set-up described in Chapter II. The silica particles used were either crushed and screened quartz of 2.3 mm nominal average size (1.18 mm - 3.36 mm) or commercially available 90% 0.73 mm (0.84-0.59 mm) Ottawa sand. The coal was crushed and screened Forestburg coal of nominal average size of 2.3 mm (1.18 mm - 3.36 mm). The experimental procedure for each sample was as follows.

A 4.54 kg (10 lbs) solids sample was prepared by thoroughly mixing approximate amounts of silica and coal particles. The mixture was then poured into the glass fluidization column. Random loose packing was obtained by gently fluidizing the solids in air and then turning the air off. The height of the column was measured. Starting from this point, the air flow into the column was increased stepwise. For each air flow (measured in the calibrated rotameter), pressure drop, column height and visual observations relating to fluidization state and mixing of coal and silica were recorded.

Plots of pressure drop versus superficial air velocity were prepared to establish the minimum fluidization velocity of the different mixtures, according to the method most generally accepted in the literature (30). The visual observations, on the other hand, were useful to establish the nature of fluidization, the extent of segregation, and the minimum superficial velocity at which segregation of coal and silica seemed to disappear.

COAL GASIFICATION EXPERIMENTS:

General Experimental Procedure

Coal crushed and screened to the desired particle size was loaded into the coal storage bin. Air was then turned on to a low flow to prevent silica particles from dropping into the combustion chamber of the propane burner while loading the inerts into reactor. One of the viewing ports of the reactor was used for charging the reactor with 0.73 mm Ottawa sand in amounts varying between 4.5 kg and 7.9 kg according to the desired bed depth. Next, water flow to the feed pipe cooling coil, the scrubber nitrogen to various parts of the coal feeding system and air to the electric furnace were turned on. Air flow into the gasifier was increased to support propane combustion and increase heat transfer to the bed of inerts. Once the bed had reached a uniform temperature of 773 - 873 K (about one hour) the coal feeder was turned on at a small feed rate while the propane feed rate was decreased until shut-off. The coal feed rate was then slowly increased until the reactor reached a temperature of approximately 1073 K. At this point the experimental conditions were set. First, steam was turned on to the desired level. Then, since it was found that coal feed rate and air flow were not independent variables, one of these two operating variables was fixed at a desired level and the other adjusted until a steady operation was reached. It was considered that the reactor had reached steady operation when both the bed temperature (measured at one point approximately in the middle of the bed) and the carbon monoxide content of the produced gas gave steady continuous readings. At this point, gas samples were taken approximately every 15-30 min and analyzed in the gas chromatograph. When sufficient data had been collected, the operating conditions were changed to obtain a new steady operation or the reactor was shut down. Each gasification run consisted of 1-4 hours of steady operation.

Bed solids samples of approximately 100 g were taken at intervals of 0.5 - 1 hour using the solids sampling port, while the ash collected in the cyclone was sampled only at the end of the run.

The shut-down procedure was as follows. Coal feeder and air were shut off while the quenching nitrogen was turned on and the steam feed rate increased. When the bed reached a temperature of 573-673 K, the bed solids were discharged through the solids sampling port. The air was then turned back on to complete the cooling of the reactor.

Coal Preparation and Analysis:

All coal samples were crushed and screened to the desired size range by B.C. Research. The coal preparation procedure was as follows. The coal samples (nominal - 12.7 mm) were first air dried for approximately a week and then screened in a vibratory screen through a set of screens of 4.76 mm, 3.36 mm and 1.0 mm. The oversize (+ 4.76 mm) was crushed by an adjustable jaw crusher in such a way as to obtain a maximum amount of sample in the 1.0 mm - 3.36 mm size range. Representative portions of the different coal samples and particle sizes were then sent to General Testing Co., Vancouver, for analysis.

For one of the coal samples (Forestburg coal) a -4.76 mm + 0.00 mm composite sample was obtained by combining the individual size fractions as shown below:

Size range (mm)	Nominal Average Size (mm)	% by weight
4.76 - 3.36	4.06	15.0
3.36 - 1.00	2.18	50.0
1.00 - 0.0	0.50	35.0
-4.76 + 0.0	$dp = \sum_i \frac{1}{(\frac{x}{dp})_i} = 1.03$	100.0

This coal sample had a nominal average size of 1.03 mm and its composition was judged as typical for a -4.76 + 0.0 mm sample of this coal crushed following the procedure outlined above.

Samples of nominal average size of 0.95 mm and 0.53 mm were also obtained from this coal by using a set of screens of 1.18 mm, 0.710 mm and 0.35 mm, and collecting the 1.18 - 0.710 mm and 0.710 - 0.355 fractions respectively.

Three types of Western Canadian coals were gasified in the small pilot plant. These had the same origin as the coal samples gasified in the spouted bed by Foong et al.(1) Ultimate and proximate analysis of the different coal samples are presented in Tables 5 and 6 respectively. Heating values are presented in Table 7 and agglomerating characteristics in Table 8.

Forestburg coal from the Luscar operation in Alberta is a sub-bituminous coal with zero free swelling index. Since this is a non-caking coal, it was used in the main part of this research in establishing the effects of the different operating parameters and most adequate conditions for the fluidized gasifier. Coleman coal from Coleman Collieries on the B.C. - Alberta border is described as a medium volatile bituminous coal of moderate caking properties with a free swelling index of 4. Sukunka coal from the Chamberlain seam of

TABLE 5

ULTIMATE ANALYSIS OF SOME WESTERN CANADIAN COAL SAMPLES

Analysis (%)	Forestburg		Coleman		Sukunka			
	-4.76 mm + 0.0 mm		3.36 mm - 1.00 mm		4.76 mm - 3.36 mm		3.36 mm - 1.00 mm	
	as received	dry	as received	dry	as received	dry	as received	dry
Carbon	50.96	66.79	77.24	78.29	78.31	79.06	79.65	80.81
Hydrogen	5.89	4.25	4.50	4.41	4.50	4.44	4.50	4.45
Sulphur	0.46	0.60	0.30	0.30	0.54	0.55	0.50	0.50
Ash	8.0	10.48	10.66	10.80	12.30	12.42	10.72	10.31
Oxygen	33.47	16.28	6.04	4.92	3.10	2.27	3.27	2.56
Nitrogen	1.22	1.60	1.26	1.28	1.25	1.26	1.36	1.37

TABLE 6

PROXIMATE ANALYSIS OF SOME WESTERN CANADIAN COAL SAMPLES

Analysis (%)	Forestburg						Coleman		Sukunka			
	-4.76 mm - 0.0 mm		4.76 mm - 3.36 mm		3.36 mm - 1.00 mm		3.36 mm - 1.00 mm		4.76 mm - 3.36 mm		3.36 mm - 1.00 mm	
	as received	dry	as received	dry	as received	dry	as received	dry	as received	dry	as received	dry
Total moisture	23.7	-	24.2	-	24.4	-	1.34	-	0.95	-	0.82	-
Ash	8.00	10.48	6.65	8.77	7.21	9.58	10.66	10.80	12.30	12.42	10.72	10.81
Volatile Matter	27.38	35.88	27.76	36.60	27.59	36.47	25.61	25.96	21.77	21.98	22.32	22.51
Fixed carbon	40.92	53.64	41.39	54.63	40.80	54.00	63.39	63.24	69.98	65.60	66.14	66.68

TABLE 7

HEATING VALUE OF SOME WESTERN CANADIAN COAL SAMPLES

Sample description	Heat value (KJ/kg)	
	"as is"	dry
Forestburg(-4.76 mm + 0.0 mm)	20694	26215
Coleman (3.36 mm - 1.00 mm)	31424	31852
Sukunka (4.76 mm - 3.36 mm)	31612	31914
Sukunka (3.36 mm - 1.00 mm)	31577	31838

TABLE 8

AGGLOMERATING CHARACTERISTICS OF SOME WESTERN CANADIAN COAL SAMPLES

Sample (a)	Free Swelling Index	Fluidity (by Gieseler Fluidity Test)					
		Initial Softening		Maximum Fluidity		Solidification	
		Temperature (K)	DDPM(b)	Temperature (K)	DDPM	Temperature (K)	DDPM
Forestburg	0	-	-	-	-	-	-
Coleman	4	725	1	739	3	765	0
Sukunka	7	723	1	761	140	795	0

(a) Coal size = 2.18 mm for all samples

(b) 1 DDPM = 1 dial division per minute = 3.6 are degree per minute in Gieseler fluidity testing aparatus (25).

the B.P. Canada property in Northeastern B.C. is a bituminous highly caking coal with a free swelling index of 7.

Coal Feed Rate:

The coal feed rate into the reactor was measured by the rate of displacement of the gas-solid interface in the coal storage bin. This could be followed through the viewing ports on the side of the bin (See Figure 3). Before loading the coal bin, the coal was weighed and its bulk density determined. By taking the time in which the coal level in the bin dropped a known distance, the volume of coal could be calculated. Since the dimensions of the bin and the coal bulk density were known, the coal mass feed rate could be calculated. This figure was checked against the feed rate calculated by dividing the amount of coal utilized during the entire run over the run duration. These two measurements were found to be in good agreement.

Fluid Flow Measurement

Except for the total wet gas flow out of the reactor, which was measured by a calibrated orifice plate, all fluid flows in and out of the reactor were measured by calibrated full view rotameters. The methods of calibration, calibration data and curves are given in Appendices II and IV respectively. Particulars of the flow metering equipment are presented in Table 9.

All gas flows are reported at the standard conditions of 294 K and 101.3 kPa.

TABLE 9

FLUID FLOW MEASUREMENT EQUIPMENT DETAIL

Stream	Flow Measurement Equipment Used
Scrubbing nitrogen	Gilmont rotameter size 14
Gasifier air	Brooks rotameter. Tube size R-10M-25-1, Float size 10-RS-64
Incinerator air	Brooks rotameter. Tube size R-10M-25-3. Float size 10-RV-64
Steam meter	Brooks rotameter. Tube size R-8M-25-2 Float size 8-RV-3
Gas sample meter	Gilmont rotameter. Size F622
Total wet gas	Orifice plate, own design. (Appendix II).

Temperature Measurement:

All temperatures were measured with calibrated Chromel-alumel thermocouples connected to a multiple channel switch and a digital display. Temperature on the reactor was measured at four or five different depths (See Table 10), at 25 mm from the reactor wall. Thermocouples were also used to measure the temperature below the fluidization grid, ambient temperature, air temperature at fluidization air rotameter, gas temperature upstream of the orifice plate, at the entrance of the gas sampling line, and at the outlet of the scrubber.

Solids Elutriation Rate:

Solids elutriated from the fluidized bed reactor were measured after each gasification experiment simply by emptying and weighing the content of the receptacle below the cyclone. By analysing the cyclone catch for ash content, both the ash and carbon elutriation rates could be calculated (see next paragraph).

Solids Analysis:

Both the solids samples withdrawn from the fluidized bed and a representative sample of the cyclone catch were analysed for ash content by incineration. The 50 - 100 g samples were weighed, put into tared crucibles and incinerated for 10 hours at 1173 K in an electric furnace. After cooling, the samples were weighed again to determine the residual ash. The weight loss was taken as carbon due to the fact that the high reactor temperatures would have driven off water and volatile matter from the solid samples.

TABLE 10

THERMOCOUPLE LOCATION IN THE FLUIDIZED BED

Distance from Fluidization Grid (m).	
0.70 m Fluid bed	0.86 m Fluid bed
0.05	0.06
0.21	0.31
0.36	0.46
0.53	0.61
-	0.76

Tar Determination:

On selected runs, the approximate tar content of the gas produced was determined by the following method. After the reactor had reached steady operation, the gas sample stream was diverted from the cleaning set of impingers in use into a second set of impingers for a fixed period of time (1 - 2 hours). Here, the tar condensed in the cold water and on the walls of the impingers. The tar was then removed by dissolving it with acetone. The acetone was then evaporated under vacuum and the residual tar weighed. The tar content of the gas was calculated by dividing the total tar thus obtained over the total gas volume passing through the impingers during the tar sampling period.

Gas Analysis:

All gas analyses were reported on a dry basis. The carbon monoxide content of the gas was continuously monitored by an infrared analyser while the remaining gases were analysed from samples taken at convenient time intervals (15 - 30 min) in a gas chromatograph. Both instruments were connected to a Watanabe chart recorder.

Continuous CO monitor:

A Beckman continuous infrared analyser, model No 864-13-4, which operated in the range of 0 - 25% CO was employed. The instrument gave a continuous signal on the recorder chart. Carbon monoxide content of the gas was obtained from a factory provided calibration chart (Figure V-1, Appendix V). The instrument was calibrated for every run by adjusting the readings of two points of the calibration curve. The zero reading was adjusted by passing through the instrument a standard gas stream containing 10% CO_2 and 90% nitrogen. The second calibration point was obtained by adjusting the instrument's reading on a standard sample containing 10% CO_2, 21.3% CO and nitrogen as balance, to the calibration curve reading (91% deflection).

The continuous CO analyser provided an excellent means of monitoring the gasifier performance as well as providing an indication of whether the reactor was operating under steady conditions.

Gas chromatograph:

At convenient time intervals, 5 cm^3 gas samples were injected into a Hewlett-Packard gas chromatograph model 5710 A. The gas chromatograph was equipped with a molecular sieve column, a poropack column and a thermal conductivity detector which resolved hydrogen, carbon dioxide, nitrogen, oxygen, methane and carbon monoxide. Typical gas chromatograph tracings are shown as Figure V-2, Appendix V. The percentage content of each gas in the sample was determined from the peak height and the corresponding calibration curves. (Figures V-3 - V-5, Appendix V). In order to avoid possible variations of the calibration curves, these were obtained for each run by analysing dilutions in air of a standard gas sample (Table V-1; Appendix V). Experience showed that carbon monoxide was best obtained from the infrared analyser and carbon dioxide by difference. Therefore, only H_2, N_2 and CH_4 were obtained from the gas chromatograph (since there was no O_2 under gasification operation conditions).

Gas Calorific Value:

The calorific value of the gas was calculated from the gross (high) heat of combustion of the three combustible components of the gas produced (hydrogen, carbon monoxide and methane), and the dry gas composition. In order to enable comparison with values reported in the literature, the gross calorific value of a unit volume of gas (h_{cg}) was calculated at the North-American standard conditions (288.6 K = 60°F and 101.6 kPa = 30 in Hg dry).

Then:

$$h_{cg}\left[= h_{H_2} \times (\%\ H_2\ v/v) + h_{CO}\ (\%\ CO\ v/v) + h_{CH_4}\ (\%\ CH_4\ v/v)\right] /100$$

where:(*)

$h_{H_2} = 12.109\ (MJ/m^3)$

$h_{CO} = 11.997\ (MJ/m^3)$

$h_{CH_4} = 37.743\ (MJ/m^3)$

(*) "Fuel Flue Gases". American Gas Association

CHAPTER IV - RESULTS

MINIMUM FLUIDIZATION VELOCITY AND SEGREGATION

The minimum fluidization velocity of 4.45 kg mixtures of coal and silica each of a size range 1.18 - 3.36 mm were determined at room temperature in a glass column of the same diameter as the coal gasifier (Figure 2). At the same time the mixing patterns were visually observed to determine conditions where segregation or slugging affected fluidization. Pressure drop across the bed, bed height and visual observations made at different air flows are presented in Tables VI-1 to VI-6, Appendix VI, for the various mixtures studied. This data is summarized in Table 11.

For all mixtures of coal and silica tested, the curves showing pressure drop across the bed as a function of the superficial air velocity (Figures 4 & 5) had a similar shape and were typical of fluid flow through a bed of solids (30). For a fixed bed, the pressure drop increased almost linearly with the fluid velocity. Upon reaching the minimum fluidization velocity the pressure drop ceased to increase and remained almost constant or decreased slightly as the fluid velocity was further increased.

The minimum fluidizaiton velocity of a mixture of 80% 3.36 - 1.18 mm silica and 20% coal of the same size was approximately 0.78 m/s (Figure 4). Once the bed became fluidized, coal segregated from the silica and floated at the top of the bed. As the superficial air velocity was increased, segregation started to disappear, and at a superficial velocity of approximately 1.10 m/s (1.41 U_{mf}) the bed was completely mixed. However, the disappearance of segregation corresponded to the commencement of slugging (Figure 4).

TABLE 11

FLUIDIZATION OF MIXTURES OF COAL AND SILICA (*)

80% 3.36-1.18mm Silica 20% Coal		80% Ottawa Sand 20% Coal		85% Ottawa Sand 15% Coal		90% Ottawa Sand 10% Coal		95% Ottawa Sand 5% Coal		100% Ottawa Sand	
Superficial air velocity (m/s)	Pressure drop (kPa)	Superficial air velocity (m/s)	Pressure drop (kPa)	Superficial air velocity (m/s)	Pressure drop (kPa)	Superficial air velocity (m/s)	Pressure drop (kPa)	Superficial air velocity (m/s)	Pressure drop (kPa)	Superficial air velocity (m/s)	Pressure drop (kPa)
0	0	0	0	0	0	0	0	0	0	0	0
0.26	0.32	0.21	1.35	0.16	1.02	0.16	1.00	0.16	1.02	0.16	1.17
0.39	0.75	0.26	1.64	0.21	1.30	0.21	1.32	0.21	1.27	0.21	1.50
0.52	1.22	0.31	2.19	0.26	1.77	0.23	1.54	0.23	1.50	0.23	1.74
0.65	1.87	0.34	2.19	0.28	2.02	0.26	1.67	0.26	1.69	0.26	2.02
0.70	1.89	0.36	2.17	0.31	2.17	0.28	1.97	0.28	1.97	0.28	2.24
0.78	2.19	0.41	2.14	0.34	2.12	0.31	2.12	0.31	2.17	0.31	2.27
0.83	1.99	0.47	2.12	0.36	2.12	0.34	2.09	0.34	2.17	0.34	2.24
0.88	1.99	0.52(a)	2.12	0.41	2.12	0.36	2.09	0.36	2.14	0.36	2.27
0.93	1.94	0.56	2.07	0.47	2.07	0.39	2.07	0.39	2.14	0.41	2.22
0.98	1.89	0.62(b)	2.07	0.52(a)	2.09	0.41	2.07	0.41	2.12	0.47	2.22
1.04(b)	1.87	0.78	1.99	0.57	2.04	0.44	2.07	0.44	2.12	0.52	2.19
1.09(a)	1.87			0.62	2.02	0.47	2.04	0.47	2.07	0.65(b)	2.14
1.14	1.84			0.67	1.97	0.52(a)	2.04	0.52(a)	2.07	0.73	2.12
				0.72(b)	1.94	0.57	2.02	0.57	2.02	0.91	2.07
				0.91	1.87	0.62(b)	1.97	0.65(b)	2.02		
						0.70	1.94	0.70	1.99		
						0.78	1.92	0.78	1.94		
						0.91	1.84	0.91	1.87		

(*) Coal and silica in size range 1.18 - 3.36 mm

(a) Minimum superficial velocity at which segregation disappears

(b) Onset of slugging.

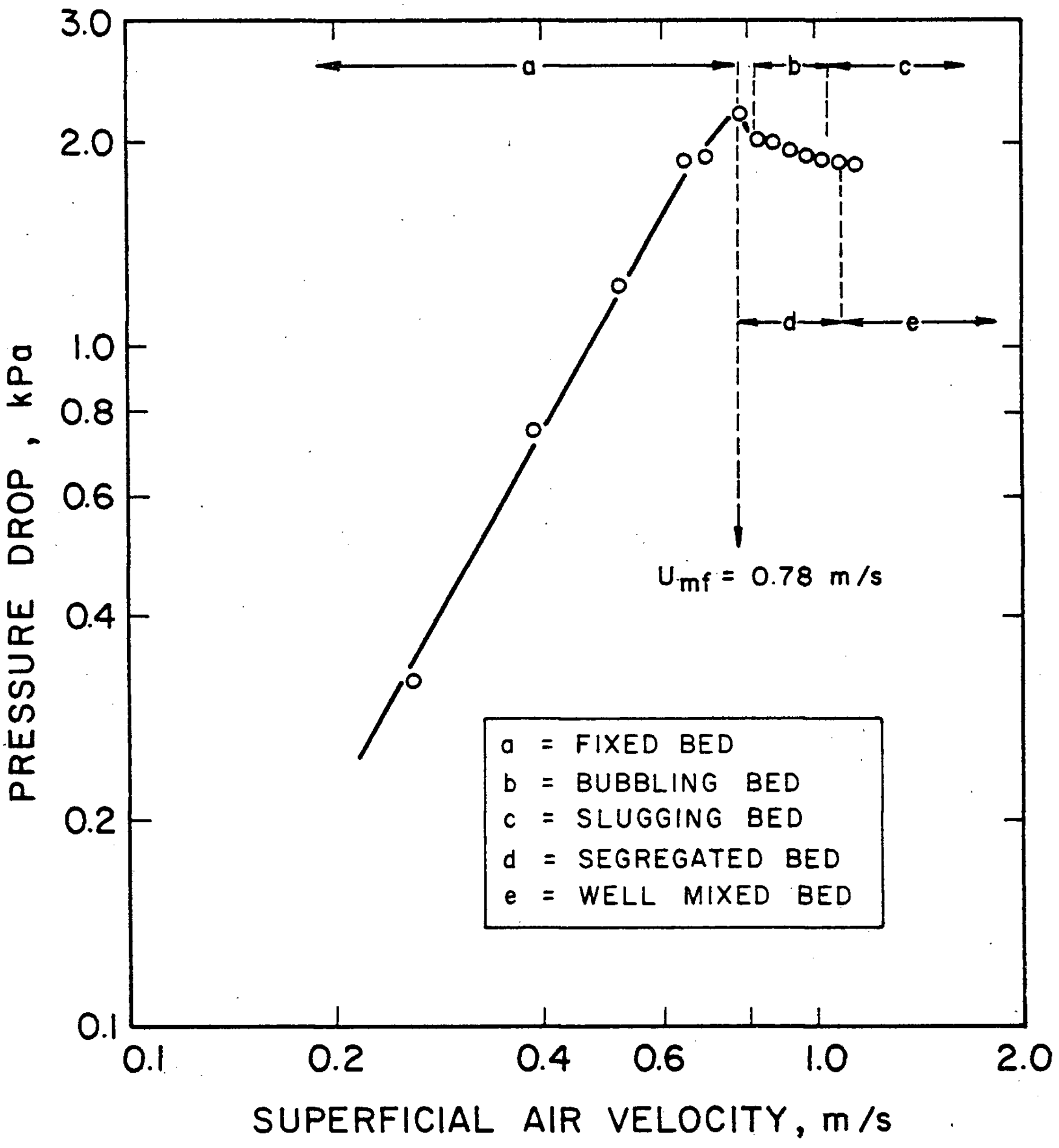


FIGURE 4: FLUIDIZATION OF A MIXTURE OF 80% 3.36-1.18 mm SILICA AND 20% 3.36-1.18 mm COAL

Substitution of the coarse crushed silica with uniformly sized 0.73 mm Ottawa sand resulted in an improvement of the mixing of coal and inerts in the fluidized bed. The minumum fluidization velocity of four different mixtures of Ottawa sand and varying amounts of coal (5% to 20%) was very similar, at approximately 0.32 m/s (Figure 5). For these mixtures very little segregation was observed, even for fluid velocities slightly above the minimum fluidization velocity, (U_{mf}). Visual observation indicated that in general, for a fluid velocity larger than 0.5 m/s (1.56 U_{mf}) excellent dispersion of coal in silica was obtained. At this velocity, the bed was vigorously bubbling, while at fluid velocities larger than 0.7 m/s for all cases the bed was slugging.

Overall, these experiments suggested that solids segregation was less intense and fluidization much smoother for mixtures of 3.36 - 1.18 mm coal and Ottawa sand than for mixtures of coal and silica of the same size. The minimum superficial velocity required for achieving good dispersion of coal in silica was in the range:

$$1.41\ U_{mf} \leq U \leq 1.56\ U_{mf}$$

which is in line with values cited in the literature $^{(33)}$. As expected for beds of the same weight, pressure drop across the bed was similar in all cases. As well the minimum fluidization velocity for mixtures of 3.36 - 1.18 mm coal in Ottawa sand were much smaller than for mixtures of coal and silica of the same (3.36 - 1.18 mm) size. This could be an important consideration in an industrial application.

As a result of these tests, it was decided to operate the fluid bed gasifier using 0.73 mm Ottawa sand as an inert dispersing medium for coal.

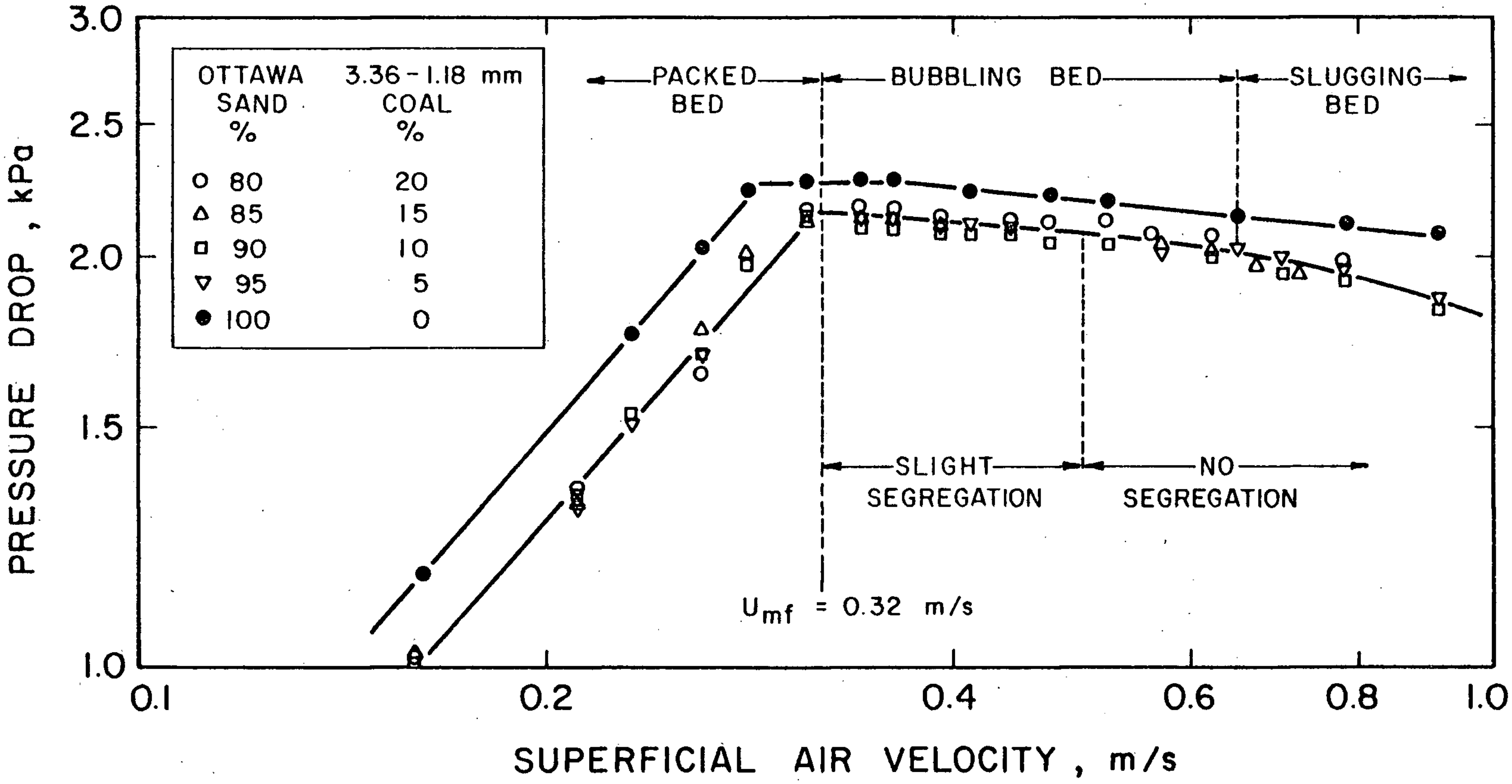


FIGURE 5: FLUIDIZATION OF MIXTURES OF 0.73 mm OTTAWA SAND AND 3.36-1.18 mm COAL

COAL GASIFICATION

The three coals described in Chapter III were gasified in a fluidized bed of 0.73 mm Ottawa sand. As outlined before, most of the gasification experiments utilized the non-caking Forestburg coal, and the data derived was used in establishing the effect of the different operating parameters on the quality of the operation and of the gas obtained. Gasification of caking coals was attempted only under selected conditions so as to establish the feasibility of gasifying caking coals in the fluidized bed and as a means of confirming some of the experimental findings on the effect of the operating variables derived in the first part of this research.

The experimental results are summarized in Tables VII-1 and VII-2, Appendix VII. The tabulated values are averages of 2 to 10 different measurements, and in general, observed variations from the reported averages are within 5%, except for gas analysis where variations of up to 10% may have occurred. Except when otherwise noted, all reported values correspond to steady operation of the gasifier. The gasifier was considered to be operating under steady conditions when there was neither a continuous decrease or increase of temperature at a rate exceeding 0.8 K/min at a typical operating temperature of 1080 K nor a continuous change in the carbon monoxide content of the gas.

The effects of the different operating parameters on the gasifier operation and on the gas quality are discussed in the following sections.

Operating Experience with the Fluidized Bed

The stability of the fluidized bed operation was found to be a delicate balance between mass and heat transfer, and both the gas composition and bed temperature were very sensitive to changes in the operating parameters. These parameters were interrelated and it was found impossible to change one operating variable and keep the others constant without introducing instability in the operation. The two main operating variables were the coal feed rate and the air flow rate. Once one of these parameters was fixed, the other as well as the bed temperature, was fixed within a relatively narrow range. A higher degree of freedom to change the steam injection rate was experienced, and this variable could be changed within a wider range without disequilibrating the system. However, an increase in steam injection rate invariably caused a decrease of the bed temperature, as well as a change in the gas composition.

The air flow into the reactor was limited at its lower end by the flow corresponding to a superficial air velocity (at the average reactor temperature) equal or larger than the previously determined minimum non-segregating velocity (0.5 m/s) (see Section 1, this Chapter). The upper limit for the air flow was approximately 0.007 - 0.008 std m^3/s or a corresponding superficial velocity of about 1.6 m/s. For values larger than this the bed began to slug excessively and overflowed into the freeboard. In practice, it was found necessary to operate at air flows resulting in superficial velocities in the range of 1.0 - 1.4 m/s; 1.2 m/s being a typical value. Under these conditions the bed was violently agitated or slugging.

On the other hand, coal feed rate was limited at its lower end by the minimum feed rate which at the given air flow would result in an operating temperature below 1175 K, the maximum safe temperature for the unlined stainless steel reactor. This corresponded to a feed rate of about 0.27 g/s of Forestburg coal (see run 10, Table VII-1, Appendix VII) at the minimum air flow for the absence of segregation in the bed. The maximum coal feed rate was determined by the coal reactivity. If coal is fed to the bed at a higher rate than it can react, it will accumulate, causing an increase of the bed's solids content, and eventually, defluidization.

The carbon content of the bed was monitored as a function of time for various gasification runs using Forestburg and Sukunka coals (Tables VII-3 and VII-4, Appendix VII) and the results plotted in Figures 6 and 7 respectively. During the gasification of Forestburg coal, carbon did not accumulate in the bed when the air to coal ratio was $\geq$ 4.93 (Figure 6). However, for an air to coal ratio of 4.01 the bed carbon content increased rapidly, to over 30%, in less than 1 hour. Therefore, the minimum air to coal ratio that allowed operation without carbon accumulation in the bed was in the range of 4.01 - 4.93, and probably slightly above 4.0 judging from operating experience. If the previously stated maximum air flow is used, the maximum coal feed rate without carbon accumulation is expected to be 1.95 - 2.39 g/s. The experimental results confirmed this since all runs with air to coal ratios below 4.0 were unstable (see Table VII-1, Appendix VII).

Sukunka coal appeared to be somewhat less reactive than Forestburg coal. Figure 7 shows that at an air to coal ratio of 4.63, initially carbon accumulates in the bed quite rapidly, with the rate of accumulation decreasing after about an hour of operation. For a slightly higher air to coal ratio (4.69) the bed carbon content oscillated between 21% and 10%, while operation at an air to coal ratio of 4.11 resulted in a somewhat unstable operation with temperature decreasing at a rate of approximately 0.8 K/min (Run No. 38).

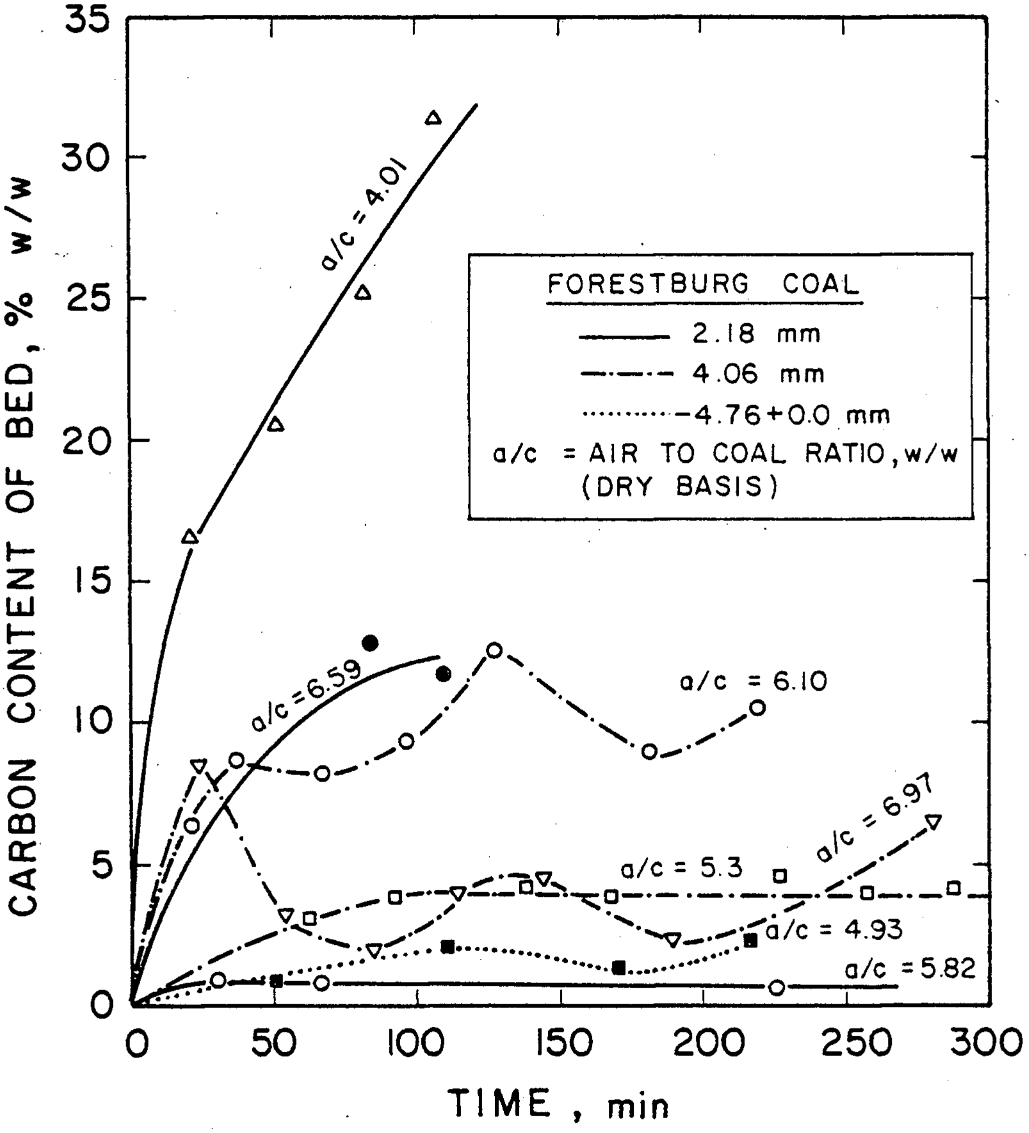


FIGURE 6: CARBON CONTENT OF FLUIDIZED BED DURING THE GASIFICATION OF FORESTBURG COAL

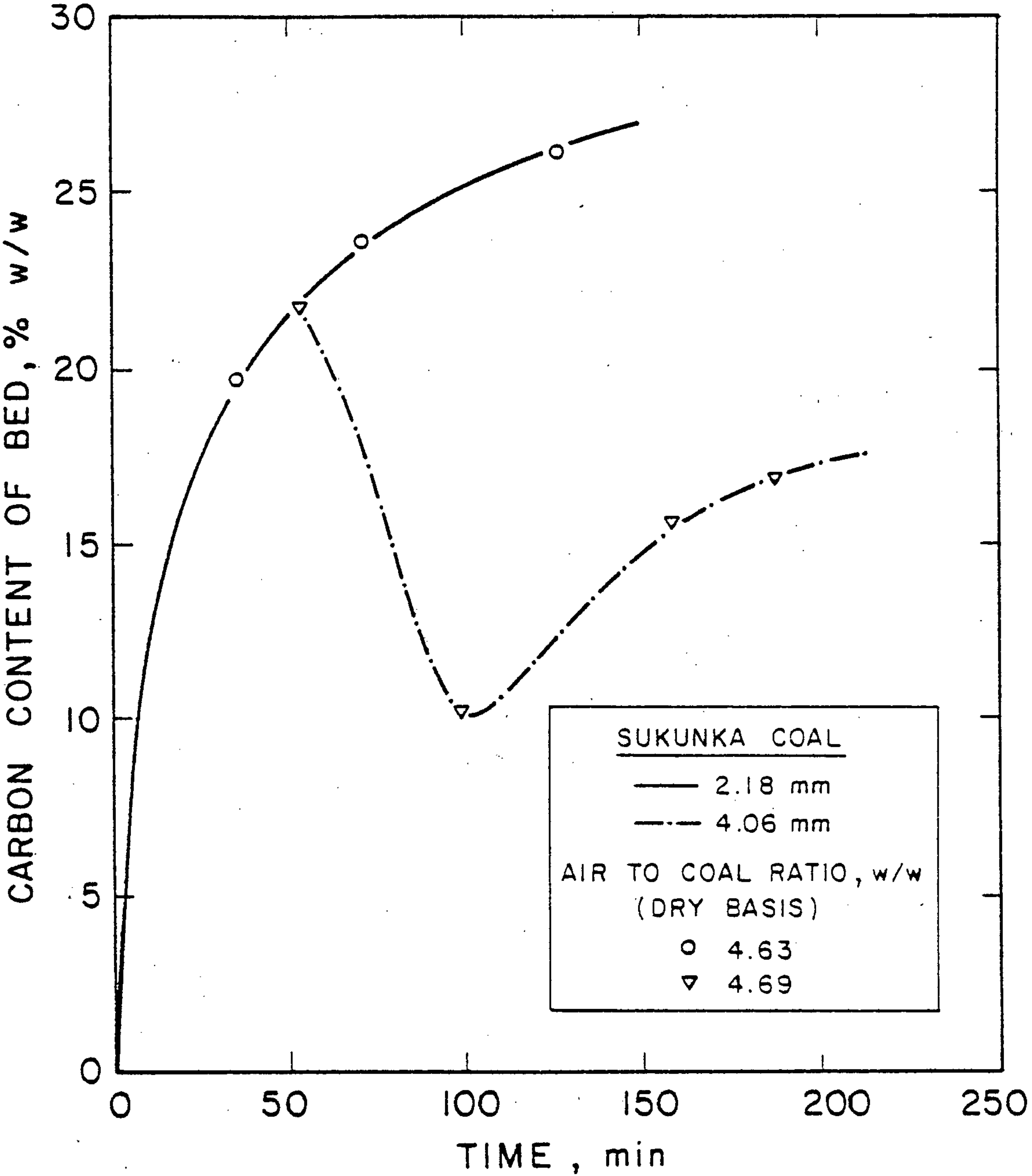


FIGURE 7: CARBON CONTENT OF FLUIDIZED BED DURING THE GASIFICATION OF SUKUNKA COAL

Effect of Temperature

Axial temperature profiles in the reactor were typical of a fluidized bed (Figure 8) i.e. fairly flat with maximum temperature differences of approximately 50 K. The maximum temperature in the bed occurred at a point approximately 13 mm below the feed point.

As mentioned earlier, one of the basic limitations of the present experimental set-up was the inability of operating at average temperatures over about 1175 K because the reactor was not refractory lined. This severely affected the quality of the gas produced since most of the data available in the literature suggests that the carbon-steam reaction does not proceed at considerable rates at temperatures below about 1273 K (see Chapter I). All of the experimental data was obtained in the 1023 - 1173 K temperature range. Within this range, the results suggest that temperature had no effect on the quality of the gas obtained from either Forestburg coal (Figure 9) or Sukunka coal (Figure 10).

Effect of Coal Quality

Forestburg coal

A total of 33 gasification runs were conducted with Forestburg coal of five different particle sizes, all below 4.76 mm (Table VII-1, Appendix VII). As expected with a non-caking coal the fluidized bed gasifier operated smoothly, and no particular problems were encountered under the wide range of operating conditions tested.

For coal feed rates of 0.27 - 2.37 g/s, air flows of 2.75 x 10^{-3} - 7.92 x 10^{-3} m^3/s and steam feed rates of 0.0 - 1.165 g/s the best gas obtained had a calorific value of 2.94 MJ/m^3 (78.9 Btu/cf)

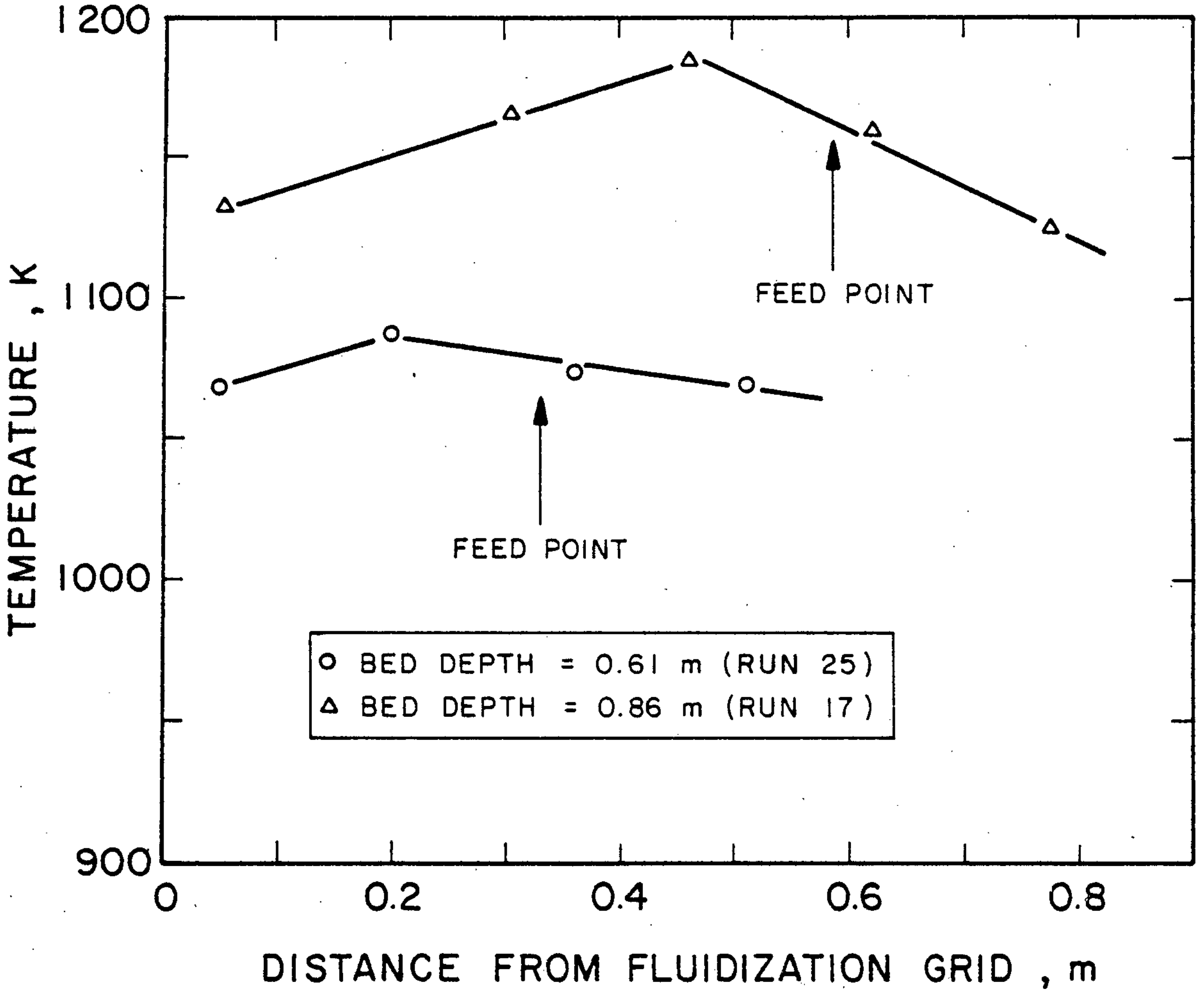


FIGURE 8: TYPICAL AXIAL TEMPERATURE PROFILES OF FLUIDIZED BED

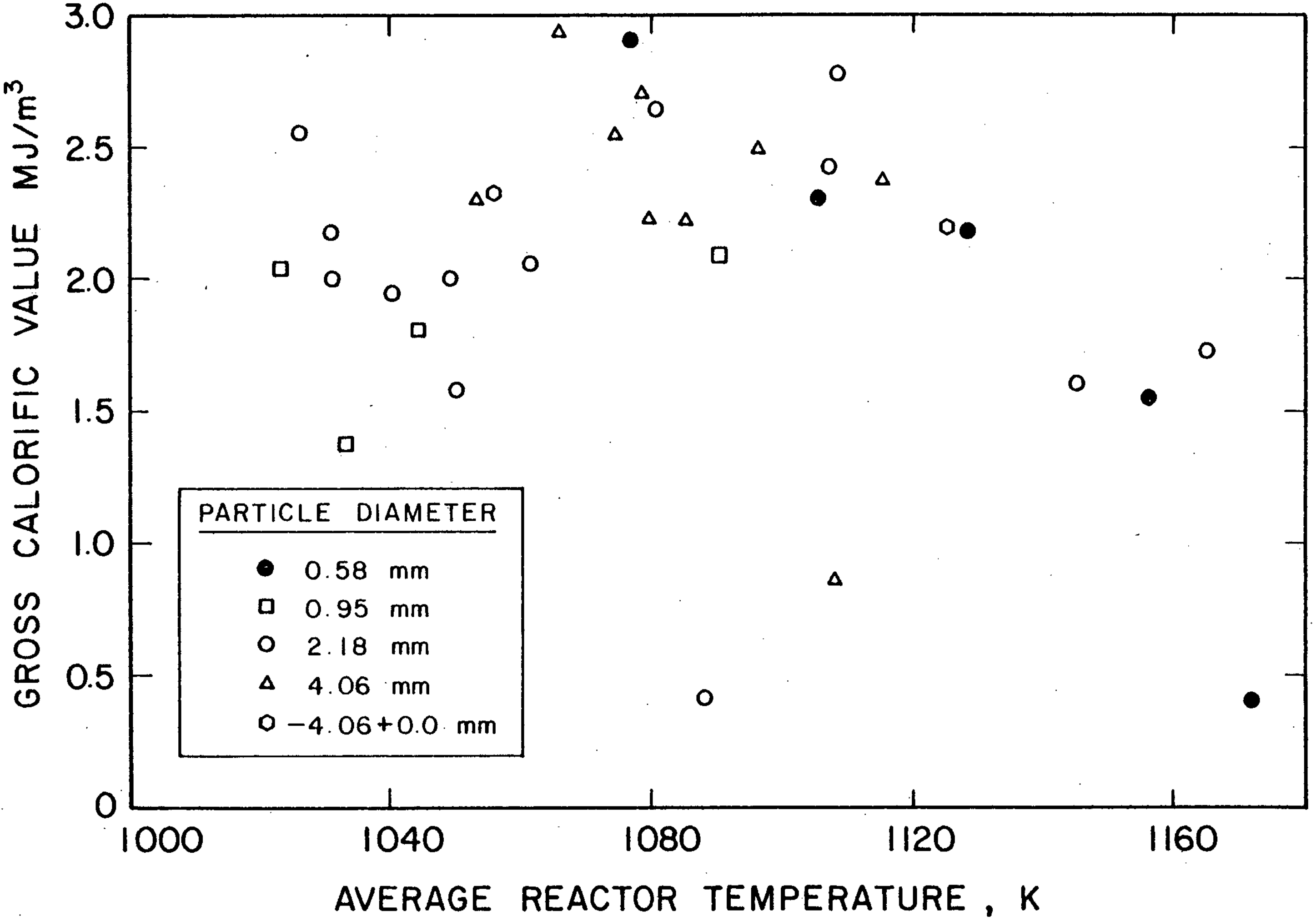


FIGURE 9: EFFECT OF TEMPERATURE ON THE GASIFICATION OF FORESTBURG COAL

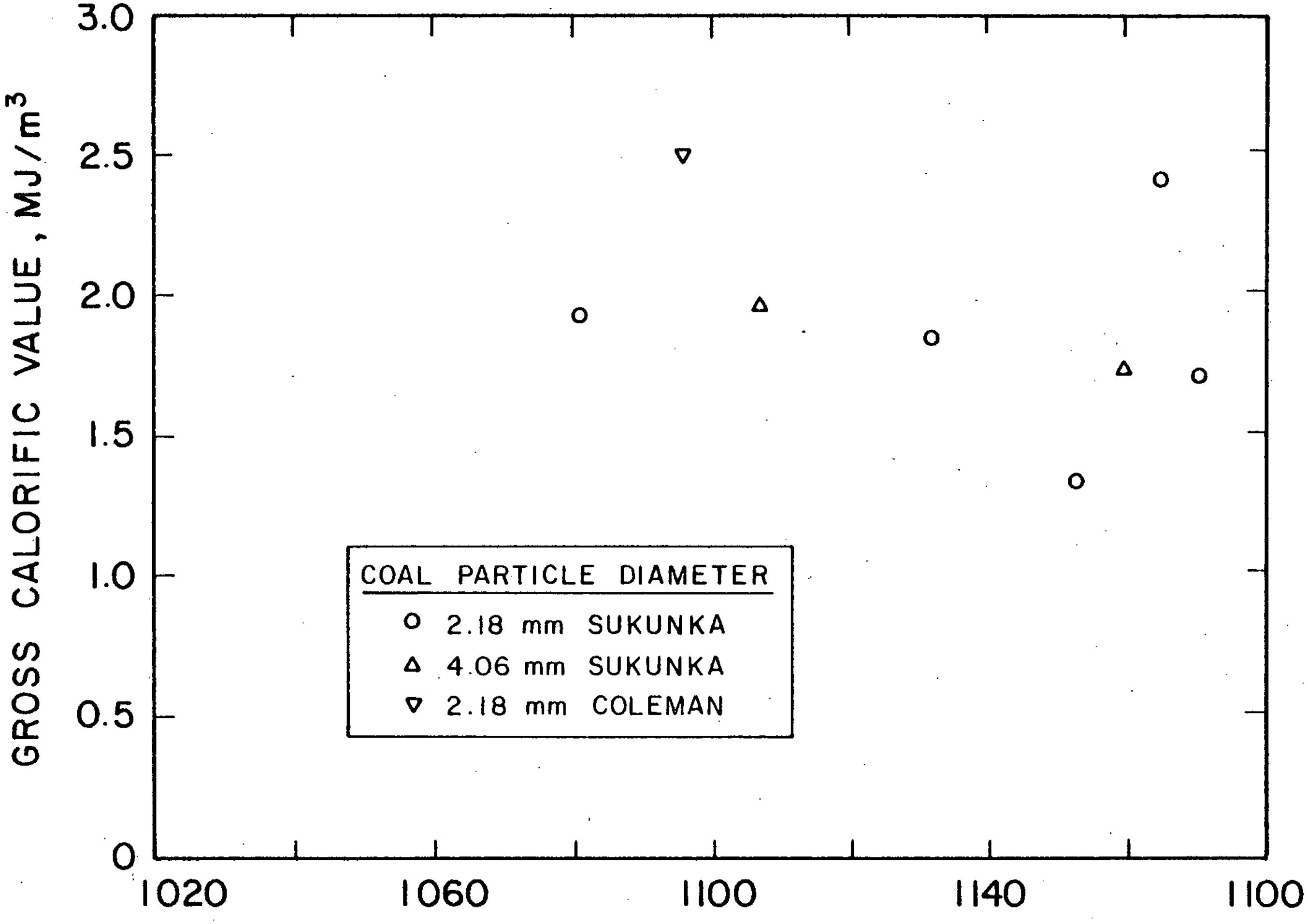


FIGURE 10: EFFECT OF TEMPERATURE ON THE GASIFICATION OF SUKUNKA COAL

and the poorest 0.41 MJ/m^3(11 Btu/cf) (Table 12). Typically the gasification of Forestburg coal produced a gas with a calorific value of 2.0 - 2.5 MJ/m^3 (53.7 - 67.1 Btu/cf)

Typical dry gas composition was as follows:

H_2 : 5.9 - 10.2%
CO : 5.9 - 12%
CH_4: 0.5 - 0.9%
CO_2: 8.2 - 14.1%
N_2 : 68.1 - 75.5%

The abnormally high nitrogen content of the gas was partly due to the use of nitrogen in the feeding system (Figure 3). Assuming that this nitrogen has only a dilution effect, the gas composition and the gross calorific value of the gas have been corrected for the introduction of nitrogen (Table VII-5, Appendix VII), which would not be necessary in an improved design of the feed system. This correction increases the concentrations of all gases but nitrogen, and the gross calorific value by percentages between 9.6 and 2.6%, depending on the air input, and are higher for lower air input. Under most operating conditions, the correction factor was between 3 and 4%. All compositions reported are uncorrected, except for those in Tables VII-5-6, Appendix VII.

Sukunka coal

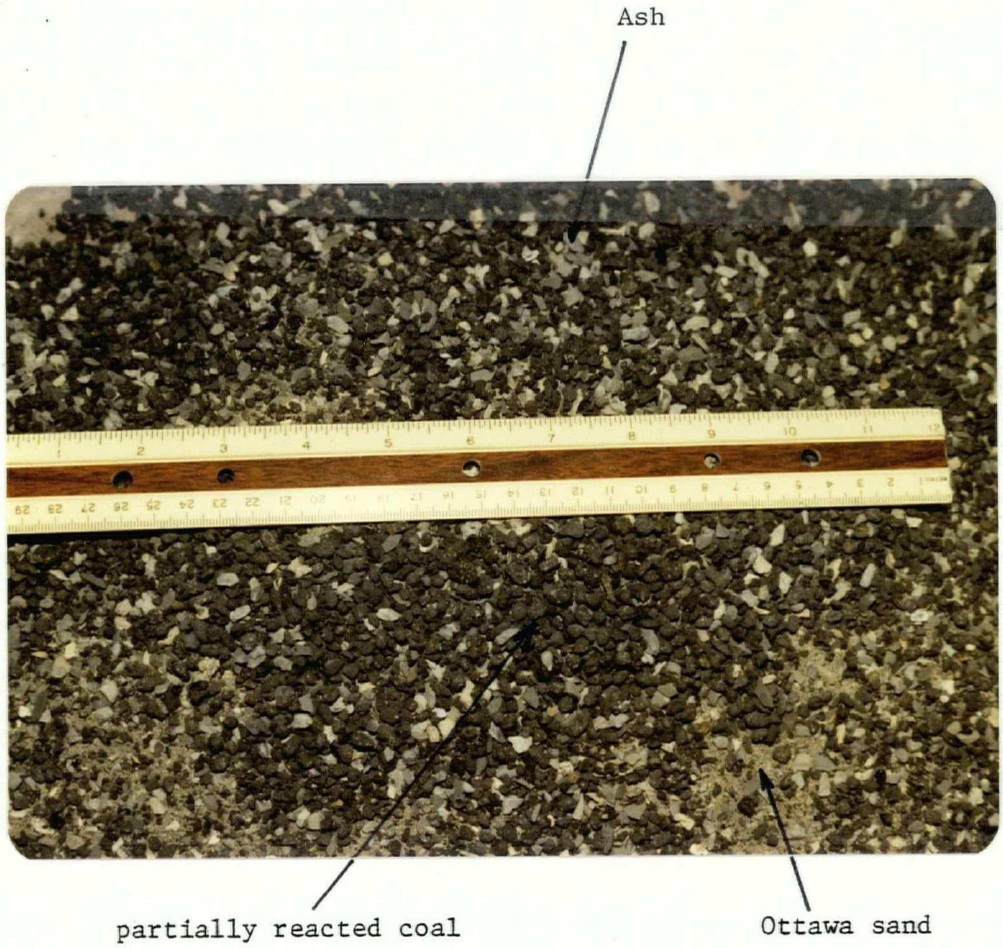
A total of seven gasification runs using either 2.18 mm or 4.06 mm Sukunka coal were carried out. (Tables VII-2, Appendix VII). This coal which has a swelling index of 7, could be gasified at feed rates of up to 1.52 g/s without agglomeration in the fluid bed (See Photograph 1) or sintering at the fluidization grid. Higher

TABLE 12

TYPICAL RESULTS FOR THE GASIFICATION OF-4.76 mm FORESTBURG COAL

Type of result	Run No. (a)	Average coal particle size (mm)	Coal feed rate (dry basis) (g/s)	Air flow (b) (m^3/s)	Steam feed rate (g/s)	Gas Composition, % (v/v) (dry Basis)					Gross gas calorific value (MJ/m^3)(c)
						H_2	CO	CH_4	CO_2	N_2	
Best	26	4.06	1.259	6.11×10^{-3}	0.763	11.6	10.9	0.6	9.1	67.8	2.94
Typical	2	0.53	1.060	4.68×10^{-3}	0.377	7.8	8.0	0.5	8.2	75.5	2.09
	3	0.53	1.060	3.03	0.395	8.7	5.9	0.7	11.9	72.8	2.03
	5	0.95	0.983	4.19	0.542	9.0	7.9	0.7	11.4	71.0	2.30
	7	0.95	0.983	5.44	0.433	7.7	8.9	0.5	12.2	70.7	2.19
	13	2.18	0.798	5.48	0.283	7.3	8.2	0.5	10.8	73.2	2.06
	14	2.18	1.185	6.51	0.542	8.6	7.2	0.7	10.9	72.6	2.17
	15	2.18	1.185	6.52	0.250	7.2	7.2	0.7	10.7	74.2	2.00
	16	2.18	1.273	4.74	0.00	5.9	12.0	0.7	9.8	71.6	2.42
	19	2.18	1.673	7.00	0.167	6.4	8.0	0.7	12.3	72.5	2.00
	22	2.18	2.369	7.92	0.526	10.2	8.2	0.9	12.6	68.1	2.56
	25	4.06	1.223	6.36	0.885	10.1	8.9	0.6	13.4	67.0	2.52
	27	4.06	1.516	6.67	0.00	7.7	10.2	0.6	12.2	69.3	2.38
	28	4.06	1.516	6.74	0.542	8.5	8.3	0.6	13.8	68.8	2.25
	29	4.06	1.516	6.75	0.748	8.7	7.6	0.7	13.9	69.1	2.23
	30	4.06	1.516	6.76	1.165	10.0	6.9	0.7	14.1	68.3	2.30
	32	-4.76+0.0	1.059	4.27	0.00	6.3	10.1	0.6	10.6	72.4	2.20
	33	-4.76+0.0	1.059	4.42	0.433	9.9	7.4	0.7	12.2	69.8	2.35
Worst	10	2.18	0.271	2.75×10^{-3}	0.283	1.0	1.8	0.2	11.5	85.5	0.41

(a) See Table VII-1, Appendix VII for details
(b) At 294 K and 101.3 kPa
(c) At 288.6 K and 101.6 kPa (North-American Gas Standard).


Ash
partially reacted coal
Ottawa sand

PHOTOGRAPH 1: Bed solids during the gasification of 4.06 mm Sukunka coal (Run No. 40) feed rates were not tried. Agglomeration of the coal in the feed pipe was experienced initially, but this problem was solved by the introduction of a small variable speed agitator into the feed pipe (Figure 3).

For coal feed rates in the range of 1.46 - 1.58 g/s, air flow of 5.43 x 10^{-3} std m^3/s and steam feed rates of 0.00 - 0.455 g/s, the Sukunka coal produced a gas with a gross calorific value in the range of 1.85 - 2.43 MJ/m^3 (49.7 - 65.2 Btu/cf) (Table 13). These values are slightly lower than those obtained with Forestburg coal in the same range of operating conditions.

Under the operating conditions described above, typical dry composition of the gases obtained for the gasification of Sunkunka coal was:

H2 : 4.3 - 6.8%
CO : 6.6 - 7.1%
CH_4: 1.0 - 2.0%
CO_2: 12.1 - 15.7%
N_2 : 70.0 - 74.8%

As can be seen from the gas composition, methane contributed an important part of the gas calorific value; and since methane is mostly produced by pyrolysis, coal feed rate played an important role in determining the gas quality (see Table VII-2, Appendix VII).

Gas composition and calorific value of the gas produced from gasification of Sukunka coal corrected for the introduction of nitrogen in the feed system are presented in Table VII-6, Appendix VII. Corrections factor were very similar, and averaged 6.10%,

TABLE 13

TYPICAL RESULTS FOR THE GASIFICATION OF - 4.76 mm CAKING COALS

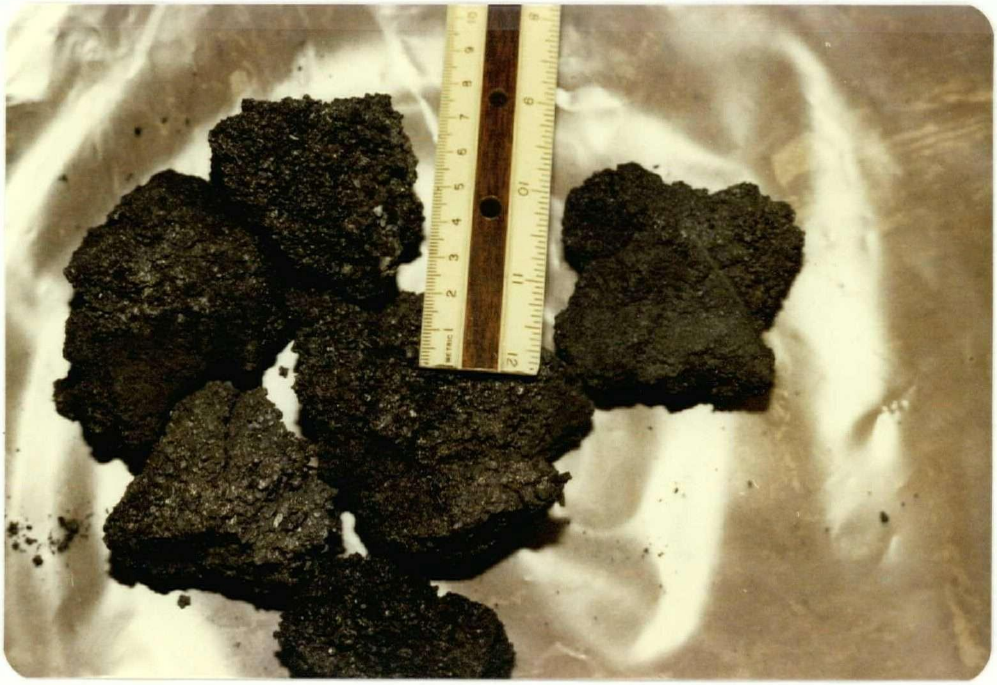
Type of coal	Run No. (a)	Average coal particle size (mm)	Coal feed rate (g/s)	Air flow std (m^3/s)	Steam feed rate (g/s)	Gas composition % v/v (dry basis)					Gross gas calorific value (MJ/m^3)
						H_2	CO	CH_4	CO_2	N_2	
Sukunka	36	2.18	1.462	5.58×10^{-3}	0.00	4.6	7.0	1.2	12.4	74.8	1.85
Sukunka	37	2.18	1.462	5.69	0.455	5.6	6.6	1.2	12.3	74.4	1.92
Sukunka	38	2.18	1.580	5.43	0.00	6.8	7.1	2.0	12.1	72.0	2.43
Sukunka	39	4.06	1.462	5.68	0.00	4.3	7.1	1.0	15.3	72.4	1.75
Sukunka	40	4.06	1.462	5.73×10^{-3}	0.358	6.2	7.0	1.0	15.7	70.0	1.97
Coleman	41	2.18	1.968	5.72×10^{-3}	0.00	7.1	7.8	1.9	10.9	72.5	2.51

(a) See Table VII-2, Appendix VII for details higher than for the Forestburg coal since introduction of more nitrogen was necessary when operating with caking coal. As a consequence, the corrected calorific values of gas produced from Forestburg and Sukunka coal were similar.

Coleman coal

Only one gasification run was carried out with Coleman coal, a caking coal of free swelling index 4. In this experiment, 2.18 mm coal was fed to the reactor at a rate of 1.97 g/s. Air flow was 5.72×10^{-3} m^3/s and steam feed rate was zero. (See Table VII-2, Appendix VII). Under these conditions, steady gasification proceeded without apparent problems producing a gas of 2.51 MJ/m^3 (67.37 Btu/cf) and the composition shown in Table 13. However, after approximately 2 hours of operation, the pressure inside the reactor started to build up, and the operation was shut down. When air was used to cool down the reactor, a fire started in the upper disengaging section which resulted in the destruction of that section. Dismantling of the reactor showed that the coal had caked (See Photograph 2) and bridged across the bed near the feed point, causing almost complete blockage of the upper part of the reactor and the consequent pressure build-up.

Analysis of the data obtained under these conditions reveals that the gas calorific value was in the range obtained for similar operating conditions for the two previously tested coals (see runs 21 and 38, Appendix VII), and that an important portion (28.5%) of the gas calorific value was produced by methane. Since in the present experimental set-up the only possibility of achieving trouble-free



PHOTOGRAPH 2: Caked Coleman coal at end of Run No. 41

operation with Coleman coal would require decreasing the coal feed rate with the consequent decrease of methane production and gas quality, it was decided not to pursue any further experiments with this type of caking coal.

Effect of Coal Particle Size

The gross calorific value of the gas obtained from narrowly sized Forestburg coal under a wide range of operating conditions is plotted as a function of the air to coal ratio (dry basis) in Figure 11. Comparison of the curves obtained for each coal size suggests that in general, an improvement on the gas quality may be expected as the coal particle size increases. However, the data is somewhat scattered and for certain operating conditions it was possible to obtain gas of similar calorific values irrespective of the coal size. Table 14 illustrates that gasification of Forestburg coal of four different sizes at air to coal ratios of 5.0 - 5.3 produced gas of a calorific value in the range of 1.94 - 2.30 MJ/m^3 and that there was no correlation with the coal size. When coal of a wider size distribution (-4.76 mm + 0.00 mm) was gasified under the above conditions, a similar gas was obtained (Run 33, Table 14). Note that although not all the experiments compared in Table 14 were performed at the same steam to coal ratio, this operating parameter did not have a substantial effect on the gas heating value as discussed later.

Little effect of coal size on gas quality was observed for Sukunka coal (Table 15). At air to coal ratios of 4.6 - 4.7 gasification of 2.18 mm and 4.06 mm Sukunka coal gave similar results.

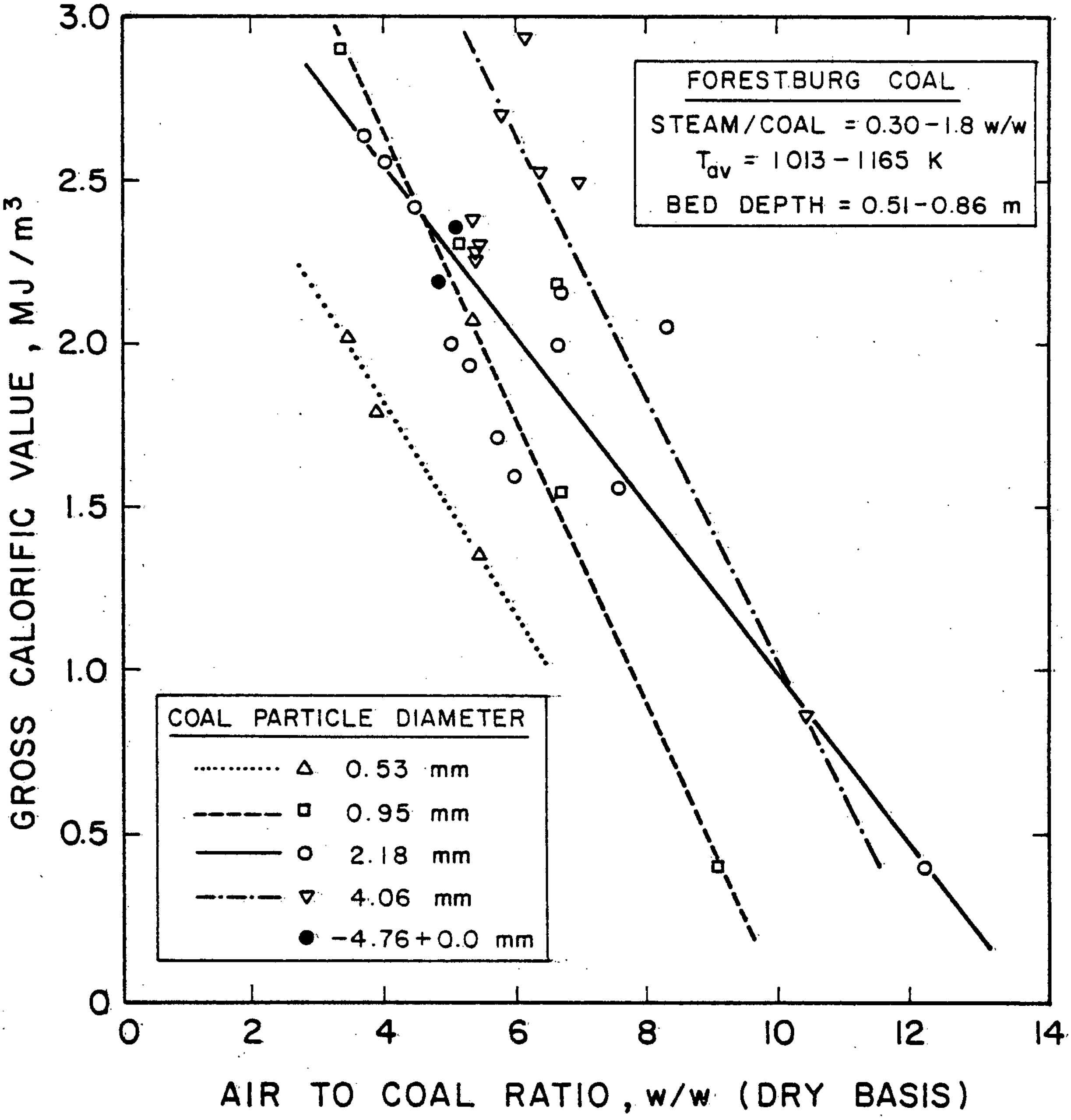


FIGURE 11: EFFECT OF AIR TO COAL RATIO AND PARTICLE SIZE ON THE CALORIFIC VALUE OF GAS PRODUCED FROM FORESTBURG COAL

TABLE 14

EFFECT OF COAL PARTICLE SIZE FOR THE GASIFICATION OF FORESTBURG COAL

Run # (a)	Nominal average size (mm)	Air to coal ratio (dry basis) (w/w)	Steam to dry coal ratio (w/w)	Gas Composition (% v/v) (dry basis)					Gas Gross heating value (MJ/m^3)
				H_2	CO	CH_4	CO_2	N_2	
$2^{(b)}$	0.53	5.29	0.436	7.8	8.0	0.5	8.2	75.5	2.09
$5^{(b)}$	0.95	5.10	0.632	9.0	7.9	0.7	11.4	71.0	2.30
19	2.18	5.01	0.436	6.4	8.0	0.7	12.3	72.5	2.00
12	2.18	5.25	0.698	7.3	6.9	0.6	10.2	75.0	1.94
28	4.06	5.33	0.690	8.5	8.3	0.6	13.8	68.8	2.25
33	-4.76+0.0	5.01	0.732	9.9	7.4	0.7	12.2	69.8	2.35

(a) See Table VII-1, Appendix VII for details

(b) Unstable operation.

TABLE 15

EFFECT OF COAL PARTICLE SIZE FOR THE GASIFICATION OF SUKUNKA COAL

Run # (a)	Nominal average size (mm)	Air to coal ratio (dry basis) (w/w)	Steam to dry coal ratio (w.w)	Gas Composition (T v/v) (dry basis)					Gas Gross calorific Value (MJ/m^3)
				H_2	CO	CH_4	CO_2	N_2	
36	2.18	4.58	0.020	4.6	7.0	1.2	12.4	74.8	1.85
39	4.06	4.66	0.021	4.3	7.1	1.0	15.3	72.4	1.75
37	2.18	4.67	0.331	5.6	6.6	1.2	12.3	74.4	1.92
40	4.06	4.71	0.261	6.2	7.0	1.0	15.7	70.0	1.97

(a) see Table VII-2, Appendix VII for details.

Although coal particle size did not have a particularly strong effect on the gas quality, it had an impact in the operation of the gasifier. When gasifying 0.53 mm and 0.95 mm Forestburg coal, the reactor was difficult to operate under stable conditions. In spite of adjustment of the operating variables, in most runs (See Table VII-1, Appendix VII) the temperature of the reactor drifted upwards or downwards at a rate in excess of 1 K/min. As the coal size increased the operation became more stable, and steady gasification with 2.18 mm and 4.06 mm coal was possible. Gasification of -4.76 mm + 0 mm Forestburg coal also went smoothly, which suggests that larger coal particles have a buffering capacity to absorb some variations in the operating conditions.

Effect of Air to Coal Ratio

The most important parameter in determining the gas calorific value appeared to be the air to coal ratio. For four different sizes of Forestburg coal a linear correlation between the gas calorific value and the air to coal ratio was found (Figure 11). In spite of some scatter and the fact that the data cover different temperatures and steam to coal ratios, a definite trend exists towards increased gas calorific value with decreasing air to coal ratios. Similar correlations have been found by other authors in the gasification of coal (34) and wood (35).

Insufficient data was generated to define such a correlation for the gasification of Sukunka coal, but the same general trend can be observed in Figure 12.

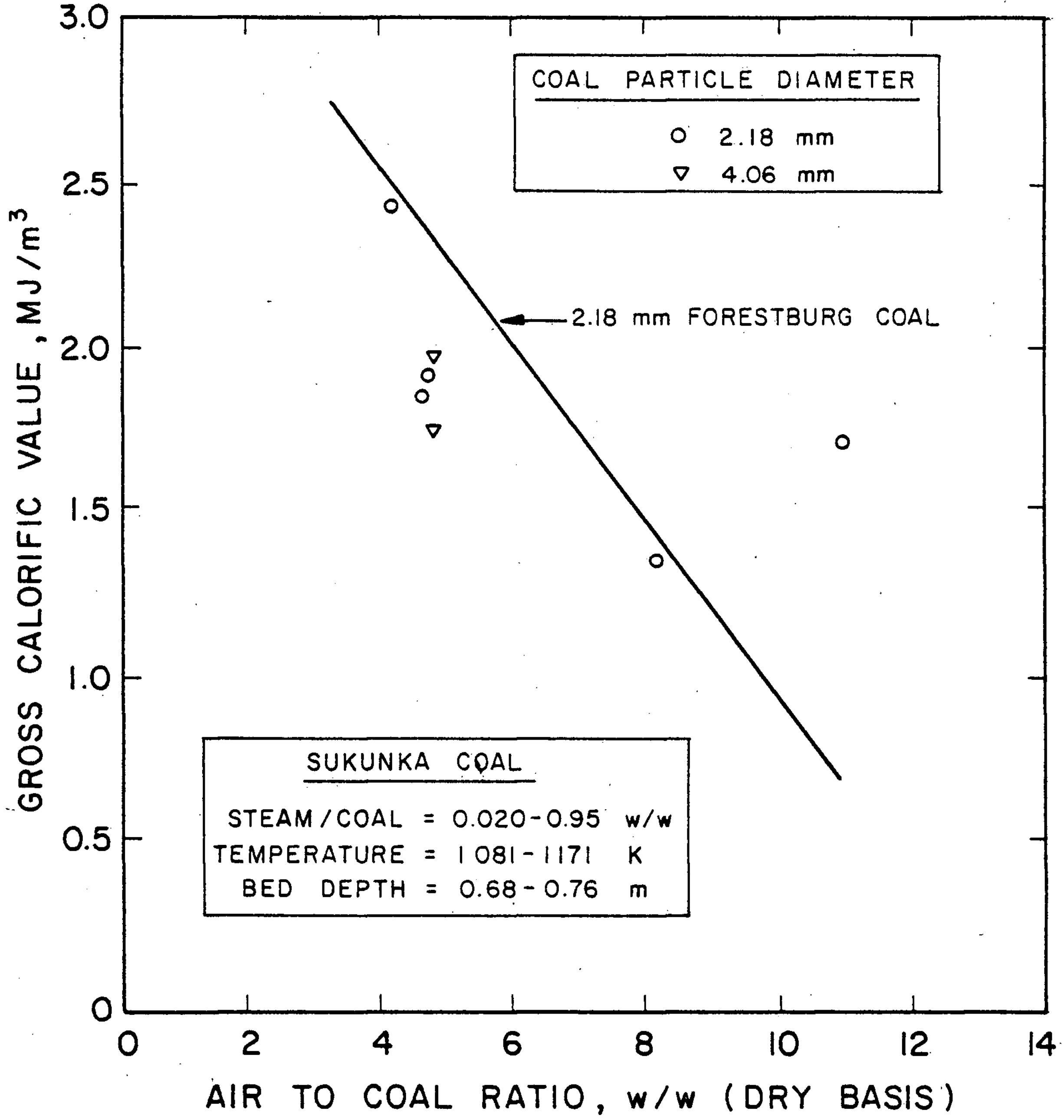


FIGURE 12: EFFECT OF AIR TO COAL RATIO ON THE GROSS CALORIFIC VALUE OF GAS PRODUCED FROM SUKUNKA COAL

The effect of the air to coal ratio on the gas composition is also of interest. Since the air to coal ratio can be changed by changing the coal feed rate, the air flow or both parameters, these effects are considered individually.

Since in coal gasification at atmospheric pressure the amount of methane formed by reaction of C and H_2 is insignificant at present operating temperatures and pressure, most of the methane produced is expected to be a product of pyrolysis, and therefore a direct function of the coal feed rate. Indeed, gasification of Forestburg coal showed that methane concentration in the gas increased with coal feed rate (Figure 13). The data in Figure 13 are scattered because not all other operating conditions are constant. However, if the methane production is calculated from the total volume of gas produced (see Table VII-7, Appendix VII) and plotted as a function of the coal feed rate, a good linear correlation is obtained (Figure 14). The slope of this line suggests that the methane yield of Forestburg coal is 2.5%. A reliable correlation could not be obtained for the gasification of caking coals due to the restricted number of experiments performed, but the available data (Table VII-8, Appendix VII) suggests a similar trend of increased methane production with coal feed rate, and also a higher methane yield for the Sukunka and Coleman coals (Figure 15).

Since steam feed rate may have an effect on the production of hydrogen and carbon monoxide, the effect of coal feed rate on the production of these gases should be studied at a constant steam to coal ratio (Table VII-9, Appendix VII). Figures 16 and 17 respectively show that the hydrogen and carbon monoxide production rates from Forestburg coal gasification at a steam to coal ratio of 0.32 - 0.34 are clearly depend-

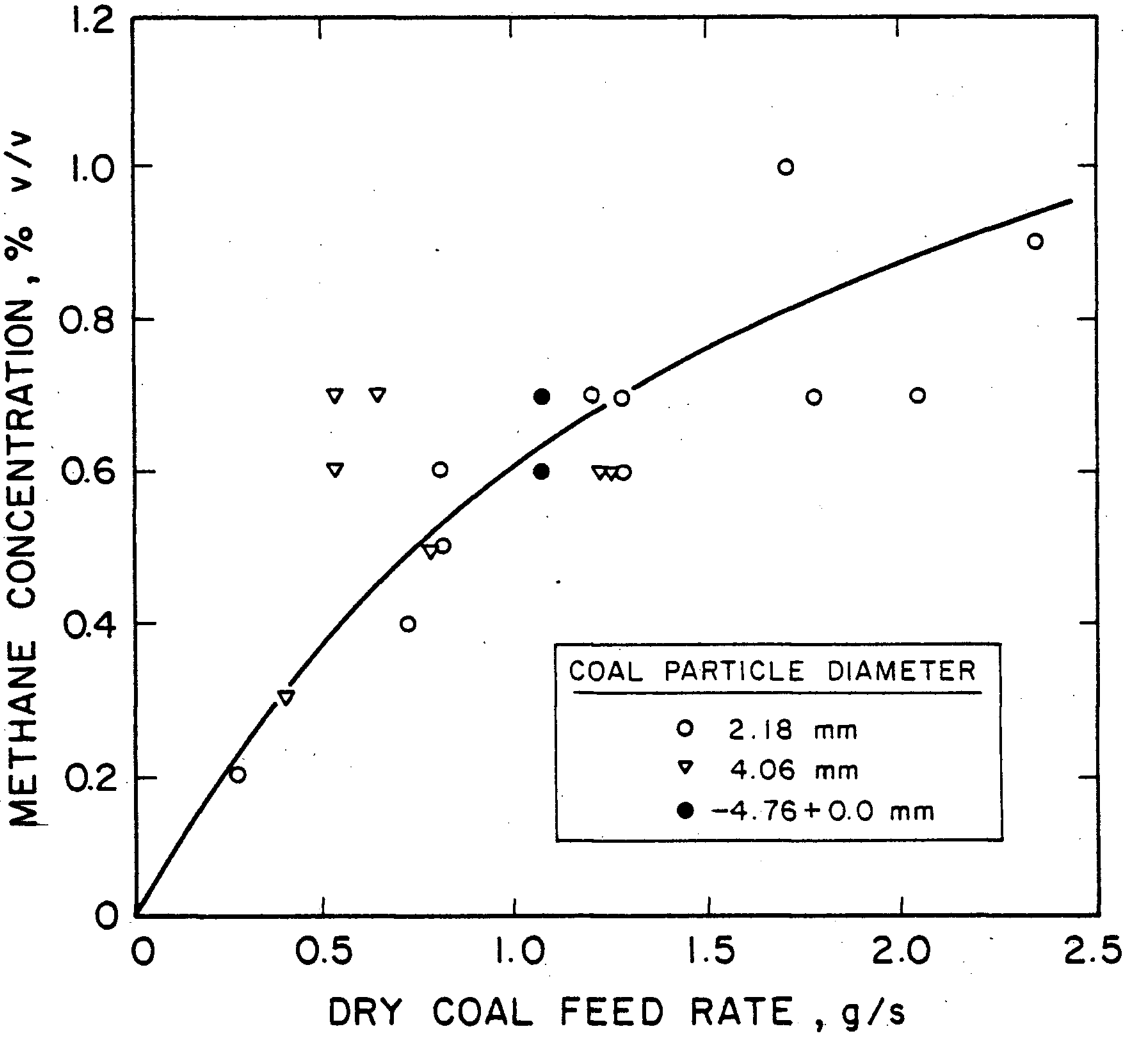


FIGURE 13: EFFECT OF COAL FEED RATE ON THE CONCENTRATION OF METHANE PRODUCED FROM FORESTBURG COAL

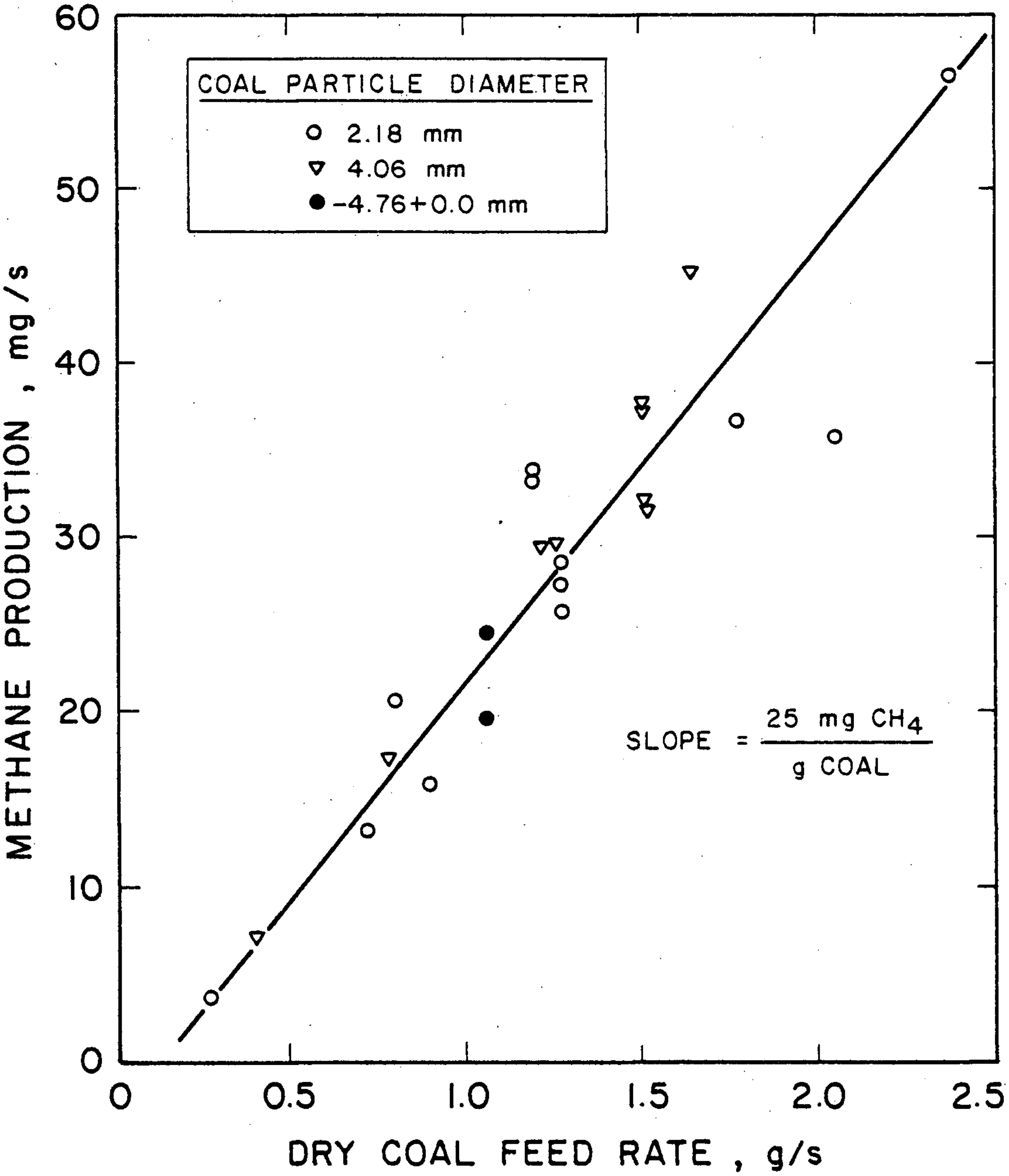


FIGURE 14: EFFECT OF COAL FEED RATE ON THE PRODUCTION OF METHANE FROM FORESTBURG COAL

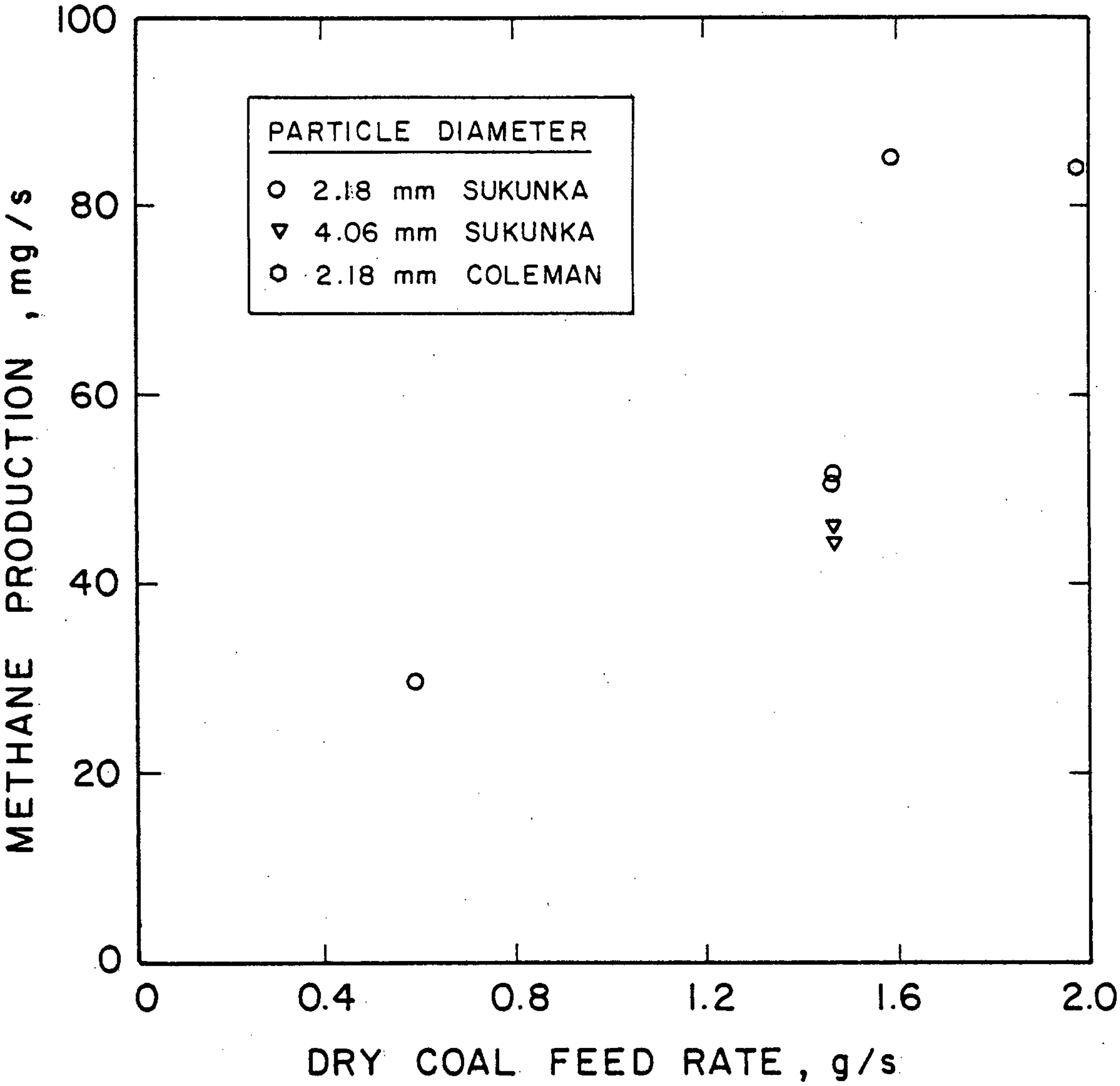


FIGURE 15: EFFECT OF COAL FEED RATE ON THE METHANE PRODUCTION FROM CAKING COALS

ent on the coal feed rate. From the slope of these curves, the hydrogen yield from Forestburg coal is calculated as 3.37% (Figure 16) which compares with a hydrogen content of 4.25% in the coal (Table 5). By a similar analysis, the carbon monoxide yield under these conditions is calculated at 52.6% (Figure 17).

Operating experience indicated that the effect of increasing the air flow into the reactor while keeping the coal and steam feed rates constant was to decrease both the hydrogen and carbon monoxide content of the gas while the carbon dioxide concentration increased. These findings are in line with results reported elsewhere (36) and are illustrated in Figure 18 for the gasification of Forestburg coal at a feed rate of 1.27 g/s and steam to coal ratio of 0.33. However, as discussed earlier in this chapter, altering the air flow rate while keeping the coal feed rate constant introduces a certain measure of instability into the operation of the reactor.

The combined effect on the gas composition of changing both the coal feed rate and the air flow rate at a constant steam to coal ratio is shown in Figures 19 and 20 for Forestburg and Sukunka coal respectively. Increasing in such a manner the air to coal ratio resulted in a decrease of the hydrogen and methane and an increase in the carbon dioxide content of the gas from both coals while the carbon monoxide content did not change appreciably. The overall result, as discussed earlier, (Figures 11 and 12) was then a decrease of the gas calorific value.

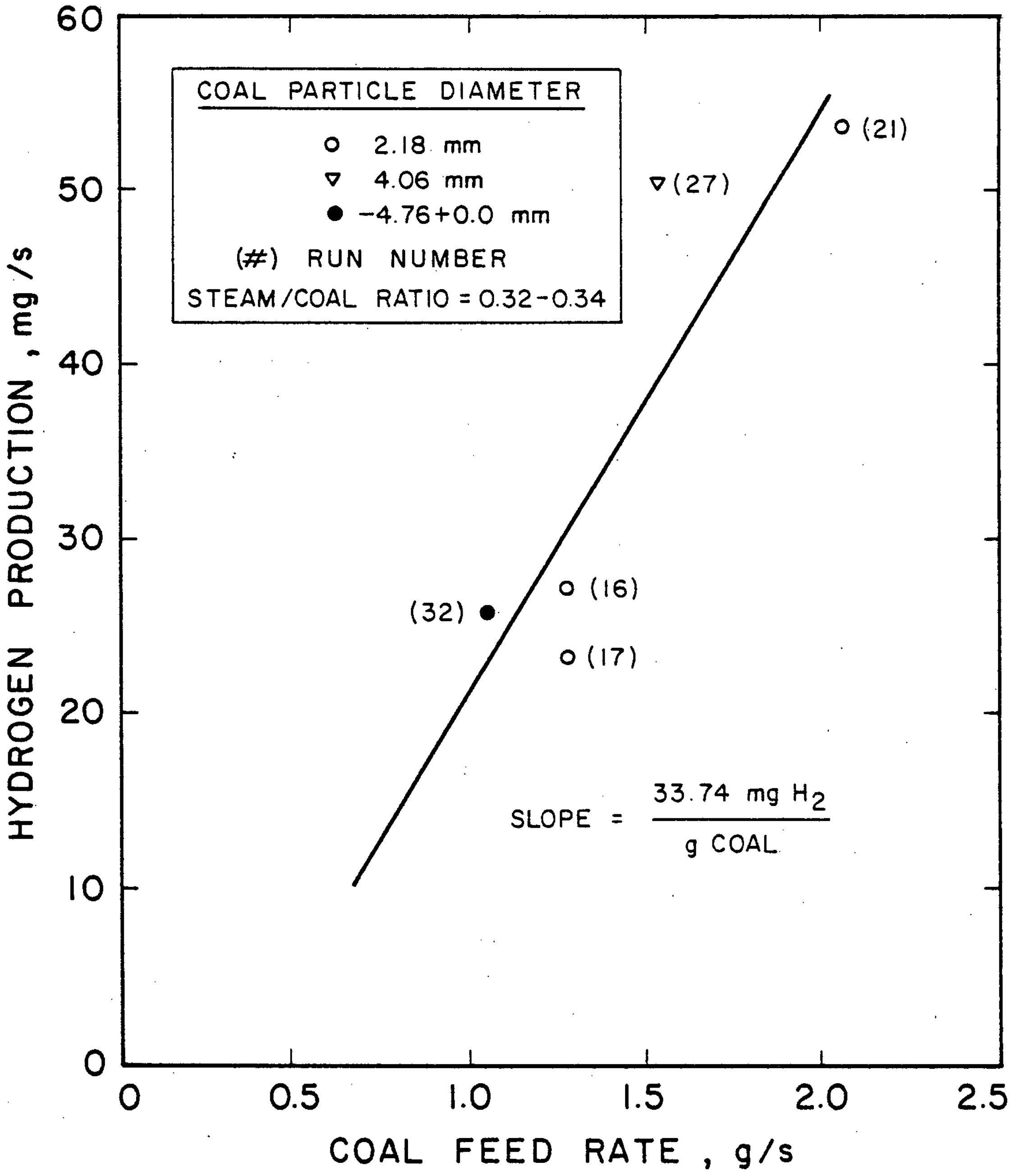


FIGURE 16: EFFECT OF COAL FEED RATE ON THE PRODUCTION OF HYDROGEN FROM FORESTBURG COAL

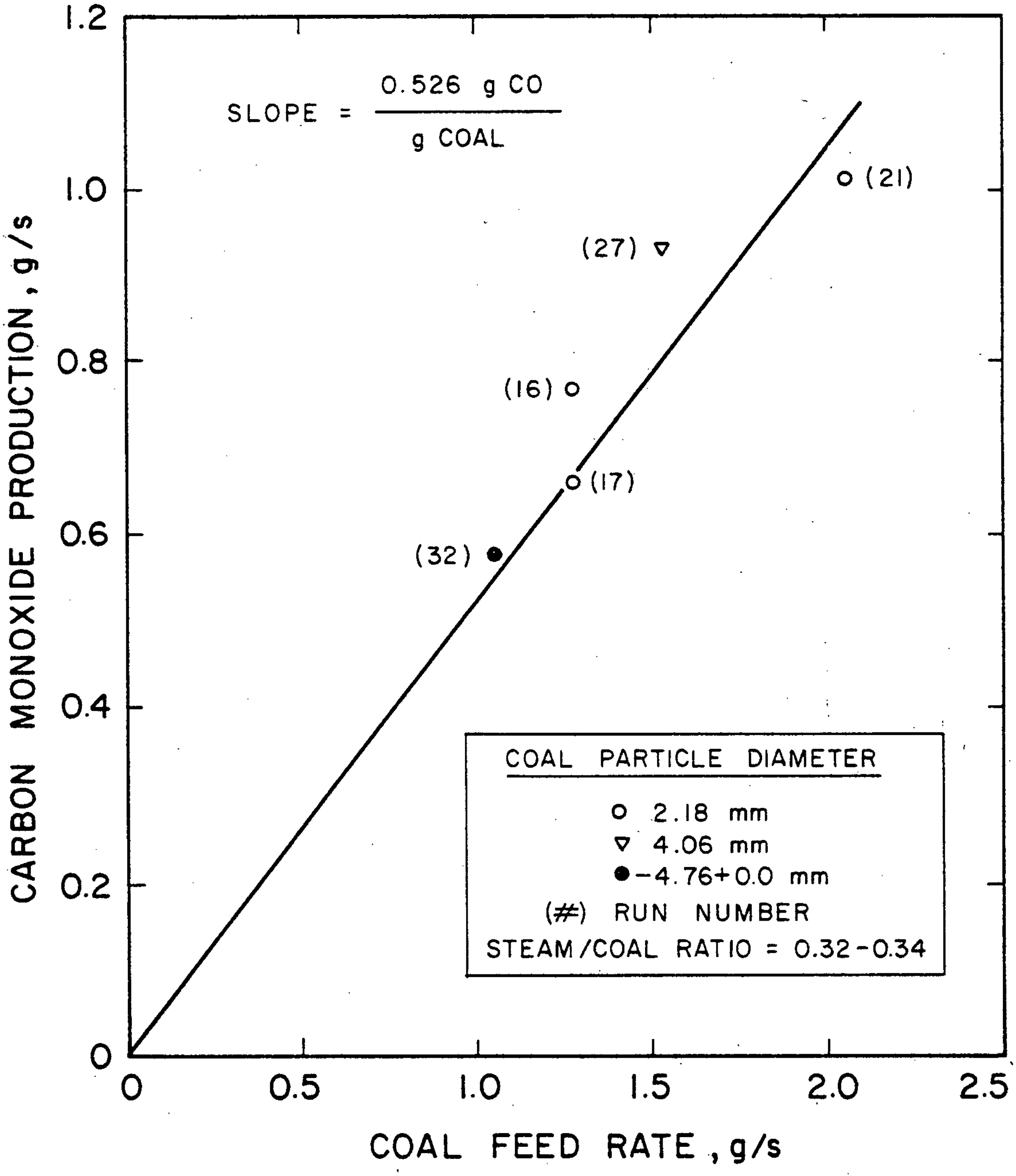


FIGURE 17: EFFECT OF COAL FEED RATE ON THE PRODUCTION OF CARBON MONOXIDE FROM FORESTBURG COAL

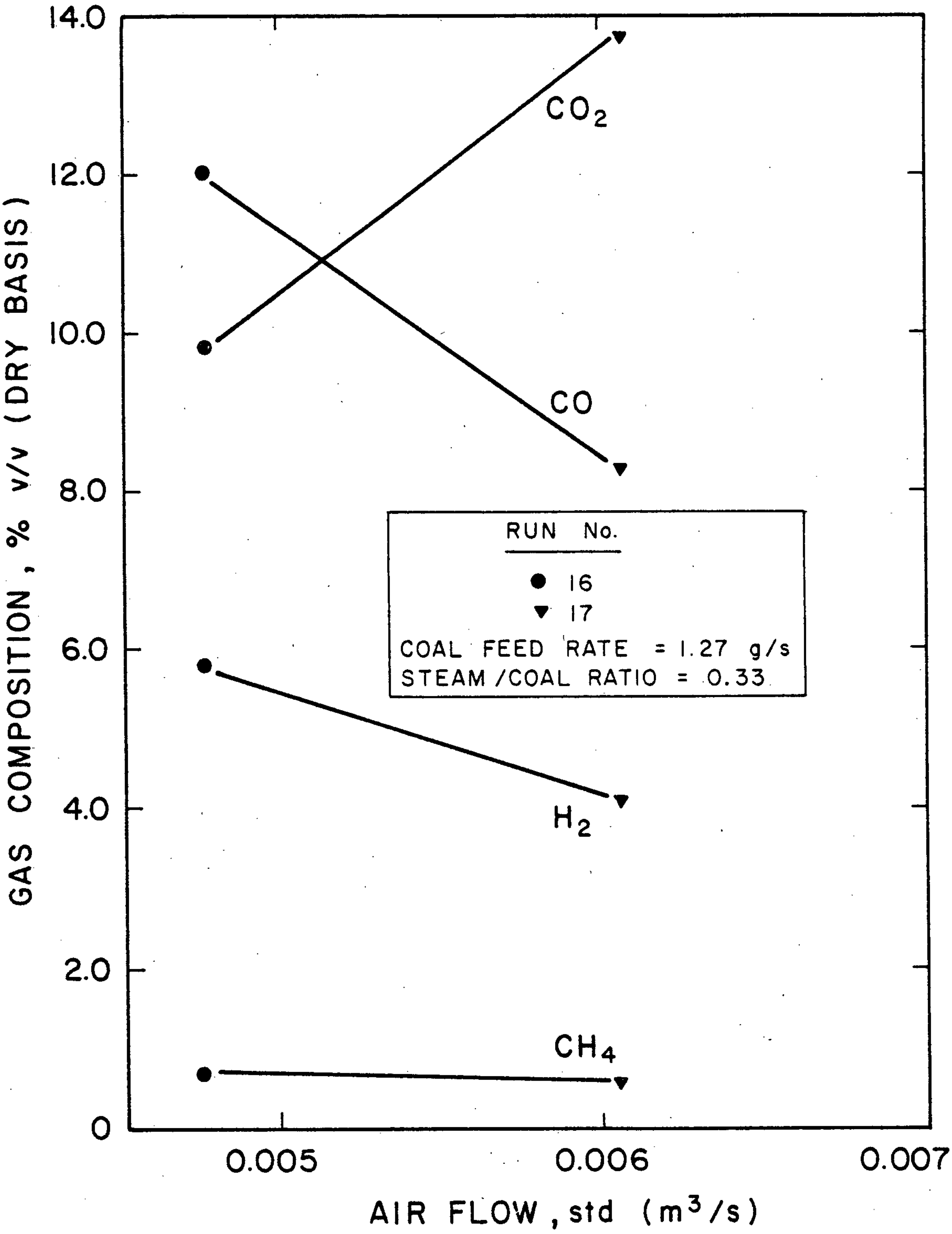


FIGURE 18: EFFECT OF AIR FLOW ON THE GAS COMPOSITION FROM FORESTBURG COAL AT A FIXED COAL FEED RATE

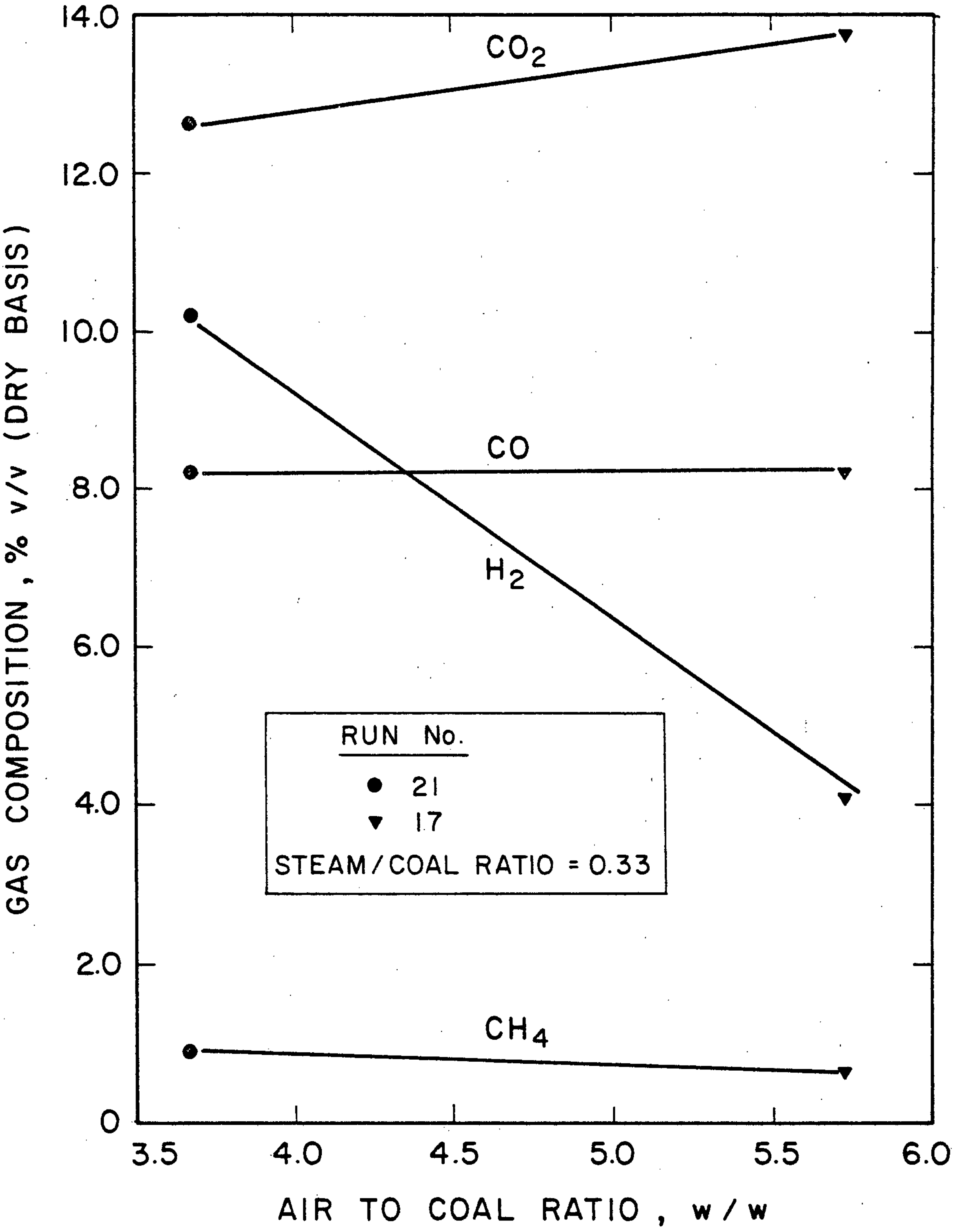


FIGURE 19: EFFECT OF AIR TO COAL RATIO ON THE GAS COMPOSITION FROM FORESTBURG COAL

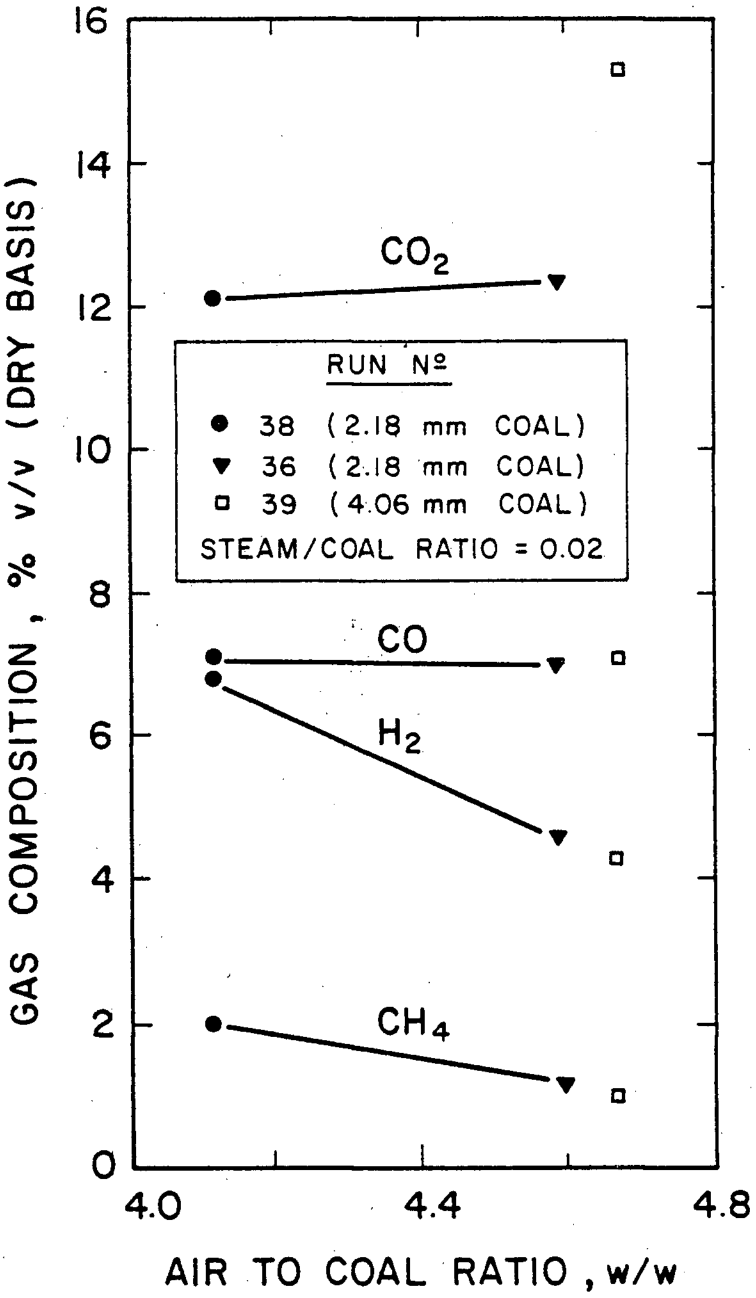


FIGURE 20: EFFECT OF THE AIR TO COAL RATIO ON THE GAS COMPOSITION FROM SUKUNKA COAL

Effect of the Steam to Coal Ratio

The effect of increasing the steam to coal ratio[*] was studied in Runs 27 to 30 using 4.06 mm diameter Forestburg coal. Other operating parameters were kept constant. The reactor temperature decreased as a consequence of the increased steam input from 1115 K at a steam to coal ratio of 0.33 to 1059 K at a steam to coal ratio of 1.10 (Table 16). As discussed earlier, a temperature change itself does not affect the gas composition. Although increasing the steam to coal ratio did not appreciably change the gas calorific value which remained in the range of 2.23 - 2.38 MJ/m^3 (Table 16), there was a marked effect on the gas composition. Figure 21 shows that the hydrogen and carbon dioxide content of the gas increased while the carbon monoxide content decreased. Methane concentration remained fairly constant. The same effect could be observed for other operating conditions with 2.18 mm Forestburg coal (compare for example runs 14 and 15 and runs 17 and 18, Table VII-1, Appendix VII), and for two different sizes of Sukunka coal (Table 17, Figure 22).

These results suggest that under the prevailing operating conditions increased concentration of water in the reactor favours the reaction

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

rather than the carbon steam reaction

$$C + H_2O \rightleftharpoons H_2 + CO$$

---

[*] Steam ratio = ratio of total water fed (i.e. steam + H_2O in coal + H_2O in air) to coal fed.

TABLE 16

EFFECT OF THE STEAM TO COAL RATIO IN THE GASIFICATION OF FORESTBURG COAL

Run #	Steam to Coal Ratio (w/w)	Average Reactor Temperature (K)	Gas Composition % (v/v) (dry basis)					Gross Calorific Value (MJ/m^3)
			H_2	CO	CH_4	CO_2	N_2	
27	0.332	1115	7.7	10.2	0.6	12.2	69.3	2.38
28	0.690	1085	8.5	8.3	0.6	13.8	68.8	2.25
29	0.826	1079	8.7	7.6	0.7	13.9	69.1	2.23
30	1.101	1059	10.0	6.9	0.7	14.1	68.3	2.30

Coal particle size: 4.06 mm

Coal feed rate: 1.516 g/s

Air flow: (6.67-6.71) x 10^{-3} std. m^3/s

Expanded bed depth 0.86 m

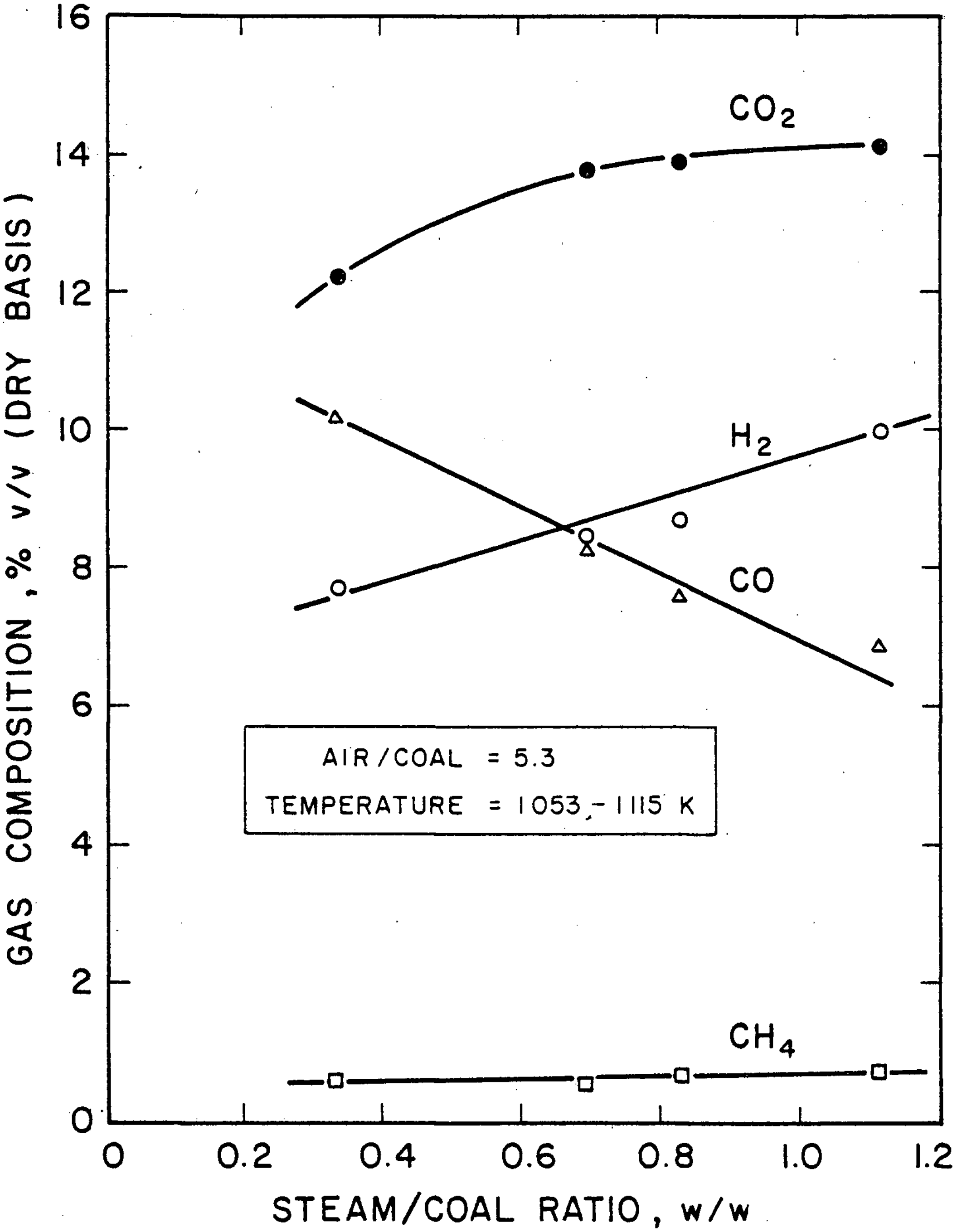


FIGURE 21: EFFECT OF THE STEAM TO COAL RATIO ON THE GAS COMPOSITION FROM 4.06 mm DIAMETER FORESTBURG COAL

TABLE 17

EFFECT OF THE STEAM TO COAL RATIO IN THE GASIFICATION OF SUKUNKA COAL

Run #	Steam to Coal Ratio (w/w)	Average Reactor Temperature (K)	Gas Composition % (v/v) (dry basis)					Gross Calorific Value (MJ/m^3)
			H_2	CO	CH_4	CO_2	N_2	
36	0.020	1132	4.6	7.0	1.2	12.4	74.8	1.85
37	0.331	1081	5.6	6.6	1.2	12.3	74.4	1.92
39	0.021	1160	4.3	7.1	1.0	15.3	72.4	1.75
40	0.261	1107	6.2	7.0	1.0	15.7	70.0	1.97

Coal particle size:	Runs 36 & 37 = 2.18 mm, Runs 39 & 40 = 4.06 mm
Coal feed rate:	1.46 g/s
Air flow:	(5.58-5.73) x 10^{-3} std m^3/s
Expanded bed depth:	0.76 m

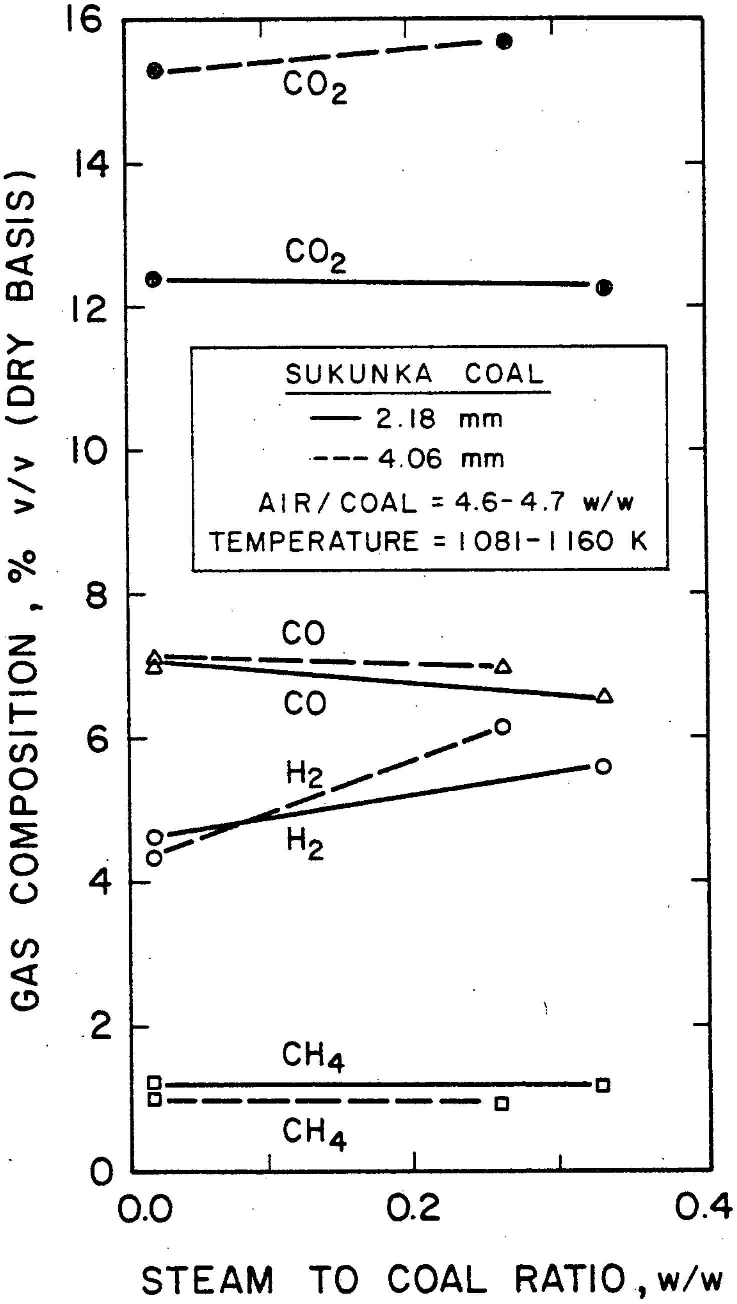


FIGURE 22: EFFECT OF THE STEAM TO COAL RATIO ON THE GAS COMPOSITION FROM SUKUNKA COAL

Effect of Bed Depth

Effects of the bed depth on the gas quality were masked by changes in other operating variables. However, inspection of the data (Table VII-1, Appendix VII) indicates that no noticeable improvement on the gas quality was obtained when the bed height varied from 0.5 to 0.86 m. This is in agreement with experimental results reported for the gasification of coal and chars at 1273 K (36). However, a deeper bed made the operation smoother and easier to control.

Carbon Content of the Bed

The carbon content of the bed is not an operating variable in itself, but rather a result of other operating conditions. The effect of coal reactivity on the bed carbon content was discussed earlier in this Chapter. In general, the average carbon bed content remained below 5% for stable gasification of Forestburg coal, (Table 18) and was not correlated with either the air to coal ratio (for conditions not exceeding the coal reactivity limit) or the temperature of the reactor. Also, there appears to be no correlation between the quality of the gas produced and the bed carbon content. Under equivalent operating conditions, the carbon content of a bed gasifying Sukunka coal was higher (Table 19) than for Forestburg coal (Table 18). This is probably due to a difference in coal reactivity.

Solids and Carbon Elutriation

Table 20 shows the size distribution of the solids elutriated from the bed for typical gasification conditions of 2.18 mm diameter Forestburg coal. Approximately 95% by weight of the solids had a size below 355 μm

TABLE 18

AVERAGE BED CARBON CONTENT DURING THE GASIFICATION OF FORESTBURG COAL

Run #	Average Coal Particle Size (mm)	Air to Coal Ratio (w/w)	Average Reactor Temperature (K)	Average Bed Carbon Content (%)
1	0.53	5.39	1033	0.44
5	0.95	5.10	1123-1088	2.29
8	0.95	6.65	1156	0.50
9	0.95	9.04	1172	0.20
10	2.18	12.16	1088	2.71
11	2.18	7.62	1050	2.90
12	2.18	5.25	1040	4.72
13	2.18	8.23	1061	3.96
14	2.18	6.59	1031	11.71
15	2.18	6.59	1031	9.89
16	2.18	4.46	1107	2.47
17	2.18	5.70	1165	0.97
18	2.18	5.94	1145	0.74
19	2.18	5.01	1049	10.36
21	2.18	3.66	1080	2.69
22	2.18	4.01	1026	16.6-31.3
23	4.06	10.36	1108	1.95
24	4.06	6.97	1096	4.44
25	4.06	6.23	1074	4.43
26	4.06	6.10	1065	10.67
27	4.06	5.28	1115	4.21
28	4.06	5.33	1085	4.31
29	4.06	5.32	1079	3.61
30	4.06	5.34	1153	4.01
31	4.06	5.68	1078	8.45
32	4.76-0.0	4.87	1125	1.52
33	4.76-0.0	5.01	1055	1.91

TABLE 19

AVERAGE BED CARBON CONTENT DURING THE GASIFICATION OF SUKUNKA COAL

Run #	Average Coal Particle Size (mm)	Air to Coal Ratio (w/w)	Average Reactor Temperature (K)	Average Bed Carbon Content (%)
34	2.18	8.12	1153	2.68
35	2.18	10.86	1171	21.4
36	2.18	4.58	1132	21.7
37	2.18	4.67	1081	27.2
38	2.18	4.11	1165	18.34
39	4.06	4.66	1160	16.2
40	4.06	4.71	1107	16.2

TABLE 20

TYPICAL SIZE DISTRIBUTION OF SOLIDS ELUTRIATED DURING THE GASIFICATION OF 2.18 mm FORESTBURG COAL (RUN NO. 20)

Particle Size (μm)	Weight Percentage (%)
+ 841	0.53
- 841 - 425	3.62
- 425 + 355	1.09
- 355 + 177	8.36
- 177 + 150	6.43
- 150 + 125	3.35
- 125 + 72	8.99
- 72 + 63	6.98
- 63 + 44	9.14
- 44	51.49

This is in good agreement with theoretical predictions where the maximum size of a solid particle elutriated is calculated from the particle terminal velocity at the prevailing operating conditions (30). From these calculations, the maximum diameter of an elutriated particle would be 727 µm for a charcoal particle and 130 µm for a silica particle.

Solids elutriation rates were expressed in terms of g of solids per g of dry coal fed into the reactor. Carbon elutriation rates were calculated from the solids elutriation rate and the carbon content of the solids elutriated. Although the data is somewhat scattered due to the fact that in some cases elutriation rates were measured as averages of several runs, in general, elutriation rates for narrowly sized Forestburg coal were in the range of 0.078-0.108 g/g coal, and seemed to increase with the average superficial velocity in the reactor and were largest for the smallest coal particle size (Table 21). When Forestburg coal sample of a wider size distribution which included fines (-4.76 + 0.0 mm) was gasified, elutriation of solids sharply increased to 0.187 g/g coal (runs 32 and 33, Table 21). The same trends with particle size and fluidization velocity were followed by carbon content of the solids elutriated, as with total solids elutriated. As a consequence the carbon elutriation rates (Table 21) increased with superficial velocity and decreased with increasing particle size.

Solids and carbon elutriation rates measured for the gasification of Sukunka coal were higher by a factor of 3 than for Forestburg coal, while for the run with Coleman coal they were intermediate between the two (Table 22). These differences existed in spite of the fact that coal ash content of all samples was similar (see Table 6) and may be an indication of the differences in coal reactivity. Indeed, during the

TABLE 21

SOLIDS AND CARBON ELUTRIATION DURING THE GASIFICATION OF FORESTBURG COAL

Run #	Average Coal Particle Size (mm)	Weighted Average Air Velocity at Reactor's Temp.(a) (m/s)	Solids Elutriation Rate (g/g coal)	Elutriated Solids Carbon Content (%)	Carbon Elutriation Rate (g/g coal)
1 & 4	0.53	1.10	0.167	62.31	0.104
2 & 3	0.53	0.80	0.102	55.14	0.056
5-9	0.95	1.15	0.078	29.50	0.023
10,12 & 13	2.18	0.79	0.084	36.79	0.031
11 & 19	2.18	1.13	0.086	50.44	0.043
14 & 15	2.18	1.25	0.093	57.99	0.054
16,17,18 & 21	2.18	1.22	0.091	29.66	0.027
22	2.18	1.52	0.093	58.48	0.054
23,25 & 31	4.06	1.21	0.083	38.14	0.032
24	4.06	0.93	0.093	30.01	0.028
26	4.06	1.28	0.091	43.35	0.039
27-30	4.06	1.35	0.108	40.96	0.044
32 & 33	-4.76+0.0	0.88	0.187	61.45	0.115

(a) When solids elutriation rate corresponds to the average of more than one run the average air velocity is the average of the air velocities weighted by the duration of the run.

TABLE 22

SOLIDS AND CARBON ELUTRIATION DURING THE GASIFICATION OF CAKING COALS

Run #	Average Coal Particle Size (mm)	Weighted Average Air Velocity At Reactor's Temp.[a] (m/s)	Solids Elutriation Rate (g/g coal)	Elutriated Solids Carbon Content (%)	Carbon Elutriation Rate (g/g coal)
34	2.18	0.80	0.227	81.66	0.185
35	2.18	1.16	0.266	85.17	0.227
36-38	2.18	1.17	0.272	80.33	0.219
30-40	4.06	1.21	0.276	61.88	0.171
41	2.18	1.17	0.139	71.86	0.100

Runs 34-40 used Sukunka Coal

Run 41 used Coleman Coal

(a) When solids elutriation rate corresponds to the average of more than one run, the average air velocity is the average of the air velocities weighted by the duration of the run.

gasification of Sukunka coal, the bed had a much higher carbon content (and thus probably a higher content of fine and light particles) than for Forestburg coal. Therefore, the elutriation rates will be also higher with the former coal.

Tar Production

Tar content measurements in the gas on some selected runs (Table 23) showed that the tar produced during gasification was notably higher for the caking coals. Tar production was highest from Sukunka coal, followed by Coleman coal while Forestburg coal yielded the lowest amount of tar. These measurements also showed a strong effect of the particle size with the smaller 2.18 mm coal producing as much as three times more tar than 4.06 mm coal. This last effect was verified for both the Forestburg and Sukunka coal. Unexpectedly, a Forestburg coal sample including fines (-4.76 mm + 0.0 mm) (see runs 31 and 32) and having a nominal average size of 1.03 mm yielded an amount of tar intermediate between the 2.18 mm and 4.06 mm Forestburg coal. Tar levels are also dependent on the coal feed location. If the coal had been fed into the bottom of the bed, lower tar levels would be expected for both coals.

TABLE 23

GAS TAR CONTENT FOR THE GASIFICATION OF SOME WESTERN CANADIAN COALS

Run #	Coal Type	Coal Particle Size (mm)	Dry Coal Feed Rate (g/s)	Air Flow Std. (m^3/s)	Steam Feed Rate (g/s)	Average Reactor Temperature (K)	Gas Tar Content (g/m^3)	Tar Production (mg/g coal)
14 & 15	Forestburg	2.18	1.185	6.52×10^{-3}	0.25-0.54	1031	0.841	5.17
24	Forestburg	4.06	0.782	4.53	0.549	1096	0.274	1.80
32 & 33	Forestburg	4.76-0.0	1.059	4.35	0.0433	1090	0.434	2.11
37-38	Sukunka	2.18	1.46-1.58	5.57	0.0455	1126	3.307	14.16
39-40	Sukunka	4.06	1.462	5.71	0.0358	1134	0.913	4.25
41	Coleman	2.18	1.968	5.72×10^{-3}	0.0	1096	2.122	7.22

CHAPTER V - DISCUSSION OF RESULTS

MASS BALANCES

In order to verify the experimental data and gain some insight into the gasification reactions, overall and detailed mass balances were carried out for 38 gasification runs. The method followed in doing such balances and a detailed sample calculation are found on Appendix VIII. Results are summarized in Tables VIII-1 to VIII-4, Appendix VIII.

Overall Mass Balances

The overall mass balances (Table VIII-1, Appendix VIII) showed that in general the outputs were short of the inputs by 2-20%, but in some cases the outputs were larger than the inputs. Overall, the average discrepancy in the mass balances was - 4.7 ± 9.2%, which is within acceptable margins of experimental error.

Inaccuracies introduced in the mass balances which contributed to the imbalance between outputs and inputs are discussed below in order of importance.

1. All mass balances were carried out at average operating conditions (Table VII-1 and VII-2, Appendix VII) and therefore do not consider variations of parameters such as air flow, reactor temperature, and gas composition, which actually occurred during a run. Calculations done at specific times during a run yielded better balances, but are not included in this thesis for the sake of brevity.

2. The water output of the reactor was measured indirectly, by difference between the measured wet gas flow and the calculated dry gas flow (see Appendix VIII). Some degree of inaccuracy is introduced here since in most gasification runs the bed was vigorously agitated or slugging, which caused considerable oscillations of the water manometer measuring the pressure drop across the orifice plate used to determine the wet gas flow.

3. Inert ash particles having a terminal velocity equal to or higher than the prevailing gas superficial velocity will not be elutriated from the reactor until such time that their size is sufficiently reduced by attrition. Therefore, at any given time there is likely to be a net accumulation of ash in the reactor. This was confirmed by the ash balances (Table VIII-4, Appendix VIII) which invariably show an ash input larger than the ash output. This accumulation term (which will increase with the coal particle size and decrease with the air flow) is not considered in the overall mass balances shown on Table VIII-1, Appendix VIII, and causes a shortfall in the total mass output.

4. In most cases, the reported solids elutriation rates represent averages of 2 - 4 gasification runs carried out under relatively similar operating conditions. Although some inaccuracy is introduced by this procedure, the contribution of the elutriated solids to the total mass output is below 5%.

5. Although tar measurements were made only for a few selected runs, these showed that the tar contribution to the total mass

output is below 1% and therefore of little consequence in the overall mass balances.

Water Balance

Water balances (Table VIII-1, Appendix VIII) indicated that in only 12 of 38 gasification runs the water output was lower than the water input. Of these, four corresponded to obvious inaccuracies in the mass balances since the wet gas flow measured was lower than the calculated dry gas flow. The balance of these 12 runs corresponded to a variety of operating conditions and it did not appear to be a specific set of conditions that favored water dissapearance in the reactor. It is therefore safe to say that in general water was formed in the reactor. Since the experimental data provides evidence of the reaction

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

occurring (see Chapter IV, effect of steam) the observed excess water can only be explained by assuming that the other water consuming reaction:

$$C + H_2O \rightleftharpoons CO + H_2$$

does not proceed to any appreciable extent under the prevailing operating conditions, and that water is formed through the reaction of oxygen and hydrogen and through the combustion of hydrocarbons at higher rates than it disappears. This is supported by both the kinetics and thermodynamics of the principal reactions involved in a gasification system (see Chapter I).

Hydrogen Balance

An important part of the hydrogen entering and leaving the reactor is water bound, and therefore total hydrogen mass balances (Table VIII-2, Appendix VIII) are very sensitive to inaccuracies in the water determination discussed earlier. Accordingly, an imbalance between total hydrogen input and output was generally obtained. Note however, that the impact of such imbalances in the overall mass balance is minor. On the other hand, the amount of hydrogen entering the reactor with the dry coal and the amount of hydrogen leaving the reactor with the dry gas are subject to less error since they only depend on coal and gas analyses and measurement of the coal feed rate and dry gas flow out of the reactor. Comparison of these two quantities (Table VIII-2, Appendix VIII) shows that in general, the hydrogen in the gas is similar to or lower than the hydrogen present in the dry coal feed. This again suggests that the formation of hydrogen through the carbon steam reaction

$$C + H_2O \rightleftharpoons H_2 + CO$$

is not very important, and that hydrogen is available for the formation of water.

Oxygen Balance

A total oxygen balance (Table VIII-3, Appendix VIII) is also very sensitive to the inaccuracies in water input and output determinations and is not of much value in interpreting the present experimental data. However, in all gasification runs, the oxygen gas content was lower than the oxygen entering the reactor with the air except in 8 runs where they were similar. In all but two runs, the oxygen contained in the dry gas was lower than the oxygen entering the reactor with the dry coal and the air. This provides additional evidence that in general some oxygen goes into the formation of water.

Carbon Balance

In general, the total carbon output was equal to or lower than the total carbon input (Table VIII-4, Appendix VIII). The major inaccuracies introduced in the carbon balances was in the determination of the carbon elutriation rates.

ENERGY BALANCES

A detailed energy balance in the fluidized bed reactor would require the knowledge of the amounts of gas produced by each of the pyrolysis, combustion and gasification reactions and consideration of the different heats of reaction at the prevailing operating conditions. Additionally, a measurement of heat losses from the reactor would be necessary. Since much of this data is not available, and in order to simplify the analysis, a simplified energy balance was carried out for various operating conditions for the gasification of Forestburg and Sukunka coals. The method considered the reactor as a "black box" and compared the heat content of the inputs (coal, air and steam) with the heat content of the outputs (sensible and calorific value of the gas, heat content of elutriated carbon and heat content of steam). The heat losses from the reactor were then calculated by difference. A detailed sample calculation is presented on Appendix IX, while energy balances for selected gasification runs are calculated in Table IX-1, Appendix IX).

The energy losses calculated for 17 gasification runs varied between 10.3% and 37.2% of the total energy input. Such variations are not only due to experimental errors and variations in the mass balances already discussed, but also to the dependence of the percentage heat losses on the temperature of the reactor and on the heat input. The mean energy losses and standard deviation were 25.8% ± 8.8%, as would be expected for a relatively small externally insulated fluidized bed. The substantial energy losses have no doubt an impact in the quality of the gas produced, and on the efficiency of the system since an important portion of the energy generated by the combustion of carbon, hydrocarbons and hydrogen will be lost to the environment instead of being available for the endothermic gasification reactions. In other words, more carbon than otherwise is necessary goes into the formation of CO_2 rather than combustible gases. The effect that reduced heat losses could have in the gasifier operation is illustrated on Figure 23. For run No. 38, which has a rather typical energy loss of 28.5%, the temperature of the gases leaving the reactor has been calculated assuming that the energy losses could be cut down to different percentages to a loss of 10%, typical of a larger scale operation. This is made under the assumption that the increased reactor temperature will not alter the specific heat and the mass of the gases leaving the reactor. Of course, this is not quite true since the change in temperature will increase the rates of the different reactions involved and thus change the gas composition. However, the point to be made here is that if the reactor temperature increases, the endothermic reactions are likely to proceed at higher

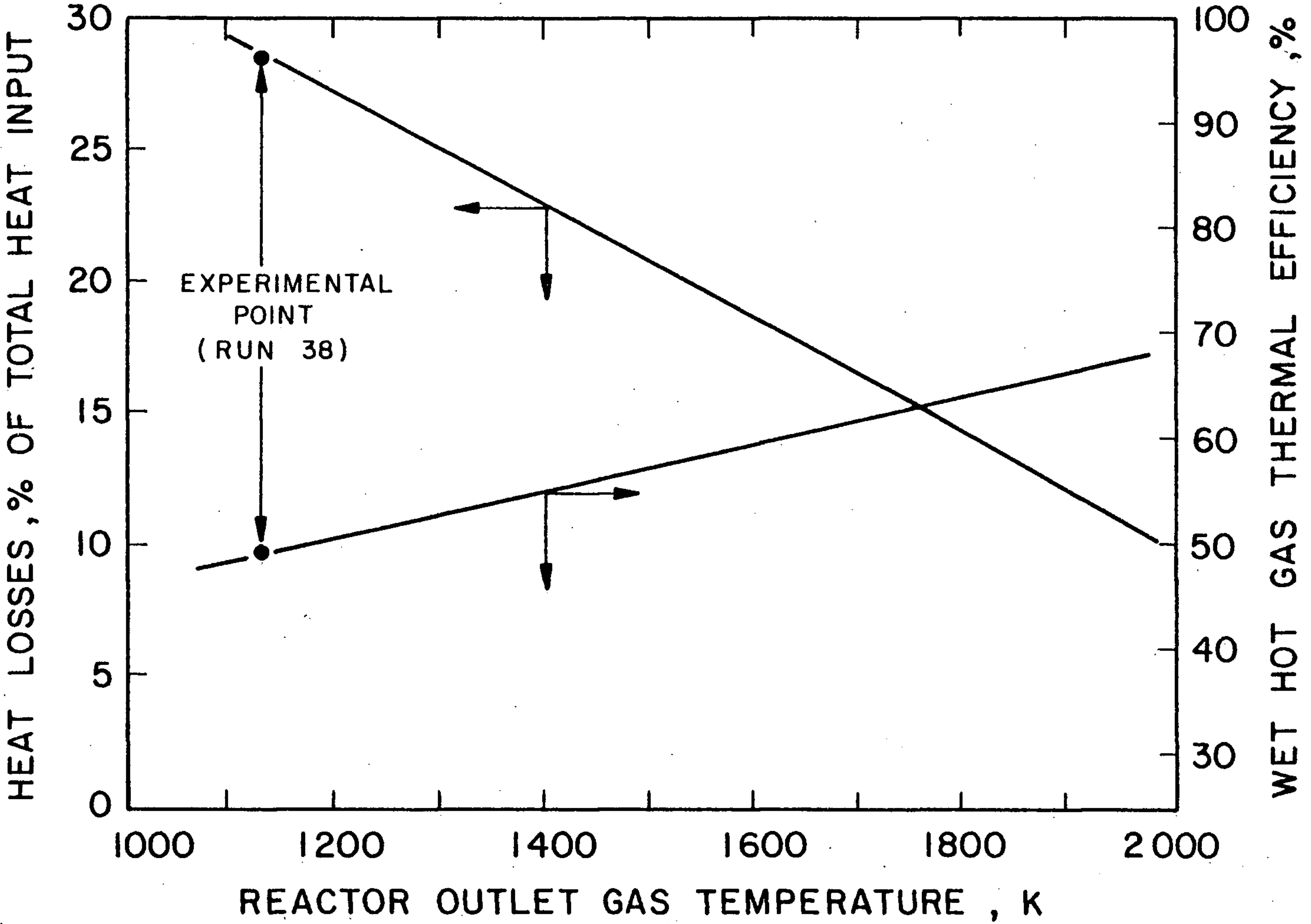


FIGURE 23: CALCULATED EFFECT OF ENERGY LOSSES ON THE TEMPERATURE OF THE GAS LEAVING THE COAL GASIFIER

rates (for instance, the reaction of carbon and steam which is known to become important only at temperatures over 1300 K). This in turn will tend to consume the extra heat and bring the reactor temperature down below the calculated value, but the quality of the gas produced could improve considerably. This factor is important when comparing results of this investigation with those from larger scale equipment where heat losses are low or with small scale studies where stronger gases are produced, by providing the reaction heat for example by external electrical heating. In the latter case one could gasify coal by steam only with no nitrogen to dilute the off-gas, and consume little coal in CO_2 production.

THERMAL EFFICIENCIES

The thermal efficiency of a gasification system is a measure of the efficiency with which the energy contained in the coal is transformed into a gaseous fuel. Two kinds of thermal efficiencies are defined here.

The first is the thermal efficiency of the clean cool gas, and is simply the ratio between the combustion heat at 288.6 K of the clean (no solids, no tars) gas produced by a unit mass of coal and the heat value of that unit mass of coal, i.e.

$$\eta_{cc} = \frac{\text{combustion heat of clean gas}}{\text{combustion heat of coal.}} \times 100$$

The clean gas thermal efficiency for the gasification of various sizes of Forestburg coal (Table 24) was generally between 40 and 50% , which is similar to the efficiency obtained for gasification of the same coal in a spouted bed (1). The data for this coal shows a general trend towards improved efficiency with increasing coal particle size. This appears to be related with a decrease of the amounts of carbon elutriated

TABLE 24

THERMAL EFFICIENCIES AND % USEFUL CARBON FOR THE GASIFICATION OF SOME WESTERN CANADIAN COALS

Run #	Thermal efficiency (%)		Useful carbon (%)
	η_{cc}	η_{raw}	
1	25.7		55.8
2	39.0		78.6
3	26.3		70.8
4	24.5		54.2
Mean ± S.D.	28.9 ± 6.8		64.9 ± 11.8
5	44.4		90.8
6	73.3		94.1
7	54.2		92.4
8	35.8		88.8
9	11.9		72.1
Mean ± S.D.	43.9 ± 22.7		87.6 ± 8.9
10	16.2		73.6
11	41.6		83.4
12	36.8	60.0	85.9
13	61.0		91.0
14	51.4	83.2	83.1
15	46.4	77.9	81.6
16	40.1	73.0	88.6
17	35.2	73.8	84.6
18	34.5	84.3	86.9
19	36.0	57.5	81.2
21	38.1	64.0	88.1
22	39.1	58.5	78.8
Mean ± S.D.	39.7 ± 10.7	70.2 ± 10.5	83.9 ± 4.8
23	30.2	61.1	82.6
24	62.9		91.8
25	57.9	70.6	90.5
26	66.1		89.9
27	47.3		84.8
28	45.5		84.3
29	44.9		84.0
30	46.9		84.6
31	61.1	80.2	90.8
Mean ± S.D.	50.9 ± 10.9	70.6 ± 9.6	87.0 ± 3.6
32	39.3		64.0
33	45.0		66.9
Mean ± S.D.	42.2 ± 4.0		65.5 ± 2.1
36	25.2	53.8	40.8
37	26.8	58.3	42.4
38	31.1	49.0	45.3
Mean ± S.D.	27.7 ± 3.1	53.7 ± 4.7	42.8 ± 2.3
39	25.8	48.2	47.0
40	29.0	52.8	51.2
Mean ± S.D.	27.4 ± 2.3	50.5 ± 3.3.	49.1 ± 3.0
41	26.8		61.3

Runs 1- 4: 0.53 mm Forestburg coal
Runs 5- 9: 0.95 mm Forestburg coal
Runs 10-22: 2.18 mm Forestburg coal
Runs 23-31: 4.06 mm Forestburg coal
Runs 32-33: 4.76 mm 0.C mm Forestburg coal
Runs 36-38: 2.18 mm Sukunka coal
Runs 39-40: 4.06 mm Sukunka coal
Runs 41: 2.18 mm Coleman coal

from the bed as the coal size increases, leaving more carbon in the bed for gasification.

The clean gas thermal efficiency for gasification of Sukunka and Coleman coal was approximately 27% (Table 24) i.e. substantially lower than for Forestburg coal. This difference is larger than the one expected from the differences in gas quality for these two coals (Tables 12 & 13), and seems to confirm the importance of carbon elutriation rates, which were substantially higher for the caking coals (see Tables 21 & 22).

A second thermal efficiency for the system, that of the wet hot gas can be defined as the ratio between the total heat value of the gas (combustion heat + sensible heat + sensible heat of steam) over the energy input to the system (heat content of coal, air and steam fed).

$$\eta_{\text{raw}} = \frac{\text{heat content of hot wet gas}}{\text{total heat input}}$$

Such efficiencies are particularly meaningful for in-situ industrial applications where the "dirty" gas can be used directly.

Computation of the raw gas thermal efficiency requires carrying out a complete energy balance, and values for selected runs are presented in Table 24. Note that since in this work the tar content of the gas was determined only on selected runs, and that the tars were not analysed, the thermal efficiency of the raw gas as defined here does not include the heat value of the tar. The average raw gas thermal efficiencies were 70 - 71% for the gasification of Forestburg coal and 50 - 54% for Sukunka coal. Again, the difference is due to the energy losses to the system in the form of elutriated carbon which were substantially higher for Sukunka coal (see Table IX-1, Appendix IX)

Full scale gasifier design calculations suggest clean gas efficiencies of 70 - 80% and hot gas efficiencies of 85 - 90%. These higher efficiencies are due to the production of a better quality gas in the absence of large heat losses, and also to the recycle of carbon elutriated from the bed.

USEFUL CARBON RATIOS

Yet another way of defining the efficiency of a gasification system is computing the useful carbon ratio or % useful carbon. This is defined(34) as the ratio between the total useful carbon, C_u, over the carbon available for gasification, C_a. The useful carbon is defined as the carbon converted to synthesis gas, C_s, ($CO + H_2$) plus the carbon converted to gaseous hydrocarbons, C_h.

then, $C_u = C_s + C_h$

On the other hand the carbon available for gasification is defined as

$$C_a = C_s + C_h + C_l + C_e$$

where C_e = unreacted carbon = elutriated carbon,

and C_l = carbon in liquid hydrocarbons,

but from the stoichiometry of the gasification reactions:

$$C + H_2O \rightleftharpoons CO + H_2$$

and $$C + 2H_2O \rightleftharpoons CO_2 + 2H_2$$

$$C + CO_2 \rightleftharpoons 2\ CO$$

it is evident that all of these reactions produce two moles of synthesis gas per mole of carbon reacted.

Accordingly:

$$C_s = S/_2$$

where S = moles of CO + H_2 produced.

Therefore:

$$\mu c = \frac{C_u}{C_a} = \frac{C_s + C_h}{C_s + C_h + C_1 + C_e}$$

$$\mu c = \frac{[\frac{1}{2}\ (\text{moles } H_2 + \text{moles CO}) + \text{moles } CH_4] \times 12}{[\frac{1}{2}(\text{moles } H_2 + \text{moles CO}) + \text{moles } CH_4] \times 12 + \text{elutriated carbon}}$$

If the amount of carbon in the liquid hydrocarbons is neglected and the sole gaseous hydrocarbon is methane, the useful carbon ratio is therefore a means of separating the combustion from the gasification reactions and assessing how much of the carbon that does not go into the formation of CO_2 goes into the formation of combustible gases. Consequently, the useful carbon ratio is a strong inverse function of the amount of elutriated carbon, and points out the importance of heat losses to the environment through elutriated carbon previously discussed in this chapter.

This is illustrated by Table 24 showing that the useful carbon ratio was higher for Forestburg coal than for the caking coals where carbon elutriation ratios are higher. For the Forestburg coal, the useful carbon ratios were lowest for the smallest particle size (0.53 mm) and the coal containing fines (-4.72 + 0.0 mm) which exhibited the highest carbon elutriation rates.

It should be noted that no attempt was made to recycle the elutriated solids in this work. In practice these fines would either be recycled to the gasifier or burned to raise steam for gasification, and would not contribute so significantly to an overall plant efficiency.

EQUILIBRIUM CONSIDERATIONS

An attempt to clarify which are the preponderant reactions in the present coal gasification system is presented below by consideration of some of the equilibrium data available in the literature.

Experimental evidence has been provided in Chapter IV (see Figure 14) that pyrolysis reactions are the main if not the sole source of methane. The observed effect of the steam ratio in the gas composition (see Chapter 4) and the mass balances points to the fact that in the present system, hydrogen is also mainly a product of coal rather than steam decomposition.

On the other hand, the energy balances showed that in a gasifier characterized by large heat losses, the combustion of carbon to CO_2 is a predominant factor in determining the gas composition.

Consider now the following reactions in a carbon steam system:

$$C + H_2O \rightleftharpoons CO + H_2 \qquad (1)$$

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad (2)$$

$$C + CO_2 \rightleftharpoons 2CO \qquad (3)$$

$$C + 2H_2 \rightleftharpoons CH_4 \qquad (4)$$

and compare the experimental product to steam ratios obtained (Table 25)) with the corresponding equilibrium ratios (Figure 24, after reference 29)

TABLE 25

PRODUCT RATIOS FOR THE GASIFICATION OF SOME WESTERN CANADIAN COALS

Run #	Reactor average Temperature (K)	Product ratios			
		CO_2/H_2O	CH_4/H_2O	H_2/H_2O	CO/H_2O
10	1088	0.445	0.018	0.039	0.070
11	1050	0.786	0.028	0.372	0.441
12	1040	0.539	0.032	0.386	0.364
13	1061	0.690	0.032	0.466	0.524
14	1031	0.614	0.039	0.484	0.406
15	1031	0.708	0.046	0.477	0.477
16	1107	0.289	0.021	0.175	0.355
17	1165	0.576	0.025	0.171	0.346
18	1145	0.360	0.014	0.115	0.143
19	1049	1.343	0.076	0.699	0.873
21	1080	0.527	0.028	0.341	0.458
22	1026	0.976	0.070	0.790	0.635
23	1108	1.984	0.044	0.415	0.503
24	1096	0.301	0.023	0.442	0.433
25	1074	1.335	0.060	1.006	0.886
26	1065	1.404	0.093	1.789	1.681
27	1115	0.503	0.025	0.318	0.421
28	1085	0.411	0.018	0.253	0.277
29	1079	0.443	0.022	0.278	0.242
30	1053	0.354	0.018	0.251	0.173
32	1125	0.233	0.013	0.139	0.223
33	1055	0.208	0.012	0.169	0.126
36	1132	0.397	0.038	0.147	0.224
37	1081	0.286	0.028	0.130	0.154
38	1165	1.427	0.236	0.802	0.837
39	1160	1.033	0.068	0.290	0.479
40	1107	0.847	0.054	0.334	0.378
41	1096	0.596	0.104	0.388	0.427

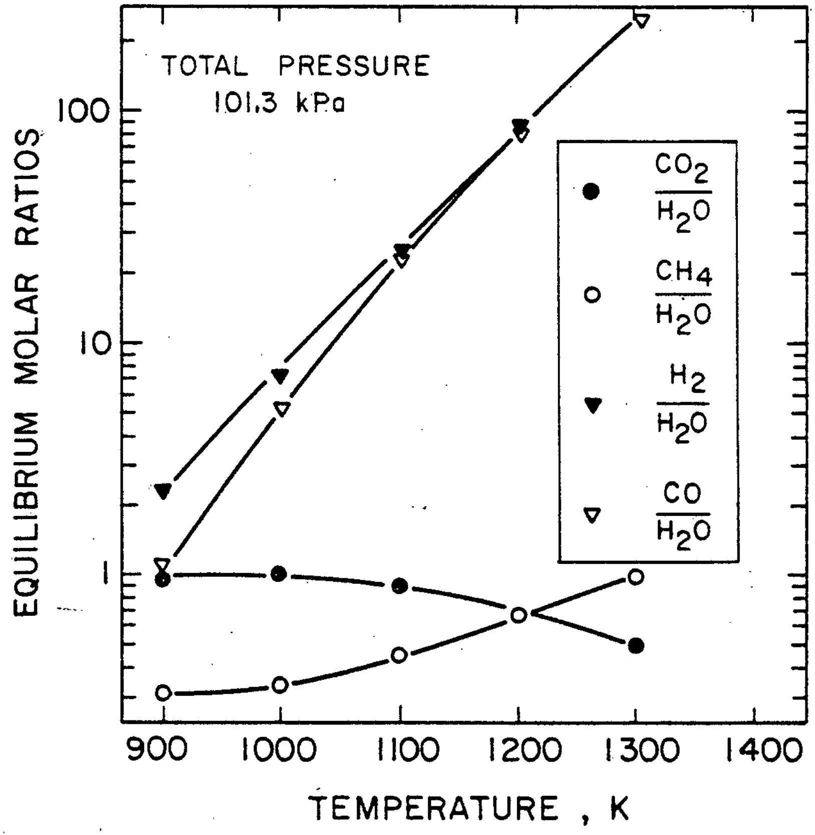


FIGURE 24: EQUILIBRIUM PRODUCT-STEAM RATIOS AS A FUNCTION OF TEMPERATURE FOR THE REACTIONS:

$C + H_2O \rightleftharpoons CO + H_2$; $CO + H_2O \rightleftharpoons CO_2 + H_2$

$C + CO_2 \rightleftharpoons 2CO$ and $C + 2H_2 \rightleftharpoons CH_4$

PERFECT GAS LAW ASSUMED

An inspection of the data shows that only the CO_2/H_2O product ratio for reaction (2) is near equilibrium. The remainder of the experimental product ratios are one or two orders of magnitude away from the equilibrium. Discounting the production of methane and hydrogen by pyrolysis, and bearing in mind the fact that the presented equilibrium data do not consider the presence of oxygen in the system, makes it even more evident that the contribution of the above reactions, perhaps with the exception of reaction (2), to the production of combustible gases is negligible.

The extent to which reaction (2) occurs is difficult to assess from the available data since one of the products of the reaction, CO_2 is mainly produced from carbon and oxygen.

Since reactions (1) and (3) do not contribute significantly to the production of carbon monoxide, this gas must be produced mainly by reaction(5)

$$C + \tfrac{1}{2} O_2 \rightleftharpoons CO \qquad (5)$$

From Figure 17, at a constant steam ratio the production of carbon monoxide was a linear function of the coal feed rate, with a carbon monoxide yield of 52.6%. This implies that under this specific operating conditions, the ratio of carbon in carbon monoxide to carbon in the coal is:

$$\frac{52.6 \times 12}{28 \times 0.6679} = 33.8\%$$

If this ratio is calculated for all gasification runs with Forestburg coal (Table 26), the average values obtained for all particle sizes but the smaller ones are in the range of 30.7 to 36.5%. This suggests that

TABLE 26

CARBON IN CARBON MONOXIDE TO CARBON FEED RATIO FOR THE GASIFICATION OF FORESTBURG COALS

Run #	$\frac{\text{Carbon in CO}}{\text{Carbon in coal}}$ (w/w)
1	14.7
2	29.1
3	14.9
4	14.3
Mean ± S.D.	18.3 ± 7.2
5	29.7
6	48.2
7	42.6
8	26.5
9	10.7
Mean ± S.D.	31.5 ± 14.7
10	13.9
11	33.1
12	25.3
13	47.4
14	33.3
15	32.7
16	38.8
17	33.1
18	26.5
19	28.1
21	31.8
22	24.4
Mean ± S.D.	30.7 ± 8.2
23	23.3
24	46.5
25	39.9
26	47.8
27	39.5
28	32.7
29	29.9
30	27.5
31	41.4
Mean ± S.D.	36.5 ± 8.6
32	35.3
33	27.7
Mean ± S.D.	31.5 ± 5.4

carbon monoxide production is a linear function of the coal feed rate, quite independently of other operating variables, in particular the steam to coal ratio. Therefore, carbon monoxide appears to be produced mainly from reaction (5). This is in agreement with other results reported in the literature for a similar system$^{(36)}$.

Lastly, the preceding discussion on the relative importance of the different reactions occuring in a coal gasification system is supported by the thermodynamic and kinetic data of a coal gasification system presented in Chapter I.

COMPARISON OF GASIFICATION RESULTS FOR FLUIDIZED AND SPOUTED BED

One of the major objectives of this research was to compare the performance of a gasifier operated as a fluidized bed and as a spouted bed with the three different Western Canadian coals. This comparison is difficult because typical operating conditions for the two types of reactor were different. In some cases it was possible to operate under similar conditions, and in other cases it was not.

Bearing in mind the limitations noted above, a comparison of the performance of the fluidized bed reactor as obtained in this work with the experimental results of Foong et al. for a spouted bed (1) is presented in Table 27 for "typical" and "similar" operating conditions with the three different coals.

In general, the data show that for typical operating conditions the gas quality obtained from the spouted bed was 40 - 56% better than from the fluidized bed gasifier. On the other hand, at similar air to coal

TABLE 27

COMPARISON OF RESULTS FOR THE GASIFICATION OF SOME WESTERN CANADIAN COALS IN A FLUIDIZED AND SPOUTED BED

Type of result	Fluidized bed (this work)			Spouted bed (1)		
	Coal feed rate (g/s)	Air to coal ratio (w/w)	Gas gross calorific value (MJ/m^3)	Coal feed rate (g/s)	Air to coal ratio (w/w)	Gas gross calorific value (MJ/m^3)
Typical of 2.18 mm Forestburg coal	1.516	5.3	2.25	2.528	2.3	3.51
Forestburg coal under similar op.. conditions	1.673	5.01	2.00	1.211	5.1	1.35-1.79
Typical of 2.18 mm Sukunka coal	1.462	4.7	1.92	1.889	2.7	2.69
Sukunka coal under similar operating conditions	1.580	4.1	2.43	1.500	4.4	1.87
Coleman coal under similar operating conditions*	1.968	3.48	2.51*	1.750	2.90	1.79

* Under this condition coal caked in the fluidized bed

ratios, which in both cases was the most important operating parameter, the gas quality of the gas produced for a fluidized bed was 12 - 48% better than from the spouted bed. Although the air to coal ratio may have an industrial significance, the ratio does not necessarily mean the same thing from a process point of view for both types of reactor. In a fluidized bed, as long as there is not extensive bubbling the air to coal ratio can be representative of the effective oxygen to carbon ratio available for reaction. In a spouted bed on the contrary, the air to coal ratio is substantially higher for the spout region than for the annulus region, and both in turn are different from the overall air to coal ratio. This stretches the meaningfulness of comparing the performance of both reactors under "similar" operating conditions.

A basic difference between the operation of both reactors was the ability to obtain lower air to coal ratios and operate at higher coal feed rates in the spouted bed. As pointed out earlier (see Chapter 2) the coal reactivity limited the minimum air to coal ratio to approximately 4.0 for Forestburg coal and to ∿ 5.0 for Sukunka coal. Since on the other hand the maximum air flow was limited by the size of the equipment and the practicality of operation under fluidizing mode to approximately 8 x 10^{-3} std m^3/s; the maximum coal feed rate was limited to 2.00 - 2.40 g/s while typical feed rates were 1.2 - 2.0 g/s. By contrast, typical feed rates for the gasification at Forestburg coal in a spouted bed were 2.50 - 3.33 g/s i.e. substantially higher. The preceding facts indicate that for a given equipment size, larger coal throughputs can be achieved with a spouted bed (Table 28). This allows the production of a gas of better quality.

TABLE 28

COMPARISON OF THE MAIN OBSERVED OPERATING CHARACTERISTICS OF A FLUIDIZED AND SPOUTED BED REACTORS

Operating variable	Fluidized bed	Spouted bed
Typical throughput (g/s m^2) for non-caking coal (Forestburg)	65.8 -109.7	137.1-182.6
Effect of coal feed rate on gas quality	critical	critical
Effect of steam ratio on gas quality	little	none
Effect of bed depth on gas quality	none	none
Observed maximum throughput (g/s m^2) for Sukunka (caking) coal	86.6	124.9
Observed maximum throughput (g/s m^2) for Coleman (caking) coal	107.9 (a)	95.9

(a) Under this condition coal severely caked in the bed.

In spite of these disimilarities, the spouted and fluidized bed have several common characteristics (Table 28). In both systems, the coal feed rate and air to coal ratio were critical in determining the gas quality while their performances were largely insensitive (other than gas composition for the fluidized bed) to the steam to coal ratio. This is no doubt a characteristic of a system operating at relatively low temperature where pyrolysis and partial combustion of carbon are the preponderant reactions in determining the gas quality. Both systems were also insensitive to changes in bed depth, probably for the same reasons mentioned above. Finally, the maximum throughput of caking coals for both systems were similar. The maximum throughput of Sukunka coal in the spouted bed was 124.9 (g/s m^2), at which point caking problems arose, while a throughput of 86.6 (g/s m^2) of this coal was processed without any problems by the fluidized bed. Higher throughputs could not be achieved in the latter system because of coal reactivity and equipment limitations rather than caking. A comparison of fluid bed versus spouted bed performance in a larger refractory lined reactor would be valuable.

The maximum throughput with Coleman coal for the spouted bed was 95.9 (g/s m^2). In the fluidized bed Coleman coal could be gasified at a throughput of 107.9 (g/s m^2) for approximately 2 hours after which time the coal was found to have caked. Gasification of this coal at lower throughputs was not attempted. Overall, it appears that the ability of gasifying caking coals in a spouted bed is mainly due to a dilution effect of the inert silica particles rather than because of the potential of the high velocity spout to break up agglomerates.

CHAPTER VI - CONCLUSIONS

Three different Western Canadian coals, one non-caking and two of the caking type of coal, were gasified under a wide range of operating conditions in a small pilot plant capable of steady operation at throughputs of up to 109.6 g/s m^2 (2.0 g/s) of dry non-caking coal. The gasifier was an air fluidized bed of 0.73 mm Ottawa sand and coal operated at atmospheric pressure. Discounting the dilution effect of some nitrogen used in the coal feeding system, all three coal tested typically produced a gas of a calorific value in the range of 2.0 - 2.6 MJ/m^3. These values are lower than for commercially available fluidized bed reactors which produce gas of a calorific value in the range of 3.91 - 4.40 MJ/m^3. The difference in performance is attributed to the inability to operate at average temperatures over about 1175 K because the reactor was not refractory lined, and to the important impact that large energy losses from the reactor (about 25% of energy input) had on the quality of the gas produced.

Analysis of the effect of the different operating variables indicated that the single most important operating parameter in determining the gas quality was the coal feed rate, while an inverse correlation between the air to coal ratio and gas quality was generally observed. Minimum air to coal ratios were limited by coal reactivity to 4.0 and about 5.0 for Forestburg and Sukunka coal respectively. Temperature in the range of 1023 - 1175 K, particle size, bed depth and steam to coal ratio did not have important effects on the gas calorific value. However, increasing the steam to coal ratio resulted in an increase of the hydrogen concentration and a decrease of the carbon monoxide concentration.

Even though the coal particle size had only minor effects on the quality of the gas produced it had an impact on the quality of the operation. Increasing the coal particle size made the operation more stable, decreased the amount of tar produced and decreased the amounts of solids and carbon elutriated from the bed thereby increasing the raw gas thermal efficiency and the useful carbon ratio. Gasification of coal with a wider size distribution which included the fines (-4.76 + 0.0 mm), as opposed to operation with narrowly sized coal, did not have a detrimental effect either on the gas quality or the stability of the operation although an increase in solids and carbon elutriation rates was observed. Tar production from this coal was less than expected from its nominal average size.

The coal characteristics also had an impact on the quality of the operation. The non-caking coal (Forestburg) and one caking coal of free swelling index of 7 (Sukunka) could be gasified without any operating problems at throughputs of 109.6 g/s m^2 and 86.62 g/s m^2 respectively. A second caking coal, with a free swelling index of 4 (Coleman) caked when gasified at a throughput of 107.9 g/s m^2. Solids and carbon elutriation rates for the gasification of Sukunka coal were higher by a factor of 3 than for Forestburg coal, while for the one run with Coleman coal they were intermediate between the two. Tar production from these three coals were in similar proportions to the elutriation rates.

Analysis of the data indicated that in the present system combustion and pyrolysis reaction prevailed over gasification reactions. Most of the hydrogen and essentially all of the methane produced arose from coal thermal decomposition while most of the carbon monoxide appeared to be produced from partial combustion of carbon. In most gasification runs, water was produced in the reactor from combustion of hydrogen and hydrocarbons. Carbon dioxide appeared to be mostly produced by the combustion of carbon, although there was experimental evidence of the reaction of carbon monoxide with water to form carbon dioxide and hydrogen. The main gasification reaction, that of carbon and steam to produce carbon monoxide and hydrogen did not appear to occur to any important extent. These conclusions were in agreement with results reported in the literature and thermodynamic and kinetic predictions for a gasification system operated at temperatures below 1175 K which allowed for relatively large energy losses from the reactor.

Comparison of the performance of the fluidized bed and a spouted bed for gasification of the three coals was difficult because it was not always possible to operate under similar conditions. A basic difference between the two gasifiers was the ability to operate at higher coal feed rates and lower air to coal ratio with the spouted bed. Since in both cases these operating parameters were the most important factors in determining the gas quality, the data showed that for "typical" operating conditions the gas quality obtained from the spouted bed was 40 - 56% better than for the fluidized bed gasifier. On the other hand, at "similar" air to coal ratios, the gas quality of the gas produced in the latter reactor was 12 - 48% better than from the spouted bed. This indicated that for a given equipment size, larger coal throughputs can probably be achieved with a spouted bed, allowing the production of a gas of better quality.

In spite of this dissimilarity, both gasifiers had several common characteristics. In both systems the coal feed rate and air to coal ratio were critical in determining the gas calorific value, while their performance was largely insensitive to the steam to coal ratio, temperature, and bed depth. This is no doubt a characteristic of a system operating at relatively low temperature where combustion and pyrolysis reactions dominate.

The ability of both systems to process caking coal was also similar. The maximum throughput of Sukunka coal in the spouted bed was 124.9 g/s m^2 while 86.62 g/s m^2 of this coal was processed without any problems in the fluidized bed. Larger throughputs could not be achieved in the latter because of coal reactivity and equipment limitations rather than caking. The maximum throughput of Coleman coal in the spouted bed was 95.9 g/s m^2 while in the fluidized bed this coal caked after 2 hours of operation at a throughput of 107.9 g/s m^2. In conclusion, it appears that the previously noted ability of a spouted bed to treat limited throughputs of caking coals is due to the dispersion of the coal in a bed of inert silica and ash rather than to the ability of the high velocity gas spout to break-up agglomerates.

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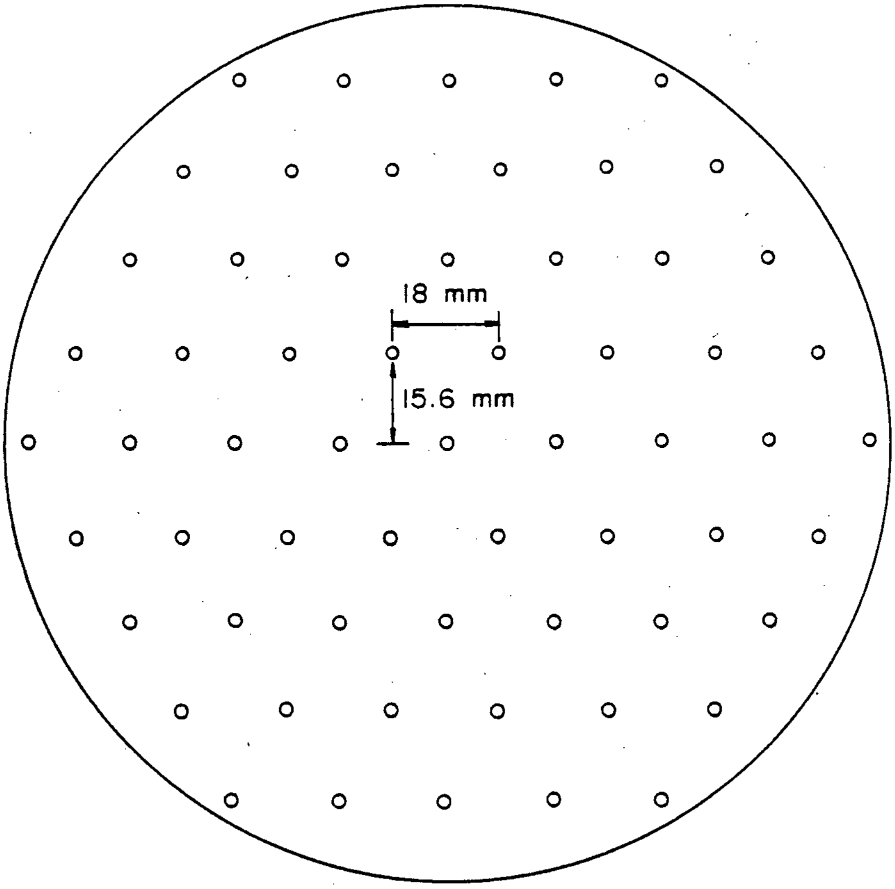
APPENDIX I

DETAILS OF FLUIDIZATION GRID



DETAIL OF PERFORATION :

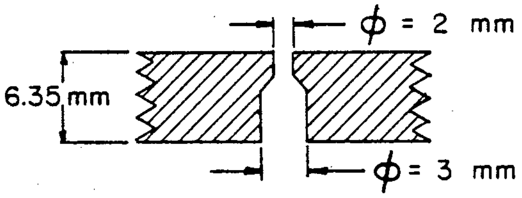

ϕ = 2 mm
6.35 mm
ϕ = 3 mm

FIGURE I-1: DETAIL OF FLUIDIZATION GRID

APPENDIX II

TOTAL GAS FLOW MEASUREMENT BY ORIFICE PLATE

TOTAL GAS FLOW MEASUREMENT BY ORIFICE PLATE

The total gas flow out of the gasifier was measured with a 19.1 mm (3/4") orifice plate of our own design (Figures II-1 and II-2) at a point between the cyclone and the gas incinerator. Pressure and temperature upstreame of the orifice plate were measured by a mercury manometer and a calibrated chromel-alumel thermocouple respectively.

The orifice plate was installed and then calibrated with air by using a Straushibe (S-type) pilot tube.

PITOT TUBE:

The pitot tube was supplied by B.C. Research and calibrated in the wind tunnel of the Department of Mechanical Engineering of U.B.C.

Air velocities in the wind tunnel were obtained by measuring the pressure drop in the tunnel with a Betz manometer. Since

$$P = \tfrac{1}{2} \rho V^2 = \rho g h$$

then $V = \sqrt{2gh}$

where V = air velocity (m/s)

g = acceleration due to gravity = 9.80665 m/s^2

H = pressure differential in m of flowing gas

h is given by: $h = h_1 \sqrt{\dfrac{\rho_{H_2O}}{\rho_{air}}}$

where: h_1 = Betz manometer pressure drop, mm H_2O

$\rho_{H_2O} = 1 \times 10^6$ g/m^3

$\rho_{air} = \dfrac{M.W. \times P}{RT}$; under calibration conditions :

$$\rho_{air} = \frac{28.91 \times 101.592 \times 10^3}{8.31439 \times 296.3} = 1.1922 \times 10^2 \text{ g/m}^3$$

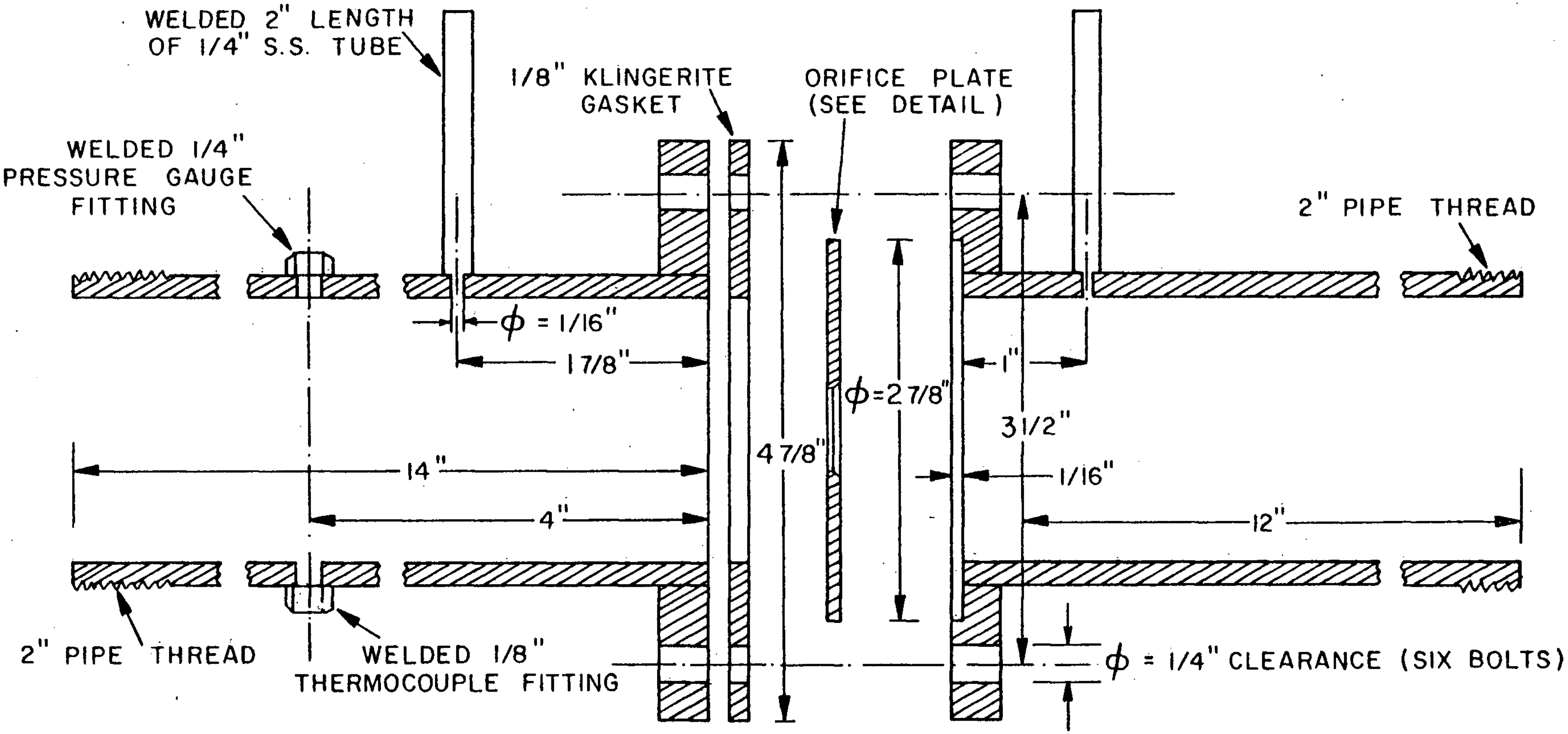

WELDED 2" LENGTH OF 1/4" S.S. TUBE
1/8" KLINGERITE GASKET
ORIFICE PLATE (SEE DETAIL)
WELDED 1/4" PRESSURE GAUGE FITTING
2" PIPE THREAD
ϕ = 1/16"
1 7/8"
1"
ϕ=2 7/8"
3 1/2"
4 7/8"
14"
4"
1/16"
12"
2" PIPE THREAD
WELDED 1/8" THERMOCOUPLE FITTING
ϕ = 1/4" CLEARANCE (SIX BOLTS)

FIGURE II-1: ORIFICE PLATE CONSTRUCTION DETAIL

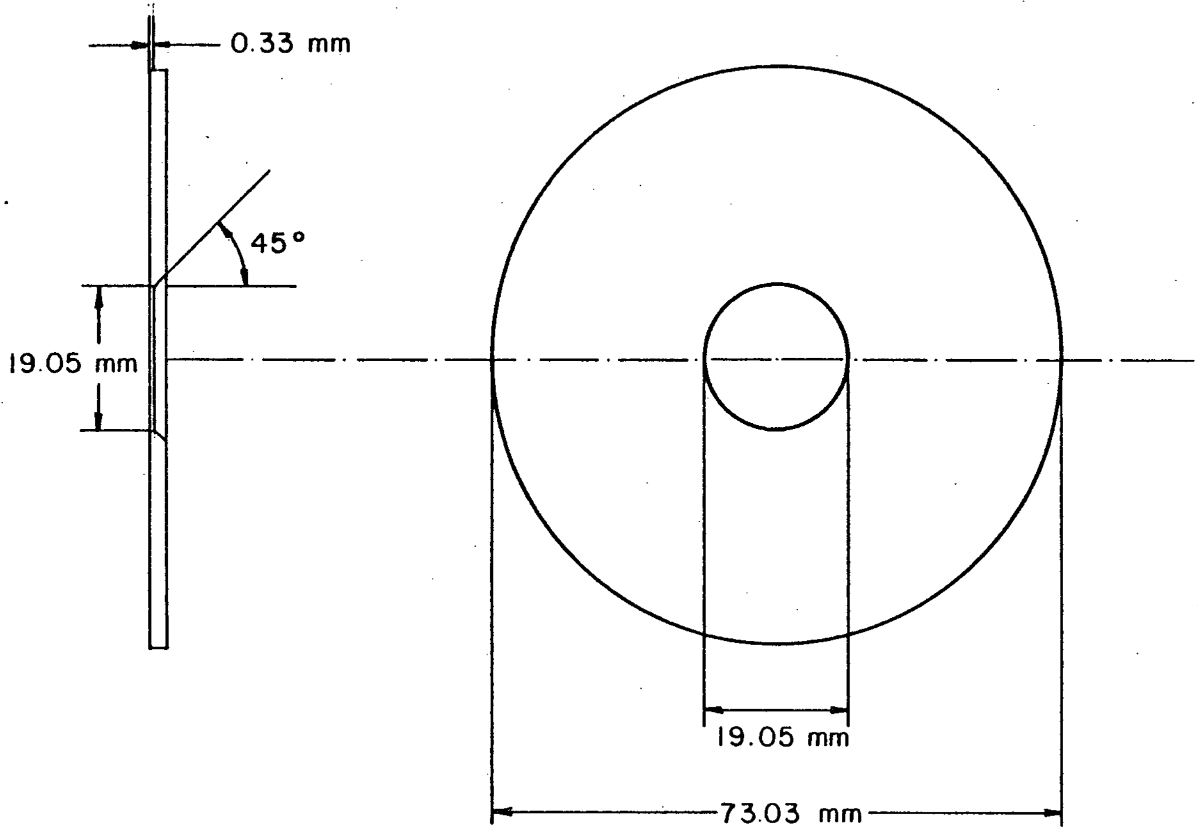

0.33 mm
45°
19.05 mm
19.05 mm
73.03 mm

FIGURE II-2: ORIFICE PLATE DETAIL

Then:

$$V = \sqrt{\frac{2 \times 9.80665 \times h_1 \times 10^{-3} \times 1.0 \times 10^{6}}{1.1922 \times 10^{3}}}$$

$$V = 4.056 \sqrt{h_1} \quad (\mathrm{m/s}) \qquad (1)$$

Similarly, from Bernouilli's equation the velocity measured by the pitot tube is related with the differential pressure ΔP by

$$V = C \sqrt{2 g \Delta P}$$

where C = discharge coefficient of pitot tube under the calibration conditions

$$V = 4.067\, C \sqrt{\Delta P} \qquad \text{or}$$

$$C = \frac{V}{4.056 \sqrt{\Delta P}} \qquad (2)$$

The discharge coefficient of the pitot tube was calculated for each calibration condition (Table II-1). The average discharge coefficient was found to be :

$$C = 0.839 \pm 0.006$$

Therefore, the calibration equation for the pitot tube is:

$$V = C \quad 2 g \sqrt{\frac{\Delta P}{\rho}} = C \sqrt{2g R} \sqrt{\frac{\Delta P \times T}{P \times M.W.}}$$

$$V = 338.81 \sqrt{\frac{\Delta P \times T}{P \times M.W}} \quad (\mathrm{m/s}) \qquad (3)$$

or, for air:

$$V = 63.013 \sqrt{\frac{\Delta P \times T}{P}} \quad (\mathrm{m/s}) \qquad (4)$$

Where: ΔP = Pitot tube pressure drop, mm H_2O

T = upstream temperature K

P = upstream pressure, Pa

The calibration curve of the pitot tube for air is presented in Figure II-3

TABLE II-1

PITOT TUBE CALIBRATION DATA

ΔP Pitot Tube mm H_2O	Betz Monometer Pressure Differential (mm H_2O)	Velocity (Eq. (1)) (m/s)	Discharge Coefficient C (Eq. (2))
3.30	2.30	6.15	0.835
5.598	3.85	7.96	0.830
7.87	5.50	9.51	0.836
10.41	7.30	10.96	0.838
13.46	9.50	12.50	0.840
19.81	14.14	15.25	0.845
22.86	16.35	16.40	0.846

Air upstream pressure = 101.592 kPa (30 in Hg)

Air upstream temperature = 296.3 K

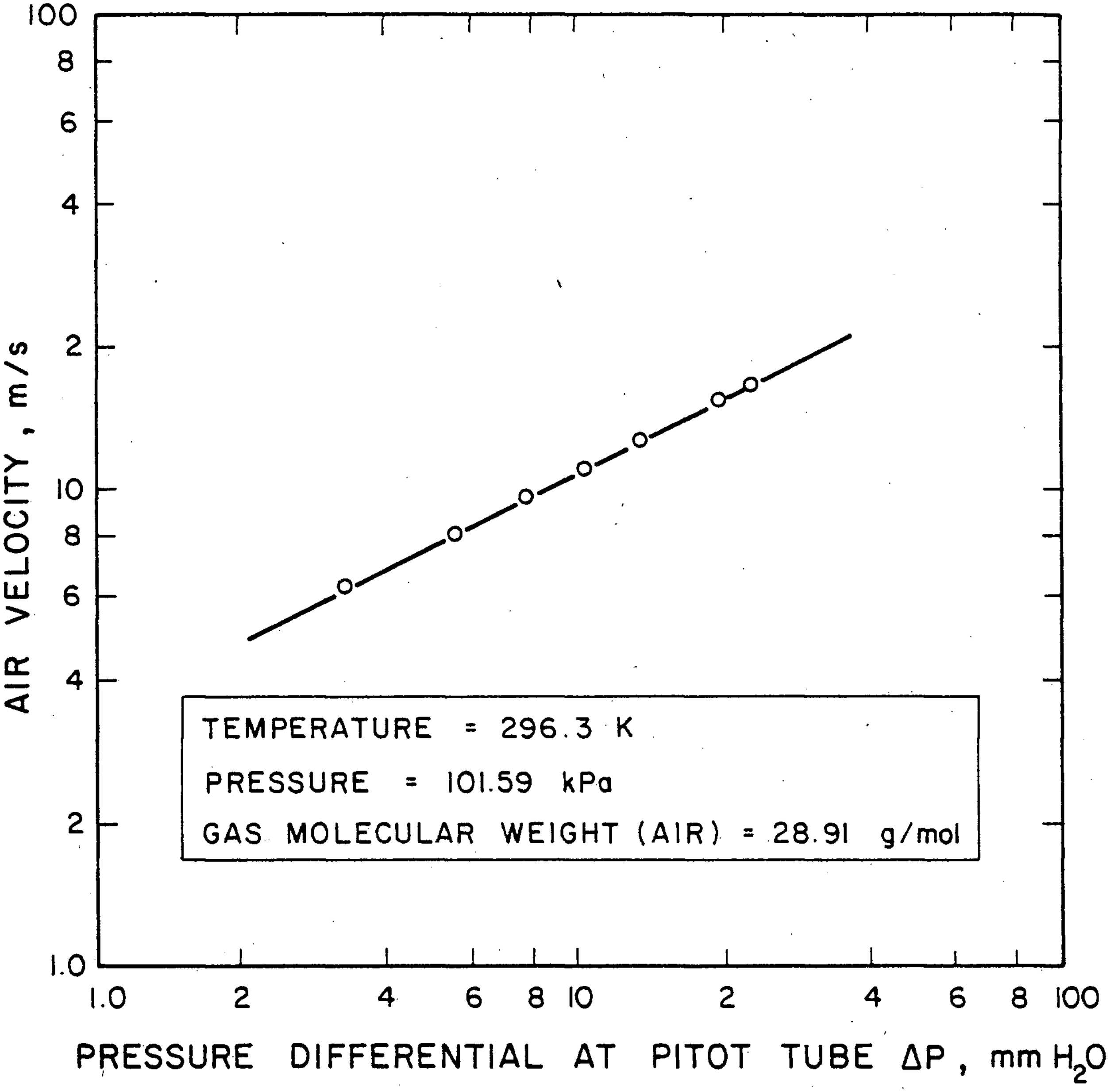


FIGURE II-3: CALIBRATION CURVE FOR THE PITOT TUBE

ORIFICE PLATE CALIBRATION

The procedure to calibrate the orifice plate was as follows. For different air flows the pressure drop across the orifice plate as well as the upstream pressure and temperature were recorded. Downstream (approximately 1 m) of the orifice plate, the air velocity at 3 different points of a cross section of the air pipe (Table II-2) were measured using the previously calibrated pitot tube. These three points were chosen according to the principle of proportional areas; i.e. each point represented the middle point over a pipe diameter of three sections of the pipe having the same area. The air flow was then calculated from the average air velocity and pipe cross sectional area (21.646 x 10^{-4} m^2). Since the flow through an orifice plate is given by:(31)

$$Q = \frac{A_2 C \sqrt{2\ gh}}{\sqrt{1 - \left(\frac{A_2}{A_1}\right)^2}} = 403.823 \quad A_2\ C \sqrt{\frac{\Delta\ P \times T}{M.W. \times P \left[1 - \left(\frac{A_2}{A_1}\right)^2\right]}}$$

where:

A_2 = orifice area = 2.852 x 10^{-4} m^2

A_1 = pipe cross area section = 21.646 x 10^{-4} m^2

M.W. = Molecular weight of gas = 28.91 g/mol for air

T = upstream temperature = 290.36 K

Δ P = pressure drop across orifice plate, mm H_2O

P = upstream pressure, Pa

C = discharge coefficient, dimentionless

Then

$$Q = 0.368C\sqrt{\frac{\Delta\ P}{P}}$$

from which the orifice plate discharge coeficient can be calculated by:

$$C = \frac{Q}{3.368}\sqrt{\frac{P}{\Delta\ P}}$$

TABLE II-2

ORIFICE PLATE CALIBRATION DATA:

ORIFICE PLATE			PITOT TUBE							Flow
ΔP (mm H_2O)	Temperature (^oC)	Relative Pressure (kPa)	inner point		middle point		outer point		Average	(m^3/s)
			ΔP (mm H_2O)	Velocity (m/s)	ΔP (mm H_2O)	Velocity (m/s)	ΔP (mm H_2O)	Velocity (m/s)	Velocity (m/s)	
39.37	17.4	0	0.381	2.07	0.508	2.39	0.584	2.57	2.34	5.07 x 10^{-3}
48.26	17.3	0	0.508	2.39	0.635	2.68	0.711	2.83	2.63	5.69
59.69	17.2	0	0.686	2.78	0.813	3.03	0.813	3.03	2.95	6.39
76.20	17.2	0.32	0.889	3.17	1.016	3.38	1.067	3.47	3.34	7.23
85.09	17.2	0.37	1.016	3.38	1.016	3.38	1.219	3.71	3.49	7.55
95.25	17.4	0.50	1.219	3.71	1.321	3.86	1.321	3.86	3.81	8.25
107.95	17.3	0.62	1.270	3.78	1.473	4.08	1.499	4.11	3.99	8.64
118.4	17.3	0.75	1.524	4.14	1.575	4.21	1.575	4.21	4.19	9.07
127.00	17.4	0.87	1.575	4.21	1.727	4.41	1.778	4.48	4.37	9.46
147.32	17.4	0.97	1.778	4.48	2.032	4.79	2.032	4.79	4.69	10.15
177.80	17.3	1.32	2.159	4.93	2.337	5.13	2.413	5.22	5.09	11.02
201.93	17.4	1.49	2.540	5.35	2.667	5.48	2.667	5.48	5.44	11.78
231.14	17.4	1.74	2.743	5.56	3.048	5.86	3.048	5.86	5.76	12.47
254.00	17.4	1.99	3.048	5.86	3.302	6.10	3.302	6.10	6.02	13.03
280.67	17.4	2.24	3.302	6.10	3.683	6.44	3.683	6.44	6.33	13.70
300.99	17.4	2.37	3.683	6.44	3.937	6.66	3.937	6.66	6.59	14.26
328.93	17.4	2.61	3.937	6.66	4.267	6.94	4.318	6.98	6.86	14.85
360.68	17.4	2.94	4.369	7.02	4.699	7.28	4.699	7.28	7.19	15.56
388.62	17.4	3.19	4.699	7.28	5.080	7.57	5.207	7.66	7.50	16.23
415.29	17.4	3.44	4.953	7.47	5.334	7.75	5.461	7.85	7.69	16.65
454.66	17.4	3.74	5.334	7.75	5.842	8.12	6.096	8.29	8.05	17.43
467.36	17.5	3.91	5.461	7.85	5.842	8.12	6.096	8.29	8.09	17.51
495.30	17.5	4.23	5.842	8.12	6.350	8.46	6.350	8.46	8.35	18.07 x 10^{-}

Atmospheric pressure: 102.269 kPa; Average temperature: 290.36 K

The average discharge coeficient was (Table II-3):

$$C = 0.722 \pm 0.007$$

Therefore, the following equation was used to measure total gas flow out of the reactor:

$$Q = 403.823 \quad A_2\ C \sqrt{\frac{\Delta\ P \times T}{M.W. \times P \quad 1 - \left(\frac{A_2}{A_1}\right)^2}}$$

$$Q = 0.084 \sqrt{\frac{\Delta\ P \times T}{M.W. \times P}} \qquad (m^3/s)$$

where Δ P, T, P were measured and M.W. is known once the gas composition is known.

TABLE II-3

DETERMINATION OF ORIFICE PLATE DISCHARGE COEFFICIENT

ΔP ($mm\ H_2O$)	Absolute Pressure (kPa)	Flow (m^3/s)	Discharge Coefficient C
39.37	102.27	5.07×10^{-3}	0.702
48.26	102.27	5.69	0.712
59.69	102.27	6.39	0.719
76.20	102.59	7.23	0.721
85.09	102.64	7.55	0.713
95.25	102.37	8.25	0.736
107.95	102.89	8.64	0.725
118.11	103.02	9.07	0.728
127.00	103.14	9.46	0.733
147.32	103.24	10.15	0.730
177.80	103.59	11.02	0.723
201.93	103.76	11.78	0.726
231.14	104.01	12.47	0.719
254.00	104.26	13.03	0.717
280.69	104.51	13.70	0.718
300.99	104.64	14.26	0.723
328.93	104.88	14.85	0.721
360.68	105.21	15.56	0.722
388.62	105.46	16.23	0.727
415.29	105.71	16.65	0.722
454.66	106.01	17.43	0.723
467.36	106.18	17.51	0.717
495.30	106.50	18.07×10^{-3}	0.720

APPENDIX III

IMPINGER SYSTEM FOR GAS SAMPLE CLEANING

IMPINGER SYSTEM FOR GAS SAMPLE CLEANING

The basic gas sample cleaning unit was an impinger as shown in Figure III-1. This was basically an hermetic (except for gas inlet and outlet) stainless steel scrubber filled with approximately 150 cm^3 of water. The impinger train consisted of two parallel sets of four impingers in series. In each set, the second impinger had a perforated diffuser section as shown in Figure III-1, while in the remaining three impingers the diffuser section was simply an open end tube. The two sets of impingers were fitted with a system of valves which allowed operation of one set while the other was by-passed. This permitted continuous operation of one set for gas cleaning while the second set was only connected when measuring tar content of the gas. The whole system was kept cool by immersion in a bath of cracked ice.

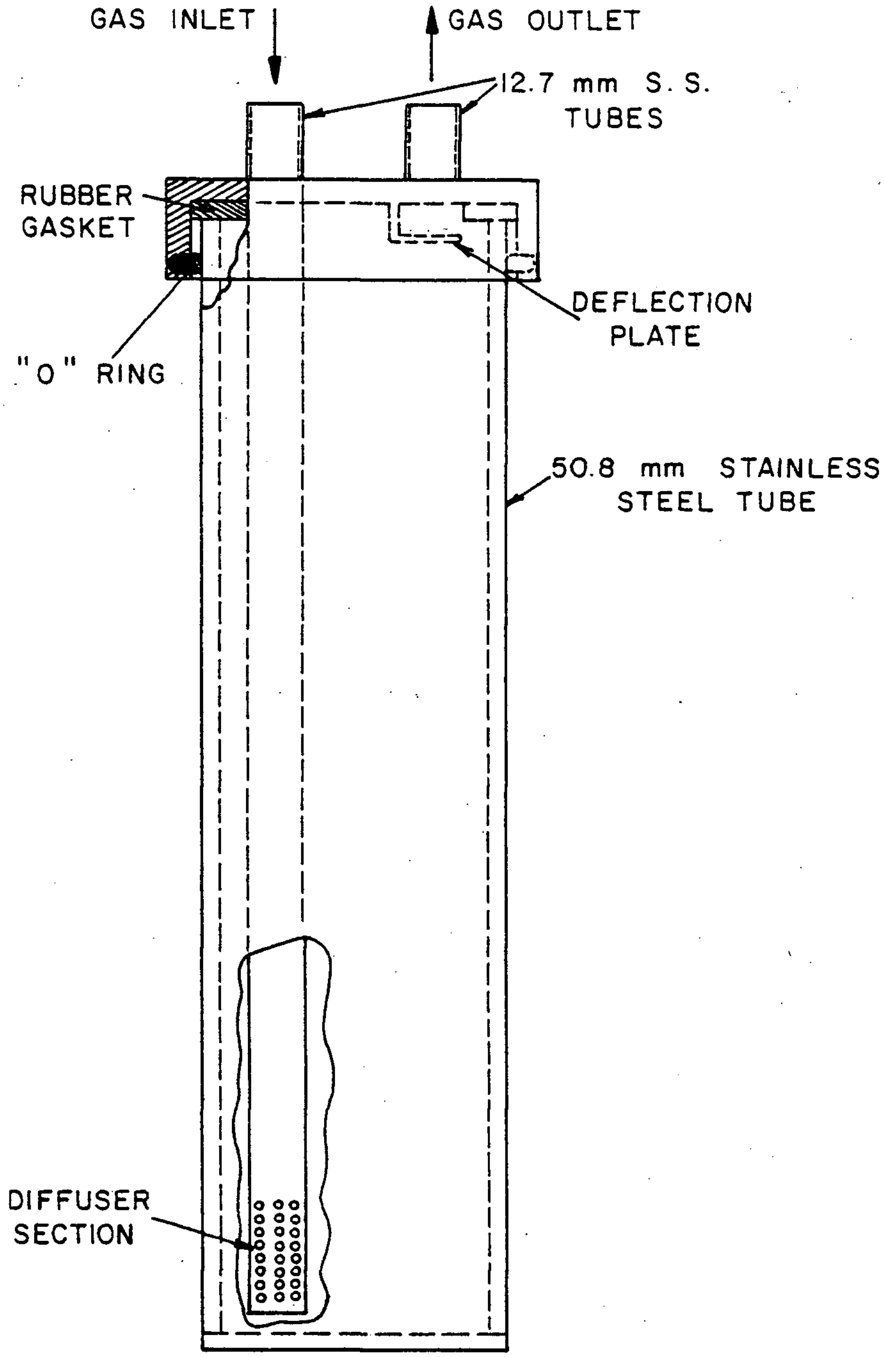

GAS INLET
GAS OUTLET
12.7 mm S. S. TUBES
RUBBER GASKET
DEFLECTION PLATE
"O" RING
50.8 mm STAINLESS STEEL TUBE
DIFFUSER SECTION

FIGURE III-1: IMPINGER DETAIL

APPENDIX IV

CALIBRATION OF ROTAMETERS

CALIBRATION OF ROTAMETERS

SCRUBBING NITROGEN ROTAMETER

This rotameter was factory calibrated in a direct reading scale.

GASIFIER AIR ROTAMETER

The gasifier rotameter was calibrated by using a dry gas meter (CHE 2856). This gas meter was in turn checked against a high precision wet test meter made available by the Water Resources Laboratory in Vancouver. The agreement between both gas meters was excellent, the U.B.C. gas meter reading (x) being related to the high precision meter reading (y) by the equation:

$$y = 0.98\ x + 0.1 \qquad (1)$$

with a correlation factor of 0.98 at standard conditions (101.3 Pa = 1 Atm. and 294 K). Just before the entrance to the rotameter, the air pressure and temperature were measured with a 170.3 kPa (10 psig). Matheson pressure gauge and a calibrated chromel-alumel thermocouple respectively. The data taken during the calibration of the fluidization air rotameter against the UBC gas meter are shown in Table IV-1. The gas meter readings (x) on the last column of Table IV-1 were then corrected according to Eq. (1) and are shown in Table IV-2. Finally the calibration curve (Figure IV-1) was drawn at standard conditions. This required calculating the air flow under standard conditions that will give the same rotameter reading as the flow under the actual measuring conditions. This is given by the following equation:(32)

$$f_1 = \sqrt{\frac{\rho_1}{\rho_2}} = \sqrt{\frac{P_1}{P_2} \times \frac{T_2}{T_1}} \qquad (2)$$

where F = volumetric flow

ρ = fluid density

P = absolute pressure

T = absolute temperature

TABLE IV-1

GASIFICATION AIR ROTAMETER CALIBRATION DATA

ROTAMETER			UBC GAS-METER			
Reading	Relative Pressure kPa	Temperature °C	Relative Pressure kPa	Temperature °C	(a) Flow (m3/s)	(b) Flow (m^3/s)
20	-	15.5	0.20	21.0	0.94x10^{-3}	0.94x10^{-3}
40	-	14.9	0.22	20.0	1.57	1.58
60	-	15.1	0.22	20.0	2.27	2.27
80	-	15.2	0.25	20.5	2.87	2.88
100	-	15.5	0.25	21.0	3.57	3.58
120	-	15.9	0.25	21.0	4.27	4.28
140	-	16.1	0.25	21.0	4.93	4.95
160	-	16.0	0.25	21.0	5.66	5.68
180	-	16.0	0.25	21.0	6.36	6.38
200	0.55	16.1	0.27	21.0	7.08	7.10
210	1.10	16.2	0.27	21.0	7.50	7.52
220	1.38	16.3	0.27	21.0	7.84	7.86
230	2.00	16.3	0.27	21.0	8.27	8.28
240	2.28	16.3	0.27	21.0	8.61	8.63
250	2.62	16.2	0.27	21.0	9.02x10^{-3}	9.04x10^{-3}

(a) At gas-meter pressure and temperature

(b) At standard conditions (101.3 kPa = 1 At; 294 K)

TABLE IV-2

CALIBRATION DATA OF GASIFICATION AIR ROTAMETER AT STANDARD CONDITIONS:

Rotameter Reading	Corrected gas meter Flow (a) (m^3/s)	Rotameter Flow (b) (m^3/s)
20	1.00×10^{-3}	0.99×10^{-3}
40	1.61	1.59
60	2.28	2.26
80	2.86	2.83
100	3.54	3.51
120	4.22	4.18
140	4.87	4.83
160	5.57	5.52
180	6.24	6.19
200	6.90	6.82
210	7.27	7.17
220	7.57	7.46
230	7.93	7.79
240	8.24	8.08
250	8.59×10^{-3}	8.41×10^{-3}

(a) at rotameter temperature and pressure (see table IV-1)

(b) at standard conditions in the rotameter

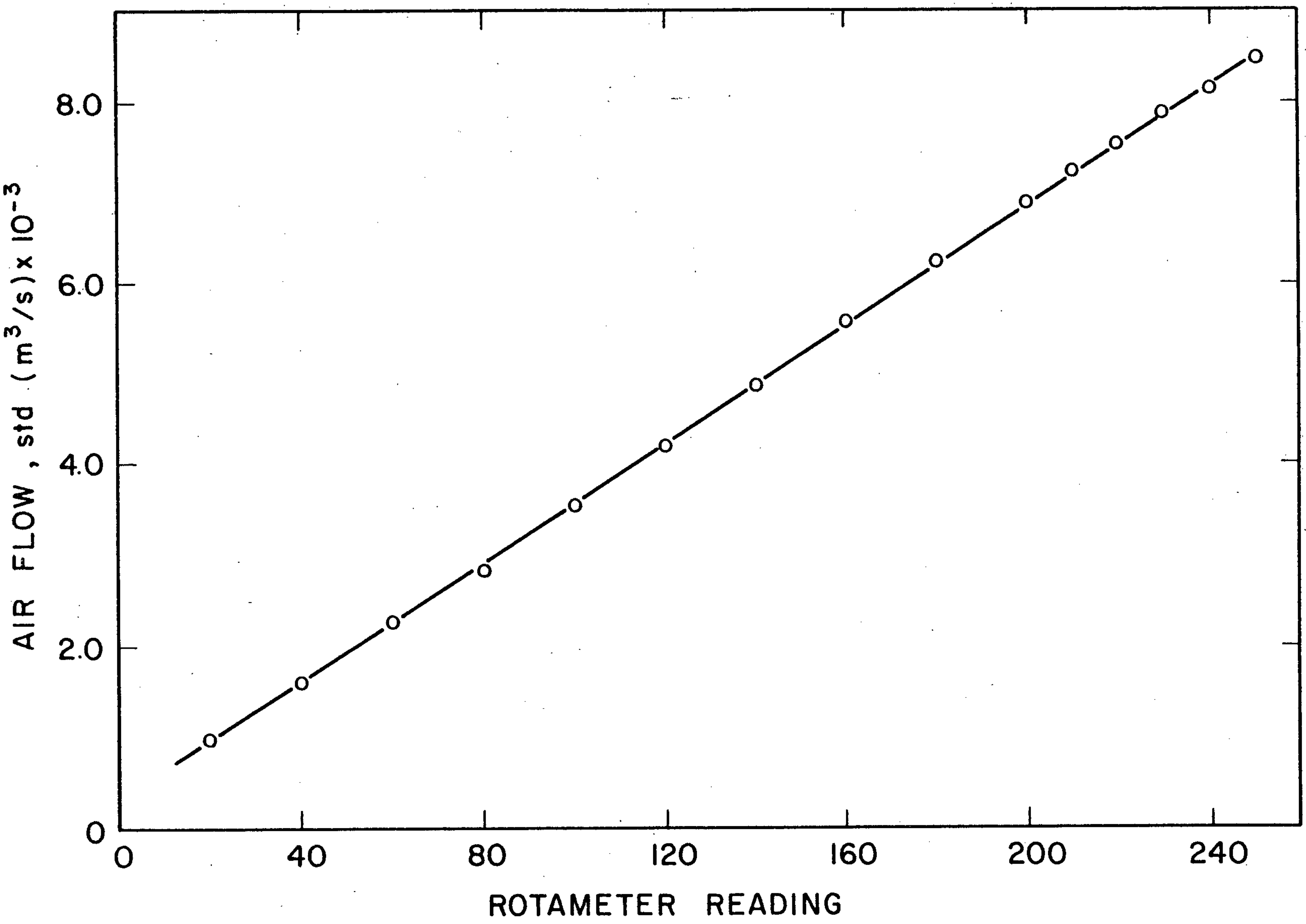


FIGURE IV-1: CALIBRATION CURVE OF GASIFICATION AIR ROTAMETER

In this case subscripts 1 and 2 denote standard conditions and rotameter actual conditions respectively. Results of these calculations are shown in Table IV-2:

Since in most of the gasification experiments the air flow into the gasifier was not metered under standard conditions, the reverse procedure as indicated here was followed to calculate actual air flow from the calibration curve. In other words, the air flow derived from the calibration curve (Figure IV-1) was corrected by equation (2).

INCINERATOR AIR ROTAMETER

This rotameter was calibrated using the same gas meter used for calibration of the gasifier air rotameter.

Since it is not necessary to meter the air into the electric furnace with extreme accuracy, and the air is metered at conditions very close to standard, no corrections allowing for temperature and pressure difference from standard conditions were made. The incinerator air rotameter calibration data is presented in Table IV-3 and calibration curve in Figure IV-2.

STEAM METER

Due to the small steam flows involved, the steam fed into the reactor was measured by a rotameter (Table 9). The rotameter was calibrated by completely condensing the steam after passing it through the rotameter. The condensed water was collected over a period of 10 min. for each rotameter reading and its volume measured in a graduated cylinder. Calibration data and curve are shown in Table IV-4 and Figure IV-3 respectively.

TABLE IV-3

INCINERATOR AIR ROTAMETER - CALIBRATION DATA

Rotameter Reading	Gas Meter			
	Relative Pressure kPa	Temperature °C	Flow (a) (m^3/s)	Flow (b) (m^3/s)
20	0.12	25	1.10×10^{-3}	1.11×10^{-3}
40	0.12	24	2.02	2.03
60	0.17	24	2.90	2.87
80	0.20	24	3.76	3.70
100	0.25	24	4.72	4.64
120	0.25	24	5.63	5.52
140	0.25	24	6.67	6.53
160	0.25	24	7.66	7.50
180	1.25	24	8.69	8.59
200	1.49	24	9.71	9.61
220	1.67	24	10.85	10.75
240	2.17	24	12.07×10^{-3}	12.00×10^{-3}

(a) At Gas meter temperature and pressure.

(b) At standard conditions (101.3 kPa, 294K) and corrected for gas meter deviation. (Eq. (1))

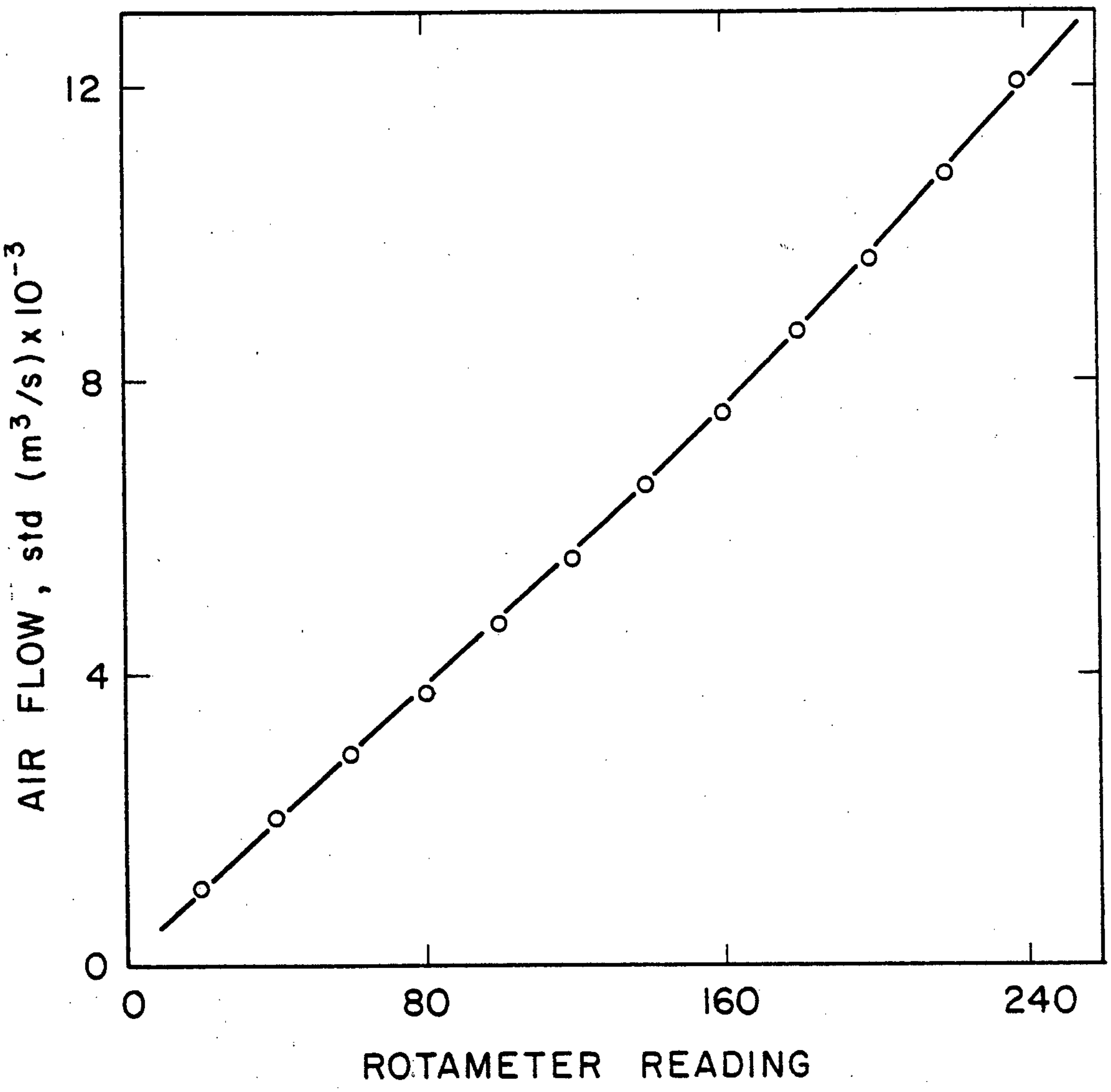


FIGURE IV-2: CALIBRATION CURVE OF INCINERATOR AIR ROTAMETER

TABLE IV-4

STEAM METER CALIBRATION DATA

Rotameter Reading	Mass of water condensed in 10 min. (g)	Flow (g/s)
0.5	243	0.41
0.75	325	0.54
1.00	360	0.60
1.25	460	0.77
1.50	660	1.10
1.75	750	1.25
2.00	880	1.47
2.25	1006	1.68
2.50	1136	1.89

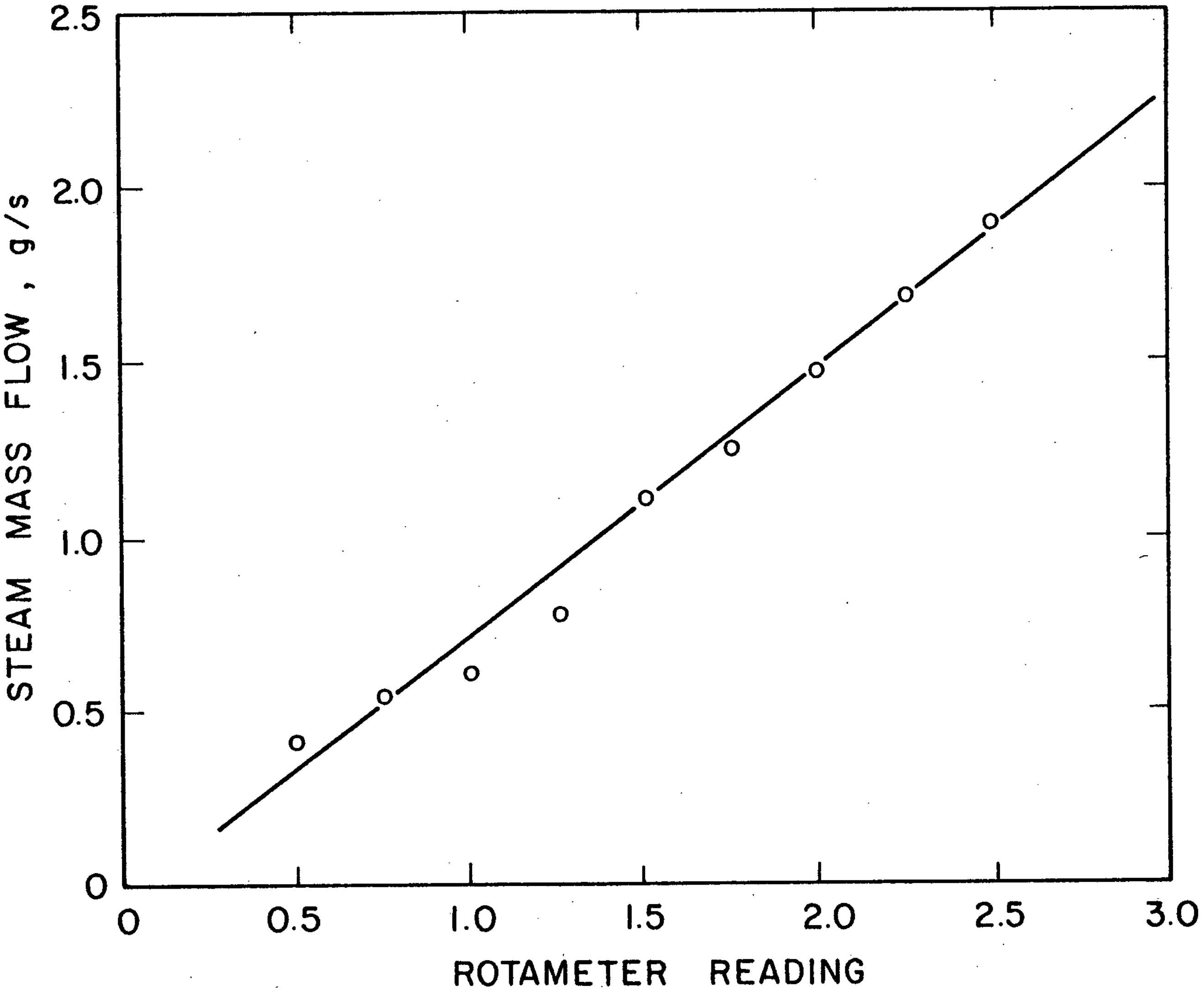


FIGURE IV-3: STEAM METER CALIBRATION CURVE

GAS SAMPLE ROTAMETER

The continuous gas sample withdrawn from the main gas line was metered with a small rotameter after it had been cleaned of solids and tars, cooled and dried (Figure 3). The rotameter was calibrated with air using a small wet test meter (0.118 x 10^{-3} (m^3/s) per revolution). The calibration data is shown in Table IV-5 and the calibration curve in Figure IV-4.

TABLE IV-5

GAS SAMPLE ROTAMETER - CALIBRATION DATA:

Rotameter Reading	GAS METER			
	Temperature °C	Absolute Pressure kPa	Flow (a) (mm^3/s)	Flow (b) (mm^3/s)
17	22.8	101.32	27.38×10^3	27.21×10^3
25	22.8	101.33	46.26	45.98
32	23.1	101.33	56.64	56.24
47	23.1	101.35	90.62	90.00
55	23.1	101.35	112.34	111.57
69	23.1	101.37	151.04×10^3	150.03×10^3

(a) At gas meter conditions.

(b) At standard conditions.

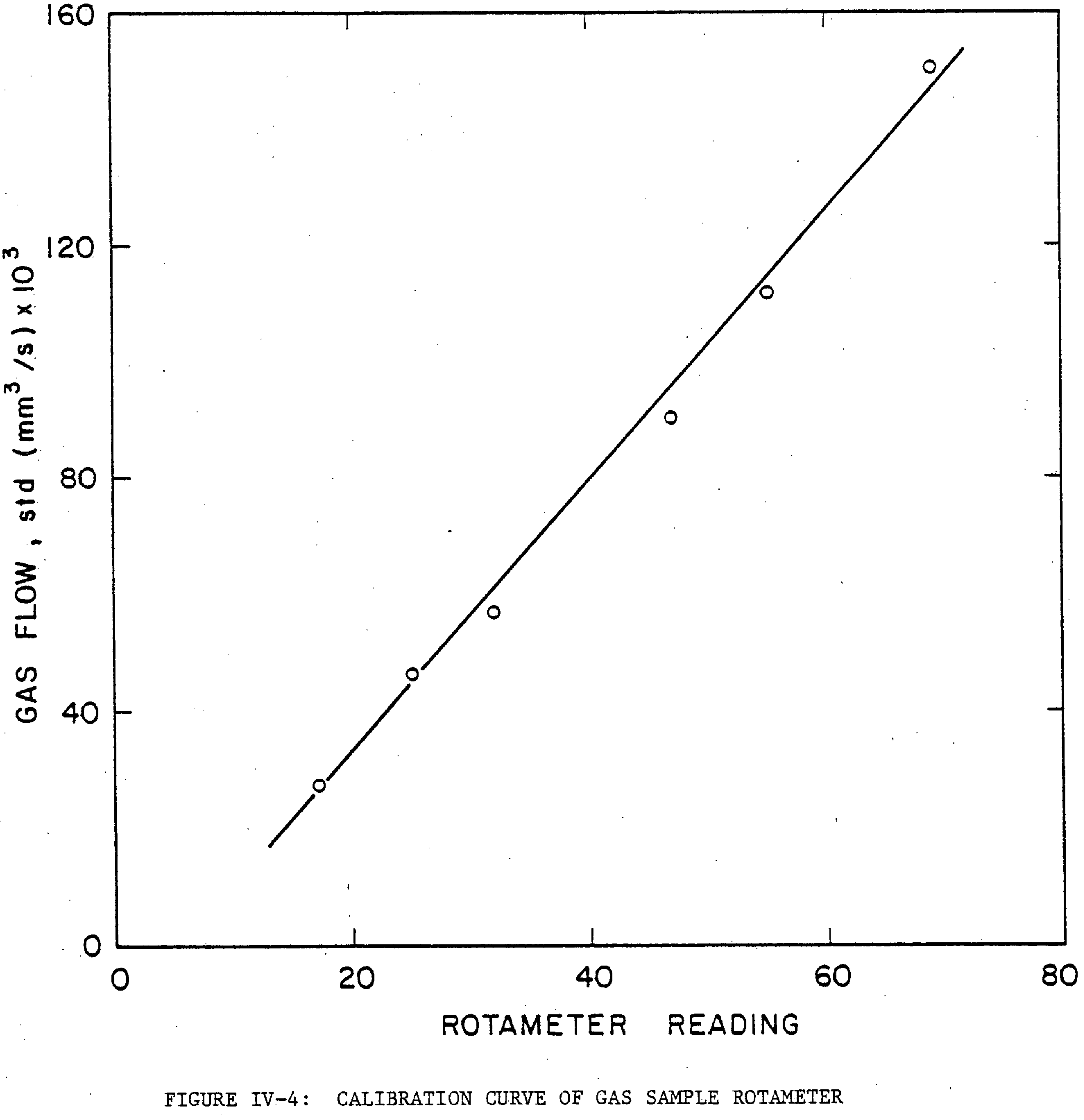


FIGURE IV-4: CALIBRATION CURVE OF GAS SAMPLE ROTAMETER

APPENDIX V

CALIBRATION CURVES FOR GAS ANALYSIS

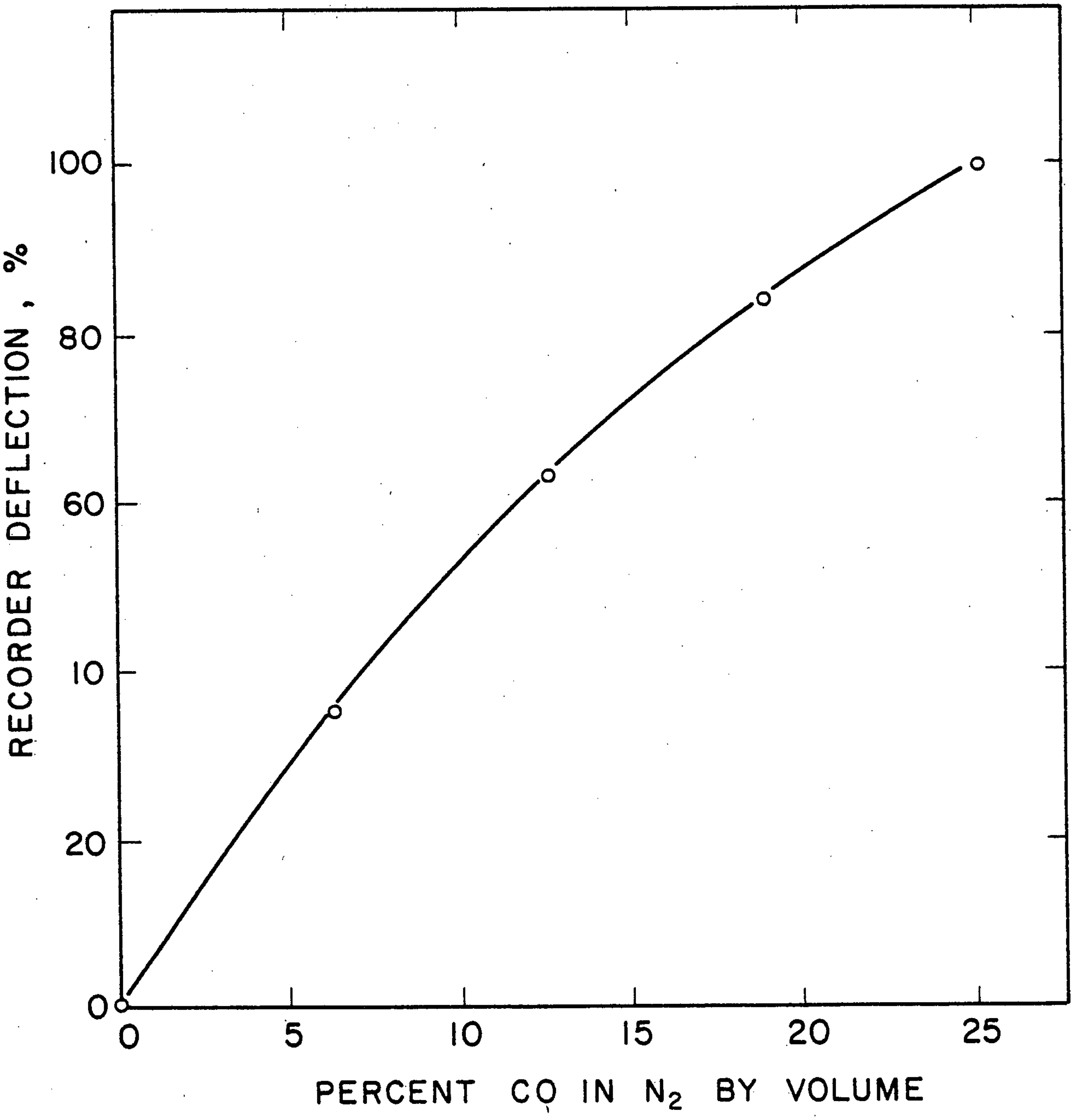


FIGURE V-1: CALIBRATION CURVE OF CONTINUOUS CO ANALYSER

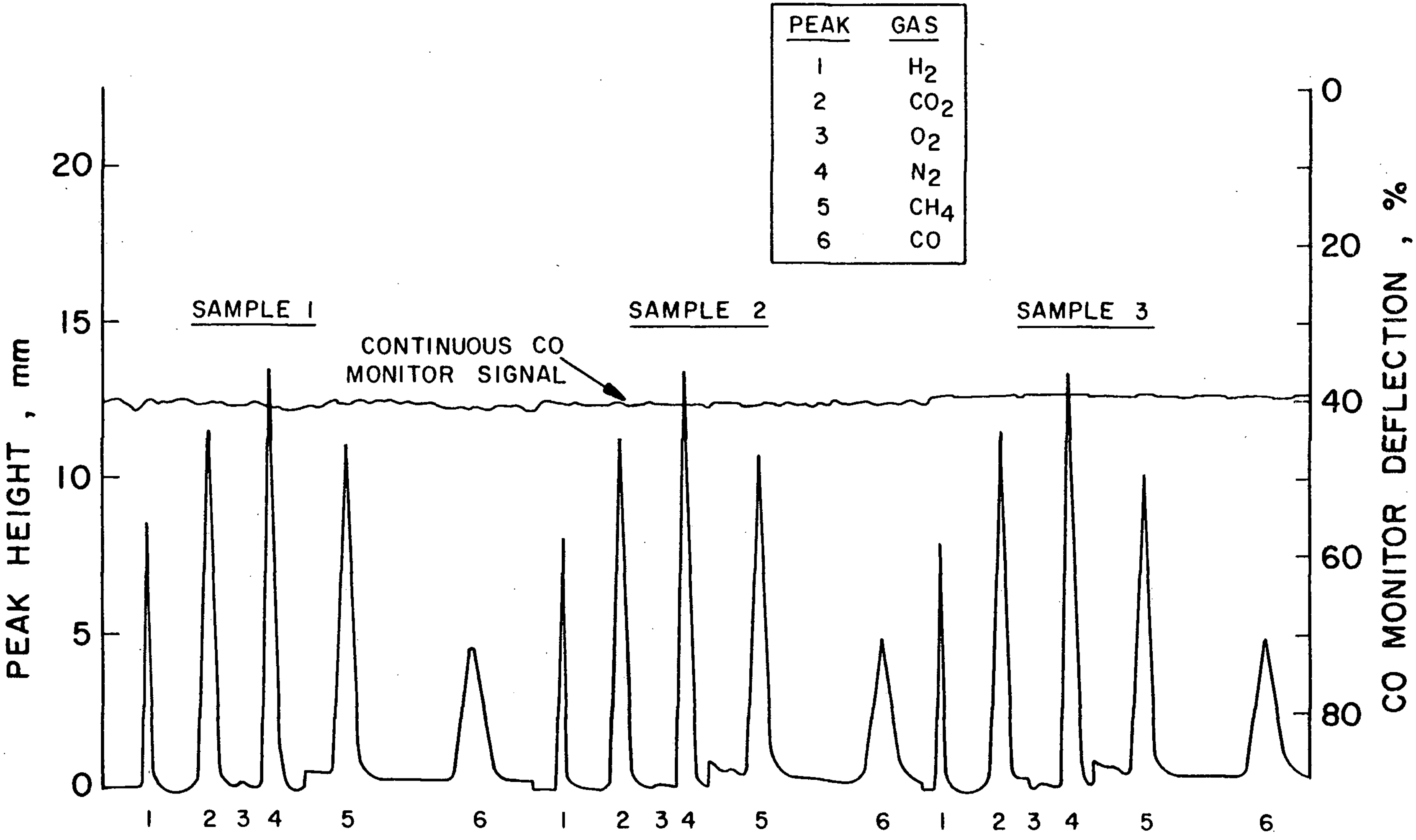


FIGURE V-2: TYPICAL GAS CHROMATOGRAPH TRACINGS, RUN NO. 38

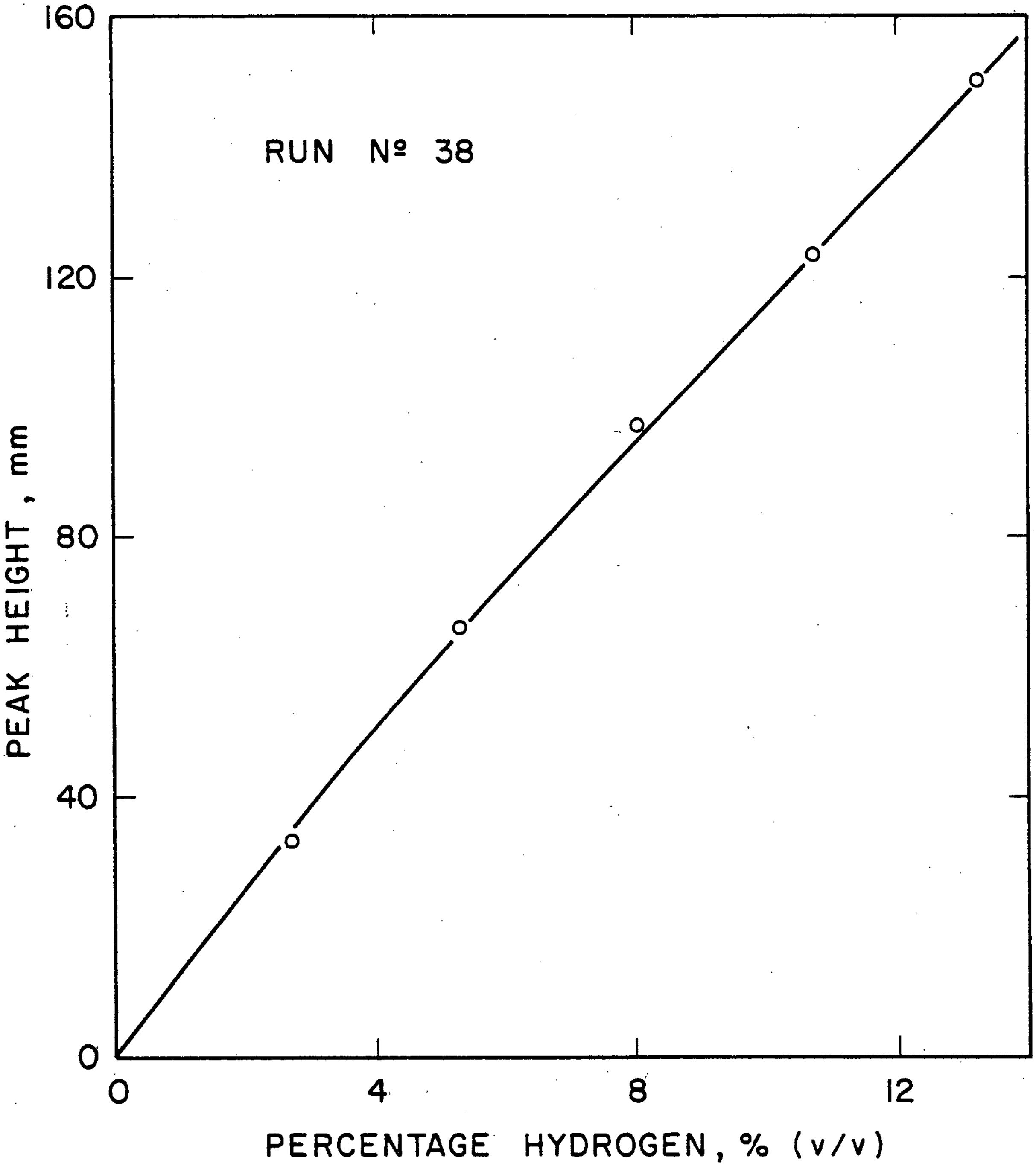


FIGURE V-3: GAS CHROMATOGRAPH HYDROGEN CALIBRATION CURVE

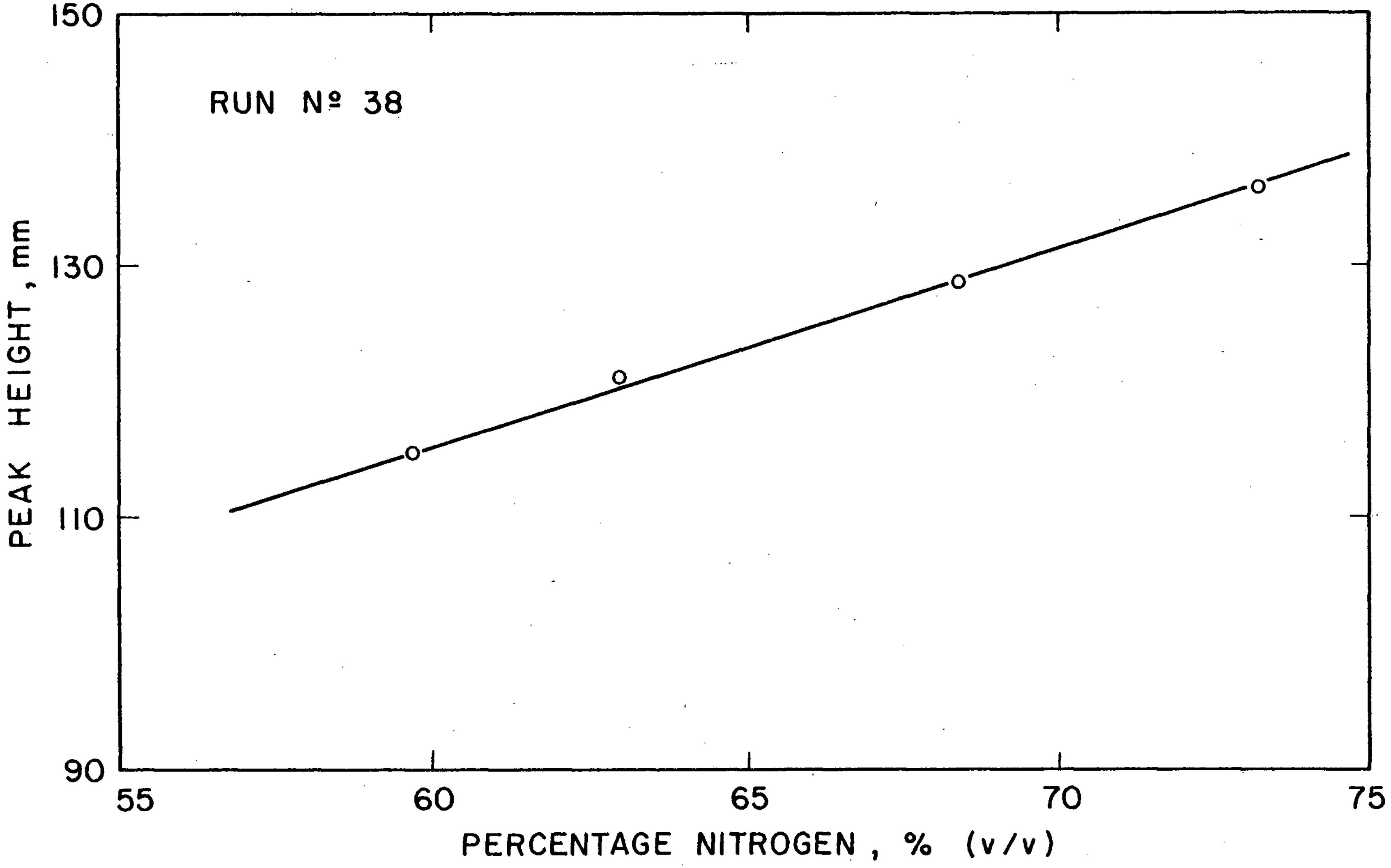


FIGURE V-4: GAS CHROMATOGRAPH NITROGEN CALIBRATION CURVE

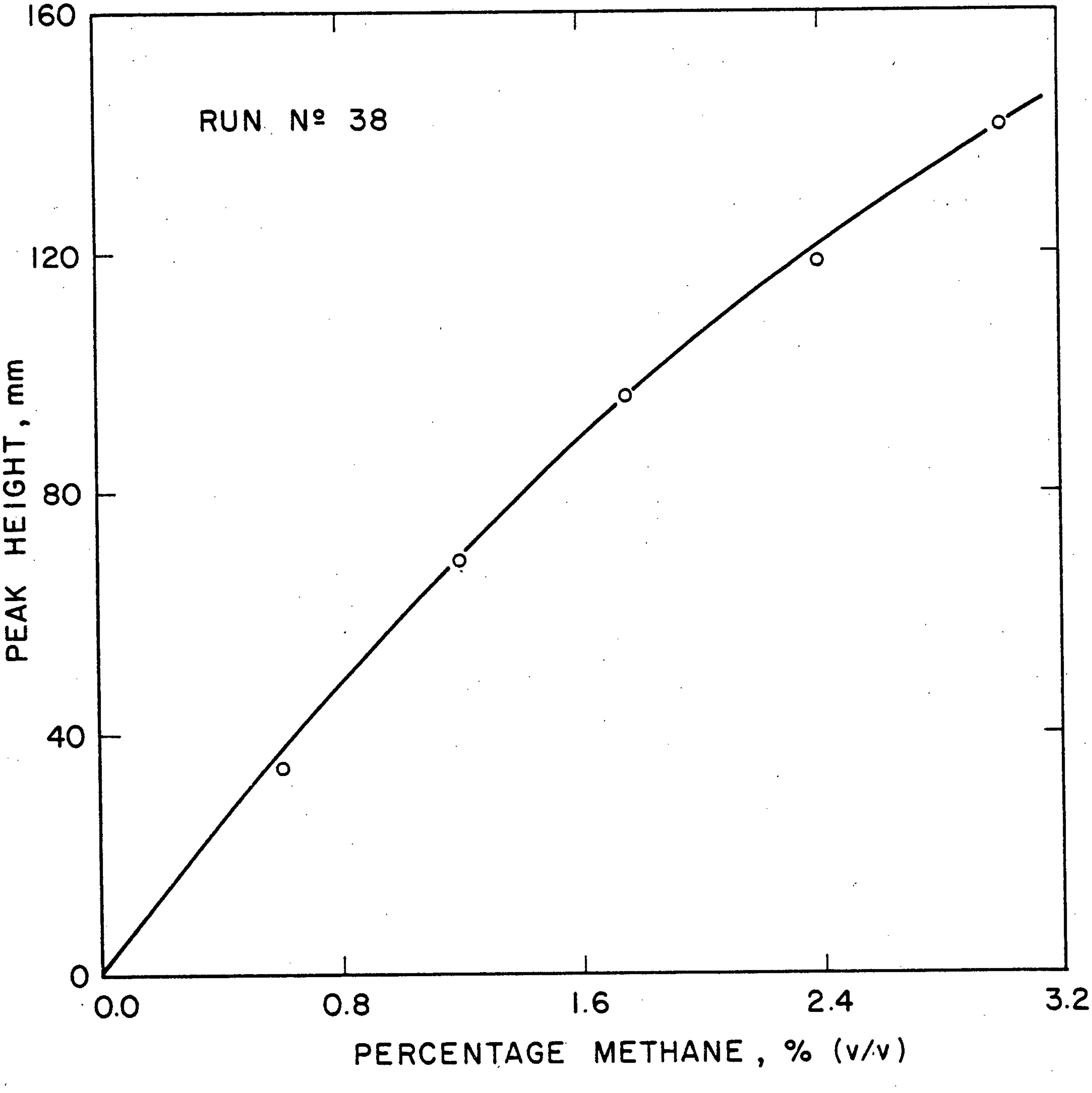


FIGURE V-5: GAS CHROMATOGRAPH METHANE CALIBRATION CURVE

TABLE V-1

COMPOSITION OF STANDARD GAS FOR GAS CHROMATOGRAPH CALIBRATION

Component	Analysis (% v/v)
H_2	13.35
CO	19.98
CO_2	9.93
CH_4	2.95
N_2	53.79

APPENDIX VI

MINIMUM FLUIDIZATION VELOCITY AND SEGREGATION

MINIMUM FLUIDIZATION VELOCITY AND SEGREGATION

TABLE VI-1 SEGREGATION DATA FOR FLUIDIZATION OF A MIXTURE OF 20% 3.36mm. 1.18mm COAL AND 80% 3.36mm. 1.18 mm SILICA

Air Flow std. (m^3/s)	Pressure Drop. across bed (Pa)	Expanded Bed Depth (m)	Observations
0	0	0.229	Packed bed. Well mixed solids.
4.72×10^{-3}	323.8	0.229	Packed bed. Well mixed solids.
7.08	747.2	0.229	Packed bed. Well mixed solids.
9.44	1220.5	0.229	Packed bed. Well mixed solids.
11.80	1868.1	0.229	Packed bed. Very fine solids fluidize at top.
12.74	1893.0	0.235	Packed bed. Finer coal particles fluidize at top.
14.16	2191.9	0.241	Packed bed. 25mm of coal fluidizes at top. Segregation starts.
15.10	1992.6	0.269	0.18m Packed bed and 0.089m of fluid bed with strong segregation of coal at top. Some bubbles.
16.05	1992.6	0.279	Similar to above but bed almost completely segregated with coal in upper 0.1 m section.
16.99	1942.8	0.292	Mildly bubbling bed. Strongly segregated.
17.94	1893.1	0.330	Vigorous bubbling. Strong segregation with some mixing at top.
18.88	1868.1	0.330	Similar to above with increased mixing. Bed starts to slug.
19.82	1868.1	0.406	Slugging bed. No segregation.
20.77×10^{-3}	1843.2	0.432	Vigorous slugging. Excellent mixing. No segregation.

TABLE VI-2: SEGREGATION DATA FOR FLUIDIZATION OF A MIXTURE 20% OF 3.36mm. 1.18mm COAL AND 80% 0.73mm OTTAWA SAND.

Air Flow std. (m^3/s)	Pressure Drop. Across Bed (Pa)	Expanded Bed Depth (m)	Observations
0	0	0.203	Packed bed. Well mixed.
3.78×10^{-3}	1345.0	0.203	Packed bed. Well mixed.
4.72	1643.9	0.203	Packed bed. Well mixed.
5.66	2191.9	0.203	Packed bed. Well mixed.
6.14	2191.9	0.203	Bubbles rise through packed bed. Some coal floats at surface.
6.61	2167.0	0.203	Similar. More frequent bubbles.
7.55	2142.9	0.229	More bubbles. Fair amount of agitation. 25 mm at top of bed enriched in coal.
8.50	2117.2	0.229	Bubbling bed. Good mixing except for some coal at bed's top.
9.44	2117.2	0.241	Vigorous bubbling. Good mixing. Slight coal enrichment at bed's top.
10.38	2067.4	0.292	Vigorous bubbling. No segregation.
11.33	2067.4	0.292	Similar. Sluggish bed.
14.16×10^{-3}	1992.6	0.373	Slugging bed. No segregation.

TABLE VI-3: SEGREGATION DATA FOR FLUIDIZATION OF A MIXTURE OF 15% 3.36 mm. 1.18 mm COAL AND 85% OTTAWA SAND

Air Flow std. (m^3/s)	Pressure Drop Across Bed (Pa)	Expanded Bed Depth (m)	Observations
0	0	0.178	Packed bed. Well mixed.
2.83×10^{-3}	1021.2	0.178	Packed bed. Well mixed.
3.78	1295.2	0.178	Packed bed. Well mixed.
4.72	1768.5	0.178	Packed bed. Well mixed.
5.19	2017.6	0.178	Packed bed. Small air stream breaks through.
5.66	2167.0	0.178	Packed bed. Some channeling. Some coal floats at bed's top.
6.14	2117.2	0.184	Gently bubbling bed. Some 6.4 mm of coal segregates at top.
6.61	2117.2	0.191	Increased bubbling. 63.5 mm of coal segregates at top.
7.55	2117.2	0.203	Bubbling bed. 63.5 mm of coal segregates at top.
8.50	2067.4	0.216	Vigorous bubbling. Enriched top layer of coal starts disappearing.
9.44	2092.3	0.241	Vigorous bubbling. Good mixing. No noticeable segregation.
10.38	2042.5	0.241	Excellent mixing in bubbling bed.
11.33	2017.6	0.241	Similar to above. But, bubble coalescense.
12.27	1967.7	0.318	Similar to above.
13.22	1942.8	0.330	Sluggish bed. No segregation.
16.52×10^{-3}	1868.0	0.419	Slugging bed.

TABLE VI-4: SEGREGATION DATA FOR FLUIDIZATION OF A MIXTURE OF 10% 3.36 mm. 1.18mmm COAL AND 90% 0.73 mm OTTAWA SAND

Air Flow std. (m^3/s)	Pressure Drop Across Bed (Pa)	Expanded Bed Depth (m)	Observations
0	0	0.171	Packed bed. No segregation.
2.83×10^{-3}	996.3	0.171	Packed bed. No segregation.
3.78	1320.1	0.171	Packed bed. No segregation.
4.25	1544.3	0.171	Packed bed. No segregation.
4.72	1668.8	0.171	Packed bed. No segregation.
5.19	1967.7	0.171	Packed bed. No segregation.
5.66	2117.2	0.171	Packed bed. Small bubbles rise up.
6.14	2092.3	0.178	Gently bubbling bed. 12.7 mm of coal at top.
6.61	2092.3	0.178	Similar to above.
7.08	2067.4	0.191	More bubbling. Some segregation.
7.55	2067.4	0.197	Still some segregation.
8.02	2067.4	0.203	Enriched coal layer at top of bed starts disappearing.
8.50	2042.5	0.216	Bubbling bed. Hardly any segregation.
9.44	2042.5	0.229	Vigorouse bubbling. No segregation.
10.38	2017.6	0.241	Same as above.
11.33	1967.7	0.254	Bubble coalescence. Slugging starts.
12.74	1942.8	0.279	Slugging bed.
14.16	1917.9	0.318	Similar to above.
16.52×10^{-3}	1843.2	0.356	Slugging bed. Some spouting.

TABLE VI-5: SEGREGATION DATA FOR FLUIDIZATION OF A MIXTURE OF 5% 3.36 mm. 1.18 mm COAL AND 95% 0.73 mm OTTAWA SAND

Air Flow std. (m^3/s)	Pressure Drop Across Bed (Pa)	Expanded Bed Depth (m)	Observations
0	0	0.170	Packed bed. Well mixed.
2.83×10^{-3}	1021.2	0.170	Packed bed. Well mixed.
3.78	1270.3	0.170	Packed bed. Well mixed.
4.25	1494.5	0.170	Packed bed. Well mixed.
4.72	1693.7	0.170	Packed bed. Well mixed.
5.19	1967.7	0.170	Packed bed. Well mixed.
5.66	2167.0	0.170	Packed bed. Well mixed.
6.14	2167.0	0.170	Small bubbles rise through bed. Fine coal floats at surface.
6.61	2142.0	0.178	Gently bubbling bed. Some segregation.
7.08	2142.0	0.178	Similar to above.
7.55	2117.2	0.178	Bubbling bed. Little segregation.
8.02	2117.2	0.191	Bubbling bed. Mixing starts.
8.50	2067.4	0.191	Vigorous bubbling. Hardly any segregation.
9.44	2067.4	0.203	Vigorous bubbling. No segregation.
10.38	2017.5	0.216	Similar to above.
11.80	2017.5	0.216	Some slugging and spouting.
12.74	1992.6	0.279	More slugging.
14.16	1942.8	0.343	Fair amount of slugging.
16.52×10^{-3}	1868.0	0.419	Slugging bed. No segregation.

TABLE VI-6: FLUIDIZATION OF 7.3 mm OTTAWA SAND

Air Flow std. (m^3/s)	Pressure Drop Across Bed (Pa)	Expanded Bed Depth (m)	Observations
0	0	0.178	Packed bed.
2.83×10^{-3}	1170.7	0.178	Packed bed.
3.78	1494.5	0.178	Packed bed.
4.25	1743.6	0.178	Packed bed.
4.72	2017.6	0.178	Packed bed.
5.19	2241.7	0.178	Packed bed.
5.66	2266.7	0.178	Bubbles rise through bed.
6.14	2241.7	0.178	More bubbles.
6.61	2266.7	0.178	Bubbling bed.
7.55	2216.8	0.191	Bubbling bed.
8.50	2216.8	0.191	Vigorous bubbling bed.
9.44	2191.9	0.203	Vigorous bubbling bed.
11.80	2142.1	0.216	Some slugging.
14.16	2117.2	0.343	More slugging.
16.52×10^{-3}	2067.4	0.419	Slugging bed.

APPENDIX VII

GASIFICATION RESULTS

TABLE VII-1 - EXPERIMENTAL RESULTS FOR FORESTBURG COAL GASIFICATION

Run #	Run Duration (h)	Average Coal Particle Size (mm)	Coal Feed Rate Dry Basis (g/s)	Air Flow(a) (m^3/s)	Steam Feed Rate (g/s)	Expanded Bed Depth (m)	Average Reactor Temperature (k)	Gas Composition (%, v/v) Dry Basis					Gross Gas Heating Value(b) (MJ/m^3)	Air Velocity At Average Reactor Temperature(c) (m/s)	Air To Coal Feed Ratio Dry Basis (w/w)	Steam To Coal Feed Ratio(d) Dry Basis (w/w)	Steam To Coal Ratio(e) Dry Basis (w/w)	Average Bed Carbon Content %	Cydone Catch Dry Basis (g/g coal)	Cyclone Catch Carbon Content %	Carbon Elutriation Rate Dry Basis (g/g coal)	Gas Tar Content (g/m^3)
								H_2	CO	CH_4	CO_2	N_2										
1	1	0.53	1.059	4.77×10^{-3}	0.472	0.61	1033	5.5	4.0	0.6	13.5	76.4	1.37	0.92	5.39	0.446	0.526	0.44	0.167	62.31	0.104	-
2*	1.5	0.53	1.060	4.68	0.377	0.61	1054-1127	7.8	8.0	0.5	8.2	73.5	2.09	0.95	5.29	0.356	0.436	-	0.102	55.14	0.056	-
3*	1	0.53	1.060	3.03	0.395	0.51	1031-1013	8.7	5.9	0.7	11.9	72.8	2.03	0.58	3.42	0.373	0.498	-	0.102	55.14	0.056	-
4*	1	0.53	2.077	6.57	0.472	0.61	1055-1034	6.7	5.4	0.9	14.1	72.9	1.80	1.28	3.80	0.227	0.340	-	0.167	62.31	0.104	-
5*	1.5	0.95	0.983	4.19	0.542	0.51	1123-1088	9.0	7.9	0.7	11.4	71.0	2.30	0.86	5.10	0.551	0.632	2.29	0.078	29.50	0.023	-
6*	0.5	0.95	0.983	5.62	0.322	0.36	1093-1059	10.9	9.8	1.1	9.1	69.1	2.91	0.55	3.31	0.328	0.411	-	0.078	29.50	0.023	-
7*	1.0	0.95	0.983	5.44	0.433	0.61	1128	7.7	8.9	0.5	12.2	70.7	2.19	1.15	6.63	0.440	0.526	-	0.078	29.50	0.023	-
8*	1.0	0.95	0.983	5.46	0.729	0.61	1156	5.4	5.9	0.5	12.2	76.0	1.55	1.18	6.65	0.742	0.825	0.50	0.078	29.50	0.023	-
9	1.5	0.95	0.983	7.42	1.667	0.61	1172	0.9	1.9	0.2	16.0	81.0	0.41	1.62	9.04	1.696	1.785	0.26	0.078	29.50	0.023	-
10*	1.7	2.18	0.271	2.75	0.283	0.43	1088	1.0	1.8	0.2	11.5	85.5	0.41	0.56	12.16	1.044	1.399	2.71	0.084	36.79	0.031	-
11	2.0	2.18	0.715	4.54	0.217	0.61	1050	5.4	6.4	0.4	11.4	76.4	1.57	0.89	7.62	0.303	0.645	2.90	0.086	50.44	0.043	-
12	1.7	2.18	0.798	3.49	0.283	0.51	1040	7.3	6.9	0.6	10.2	75.0	1.94	0.67	5.25	0.355	0.698	4.72	0.084	36.79	0.031	-
13	2.0	2.18	0.798	5.48	0.283	0.51	1061	7.3	8.2	0.5	10.8	73.2	2.06	1.08	8.23	0.355	0.698	3.96	0.084	36.79	0.031	-
14	1.0	2.18	1.185	6.51	0.542	0.61	1031	8.6	7.2	0.7	10.9	72.6	2.17	1.25	6.59	0.457	0.797	11.71	0.093	57.99	0.054	0.841
15	1.0	2.18	1.185	6.52	0.250	0.61	1031	7.2	7.2	0.7	10.7	74.2	2.00	1.25	6.59	0.211	0.550	9.89	0.093	57.99	0.054	0.841
16	1.5	2.18	1.273	4.74	0.0	0.69	1107	5.9	12.0	0.7	9.8	71.6	2.42	0.98	4.46	0.0	0.334	2.47	0.091	29.96	0.027	-
17	1.0	2.18	1.273	6.06	0.0	0.86	1165	4.1	8.3	0.6	13.8	73.2	1.72	1.32	5.70	0.0	0.337	0.97	0.091	29.96	0.027	-
18	1.3	2.18	1.273	6.32	0.800	0.86	1145	5.1	6.3	0.6	15.9	72.1	1.60	1.35	5.94	0.628	0.966	0.74	0.091	29.96	0.027	-
19	2.0	2.18	1.673	7.00	0.167	0.61	1049	6.4	8.0	0.7	12.3	72.5	2.00	1.37	5.01	0.100	0.436	10.36	0.086	50.44	0.043	-
20	1.0	2.18	1.699	5.89	0.258	0.50	1108	11.0	8.8	1.0	21.4	57.8	2.77	1.22	4.15	0.211	0.544	-	-	-	-	-
21*	1.0	2.18	2.046	6.25	0.0	0.86	1080	8.4	11.3	0.7	13.0	66.6	2.64	1.26	3.66	0.0	0.332	2.69	0.091	29.96	0.027	-
22	2.0	2.18	2.369	7.92	0.526	0.66	1026	10.2	8.2	0.9	12.6	68.1	2.56	1.52	4.01	0.222	0.555	16.57-31.27	0.093	58.48	0.054	-
23	2.0	4.06	0.397	3.43	0.570	0.40	1108	2.8	3.4	0.3	13.4	80.1	0.86	0.71	10.36	1.436	1.781	1.95	0.083	38.14	0.032	-
24	5.0	4.06	0.782	4.53	0.549	0.61	1096	9.7	9.5	0.5	6.6	73.7	2.50	0.93	6.97	0.702	1.038	4.44	0.093	30.01	0.029	0.274
25	1.3	4.06	1.223	6.36	0.885	0.61	1074	10.1	8.9	0.6	13.4	67.0	2.52	1.27	6.23	0.724	1.059	4.43	0.083	38.14	0.032	-
26	3.0	4.06	1.259	6.11	0.763	0.61	1065	11.6	10.9	0.6	9.1	67.8	2.94	1.28	6.10	0.606	0.940	10.67	0.091	43.35	0.039	-
27	0.7	4.06	1.516	6.67	0.0	0.86	1115	7.7	10.2	0.6	12.2	69.3	2.38	1.39	5.28	0.0	0.332	6.21	0.108	40.96	0.044	-
28	1.0	4.06	1.516	6.74	0.542	0.86	1085	8.5	8.3	0.6	13.8	68.8	2.25	1.36	5.33	0.358	0.690	4.31	0.108	40.96	0.044	-
29	2.0	4.06	1.516	6.75	0.748	0.86	1079	8.7	7.6	0.7	13.9	69.1	2.23	1.35	5.32	0.493	0.826	3.51	0.108	40.96	0.044	-
30	1.5	4.06	1.516	6.76	1.165	0.86	1053	10.0	6.9	0.7	14.1	68.3	2.30	1.33	5.34	0.768	1.101	4.01	0.108	40.96	0.044	-
31	2.6	4.06	1.642	7.75	0.950	0.61	1078	10.9	9.4	0.7	14.0	65.0	2.71	1.56	5.68	0.579	0.912	8.45	0.083	38.14	0.032	-
32	2.0	−4.76+0.0	1.059	4.27	0.0	0.66	1125	6.3	10.1	0.6	10.6	72.4	2.20	0.89	4.84	0.0	0.323	1.52	0.187	61.45	0.115	0.434
33	2.0	−4.76+0.0	1.059	4.42×10^{-3}	0.433	0.69	1055	9.9	7.4	0.7	12.2	69.8	2.35	0.87	5.01	0.409	0.732	1.91	0.187	61.45	0.115	0.434

(*) Unstable condition

(a) All gas flows at 294 K and 101.3 KPa

(b) At North American standard conditions: 288.6 K and 101.6 KPa

(c) Superficial velocity of the air and not the gas

(d) Ratio of steam fed to dry coal fed

(e) Ratio of total water fed (i.e. steam + H_2O in coal + H_2O in air) to coal fed (see Appendix VIII)

TABLE VII-2

EXPERIMENTAL RESULTS FOR FORESTBURG COAL GASIFICATION

Run No.	Run Duration (h)	Average Coal Particle Size (mm)	Coal Feed Rate Dry Basis (g/s)	Air Flow(a) (m^3/s)	Steam Feed Rate (g/s)	Expanded Bed Depth (m)	Average Reactor Temp. (k)	Gas Composition (%) v/v Dry Basis					Gross Gas Heating Value(b) (MJ/m^3)	Air Velocity at Average Reactor Temp.(c) (m/s)	Air To Coal Feed Ratio Dry Basis (w/w)	Steam To Coal Feed Ratio(d) Dry Basis (w/w)	Steam To Coal Ratio(e) Dry Basis (w/w)	Average Bed Carbon Content (%)	Cyclone Catch Dry Basis (g/g coal)	Cyclone Catch Carbon Content (%)	Carbon Elutriation Rate Dry Basis (g/g coal)	Gas Tar Content (g/m^3)
								H_2	CO	CH_4	CO_2	N_2										
34	2.0	2.18	0.554	3.69×10^{-3}	0.508	0.68	1153	4.2	4.9	0.7	13.7	76.6	1.36	0.80	8.12	0.917	0.045	2.68	0.227	81.66	0.185	-
35	1.0	2.18	0.588	5.33	0.358	0.68	1171	5.5	6.6	0.7	11.7	75.5	1.72	1.16	10.86	0.609	0.645	21.4	0.266	85.17	0.227	-
36	1.4	2.18	1.462	5.58	0.0	0.76	1132	4.6	7.0	1.2	12.4	74.8	1.85	1.18	4.58	0.0	0.020	21.7	0.272	80.33	0.219	3.307
37	1.5	2.18	1.462	5.69	0.455	0.76	1081	5.6	6.6	1.2	12.3	74.4	1.92	1.15	4.67	0.310	0.331	27.2	0.272	80.33	0.219	3.307
38*	1.1	2.18	1.580	5.43	0.0	0.76	1165	6.8	7.1	2.0	12.1	72.0	2.43	1.18	4.11	0.0	0.018	18.34	0.272	80.33	0.219	3.307
39	2.0	4.06	1.462	5.68	0.0	0.76	1160	4.3	7.1	1.0	15.3	72.4	1.75	1.23	4.66	0.0	0.021	16.2	0.276	61.88	0.171	0.913
40	1.5	4.06	1.462	5.73	0.358	0.76	1107	6.2	7.0	1.0	15.7	70.0	1.97	1.18	4.71	0.245	0.261	16.2	0.276	61.88	0.171	0.913
								COLEMAN COAL														
41	1.0	2.18	1.968	5.72×10^{-3}	0.0	0.76	1096	7.1	7.8	1.9	10.9	72.5	2.51	1.17	3.48	0.0	0.022	26.8	0.139	71.86	0.100	2.122

(*) Unstable condition
(a) All gas flows at 294 K and 101.3 kPa
(b) At North American Standard conditions: 288.6 K and 101.6 kPa
(c) Superficial velocity of the air and not the gas
(d) Ratio of steam fed to dry coal fed
(e) Ratio of coal water fed (i.e. steam + H_2O in coal + H_2O in air) to coal fed (See Appendix VIII).

TABLE VII-3

BED CARBON CONTENT AS A FUNCTION OF TIME FOR THE GASIFICATION OF FORESTBURG COAL

Run #	Average Coal Particle Size (mm)	Air to Coal Ratio (Dry Basis) (w/w)	Time (min)	Bed Carbon Content (% w/w)
14 & 15	2.18	6.59	22 46 84 109	6.44 8.56 12.86 11.71
17 & 18	2.18	5.82	30 65 225	1.06 0.87 0.74
22	2.18	4.01	22 52 82 107	16.57 20.54 25.16 31.27
24	4.06	6.97	24 56 84 114 144 189 279	8.54 3.15 2.01 4.10 4.54 2.28 6.48
26	4.06	6.[illegible] 6.10	37 67 97 127 182 219	8.72 8.24 9.36 12.61 8.94 10.45
27-30	4.06	5.30	62 92 137 167 227 287 287	3.08 3.94 4.15 3.86 4.63 3.99 4.21
32 & 33	-4.76 + 0.0	4.93	50 110 170 215	0.94 2.09 1.39 2.42

TABLE VII-4

BED CARBON CONTENT AS A FUNCTION OF TIME FOR THE GASIFICATION OF SUKUNKA COAL

Run #	Average Coal Particle Size (mm)	Air to Coal Ratio (Dry Basis) (w/w)	Time (min)	Bed Carbon Content (%, v/v)
36 & 37	2.18	4.63	35	19.66
			70	23.71
			125	27.21
39 & 40	4.06	4.69	52	21.79
			97	10.20
			157	15.56
			187	16.87

TABLE VII-5

GAS COMPOSITION AND GROSS CALORIFIC VALUES FROM FORESTBURG COAL GASIFICATION CORRECTED FOR THE INTRODUCTION OF PURGING NITROGEN

Run #	Gas Composition (%; v/v) (Dry Basis)					Gross Calorific Value (MJ/m^3)
	H_2	CO	CH_4	CO_2	N_2	
1	5.77	4.20	0.63	14.17	75.23	1.44
2*	8.19	8.40	0.53	8.61	74.27	2.19
3*	9.34	6.33	0.75	12.77	70.81	2.18
4*	6.93	5.59	0.93	14.58	71.97	1.86
5*	9.47	8.31	0.74	11.99	69.49	2.42
6*	11.33	10.19	1.14	9.46	67.88	3.02
7*	8.01	9.26	0.52	12.69	69.52	2.28
8*	5.63	6.15	0.52	12.73	74.97	1.62
9	0.93	1.96	0.21	16.54	80.36	0.42
10	1.10	1.97	0.22	12.61	84.1	0.45
11	5.68	6.73	0.42	11.99	75.18	1.65
12	7.78	7.35	0.64	10.87	73.36	2.07
13	7.60	8.54	0.52	11.25	72.09	2.15
14	8.90	7.45	0.72	11.28	71.65	2.24
15	7.45	7.45	0.72	11.08	73.30	2.07
16	6.17	12.56	0.73	10.26	70.28	2.53
17	4.25	8.61	0.62	14.32	72.20	1.78
18	5.28	6.52	0.62	16.46	71.12	1.66
19	6.61	8.26	0.72	12.69	71.72	2.06
21*	8.68	11.67	0.72	13.43	65.50	2.73
22	10.47	8.42	0.92	12.93	67.26	2.63
23	3.05	3.65	0.32	14.37	78.66	0.92
24	10.19	9.98	0.53	6.93	72.37	2.63
25	10.45	9.21	0.62	13.86	65.86	2.61
26	12.00	11.27	0.62	9.41	66.70	3.04
27	7.95	10.53	0.62	12.59	68.31	2.46
28	8.77	8.56	0.62	14.24	67.81	2.32
29	8.98	7.84	0.72	14.34	68.12	2.30
30	10.31	7.12	0.72	14.54	67.31	2.37
31	11.18	9.64	0.72	14.36	64.10	2.78
32	6.63	10.63	0.63	11.15	70.96	2.31
33	10.42	7.79	0.74	12.84	68.21	2.47

* Unstable condition

TABLE VII-6

GAS COMPOSITION AND GROSS CALORIFIC VALUES FROM CAKING COAL GASIFICATION CORRECTED FOR THE INTRODUCTION OF PURGING NITROGEN

Run #	Gas Composition (%, v/v) (Dry Basis)					Gross Calorific Value (MJ/m^3)
	H_2	CO	CH_4	CO_2	N_2	
35	5.85	7.02	0.74	12.44	73.95	1.83
36	4.89	7.44	1.28	13.18	73.21	1.97
37	5.94	7.00	1.27	13.05	72.74	2.04
38*	7.22	7.54	2.12	12.85	70.27	2.58
39	4.56	7.52	1.06	16.21	70.65	1.85
40	6.55	7.40	1.06	16.60	68.39	2.08
41	7.52	8.26	2.01	11.55	70.66	2.66

* Unstable condition

Runs 35-40: Sukunka Coal

Run 41: Coleman Coal

TABLE VII-7

METHANE PRODUCTION FROM THE GASIFICATION OF FORESTBURG COAL

Run #	Coal Particle Size (mm)	Dry Coal Feed Rate (g/s)	Volume of Gas Produced (a) (m^3/s)	CH_4 Concentration In Gas % (v/v)	CH_4 Production (mg/s)
10	2.18	0.271	2.80x10^{-3}	0.2	3.71
11	2.18	0.715	4.97	0.4	13.19
12	2.18	0.798	3.97	0.6	15.80
13	2.18	0.798	6.19	0.5	20.53
14	2.18	1.185	7.35	0.7	34.13
15	2.18	1.185	7.21	0.7	33.48
16	2.18	1.273	5.53	0.7	25.68
17	2.18	1.273	6.82	0.6	27.14
18	2.18	1.273	7.20	0.6	28.66
19	2.18	1.673	7.90	0.7	36.68
21	2.18	2.046	7.73	0.7	35.89
22	2.18	2.369	9.48	0.9	56.60
23	4.06	0.397	3.65	0.3	7.26
24	4.06	0.782	5.14	0.5	17.05
25	4.06	1.223	7.37	0.6	29.33
26	4.06	1.259	7.42	0.6	29.13
27	4.06	1.516	7.89	0.6	31.40
28	4.06	1.516	8.03	0.6	31.96
29	4.06	1.516	8.00	0.7	37.15
30	4.06	1.516	8.11	0.7	37.66
31	4.06	1.642	9.71	0.7	45.09
32	-4.76+0.0	1.059	4.96	0.6	19.74
33	-4.76+0.0	1.059	5.31x10^{-3}	0.7	24.66

(a) Calculated from mass balance,see Table VIII-1

TABLE VII-8

METHANE PRODUCTION FROM THE GASIFICATION OF CAKING COALS

Run #	Particle Size (mm)	Dry Coal Feed Rate (g/s)	Volume of Gas Produced (a) (m^3/s)	CH_4 Concentration In Gas % (v/v)	CH_4 Production (mg/s)
34	2.18	0.544	-	0.7	-
35	2.18	0.588	6.25×10^{-3}	0.7	29.16
36	2.18	1.462	6.34	1.2	50.47
37	2.18	1.462	6.49	1.2	51.66
38	2.18	1.580	6.43	2.0	85.31
39	4.06	1.462	6.66	1.0	44.18
40	4.06	1.462	6.94	1.0	46.04
41	2.18	1.968	6.70×10^{-3}	1.9	84.4

(a) Calculated from mass balance, see Table VIII-1

Runs 34-40: Sukunka Coal

R-n 41: Coleman Coal

TABLE VII-9

HYDROGEN AND CARBON MONOXIDE PRODUCTION FROM THE GASIFICATION OF FORESTBURG COAL AT A STEAM TO COAL RATIO OF 0.32-0.34

Run #	Coal Particle Size (mm)	Dry Coal Feed Rate (g/s)	Volume of Gas Produced(a) (m^3/s)	H_2 Conc. in Gas % (v/v)	H_2 Production (mg/s)	CO Conc. in Gas % (v/v)	CO Production (g/s)
16	2.18	1.273	5.53×10^{-3}	5.9	27.05	12.0	0.77
17	2.18	1.273	6.82	4.1	23.19	8.3	0.66
21	2.18	2.046	7.73	8.4	53.84	11.3	1.01
27	4.06	1.516	7.89	7.7	50.38	10.2	0.93
32	4.76-0.0	1.059	4.96×10^{-3}	6.3	25.91	10.1	0.58

(a) Calculated from mass balance, see Table VIII-1

APPENDIX VIII

MASS BALANCES

MASS BALANCES

GENERAL PROCEDURE:

The procedure consists in calculating the dry volume of gas produced during gasification through a nitrogen balance, and determining the volume of water vapour coming out of the reactor by comparing the dry volume of gas with the measured volume of wet gas. Overall mass balances and mass balance for the different elements can then be carried out. The calculation procedure is detailed below with specific application to Run No. 38. Mass balances for all gasification runs are tabulated in Tables VIII-1- VII-4, Appendix VIII.

SAMPLE CALCULATION RUN NO. 38

Data For Run No. 38

Basis : 1 second

Inputs: : Dry coal feed rate: r_c = 1.580 (g/s) of 2.18 mm Sukunka coal

Coal analysis: See Table 2

Air flow: F= 5.43 10^{-3} std (m^3/s)

Steam feed rate: S= 0.0 (g/s)

Purging Nitrogen: N= 0.435 (g/s)

Outputs: Wet Gas: Orifice upstream pressure: P= 102.83 kPa

Orifice upstream temperature: T= 637 K

Pressure drop across orifice plate: ΔP= 138.01 mm H_2O

Dry Gas composition:

H_2 = 6.8% (v/v)

CO = 7.1%

CH_4 = 2.0%

CO_2 = 12.1%

N_2 = 72.0%

Molecular weight dry gas :$M.W._{dry}$ = 27.83 (g/mol)

Gas composition by elements, dry weight basis:

O = 17.98%

N = 72.19%

H = 0.77%

C = 9.11%

Gas tar content: 3.307 (g/std m^3) dry gas

Cyclone catch: 0.272 g/g dry coal fed

Cyclone catch carbon content: 80.33%

Calculations:

1.- N_2 Mass balance:

Nitrogen input = m

$$= \text{Air flow} \times \rho_{air} \times (\%\ N_2)_{air}(w/w)/100 + \text{coal feed rate} \times (\%N_2 \text{ in coal})(w/w)/100 + \text{purging } N_2 \text{ input}$$

$$= 5.43 \times 10^{-3} \times \frac{28.91}{24.12 \times 10^{-3}} \times 0.7553 + 1.580 \times 0.0126 + 0.435$$

m = 5.372 (g/s).

2.- Total dry gas flow (V) out of the reactor, based on nitrogen input:

$$\dot{V} = \frac{m \times 100.0}{(\%N_2 v/v)_{dry\ gas} \times \rho\, N_2} = \frac{5.372 \times 24.12 \times 10^{-3}}{0.72 \times 28}$$

$$= 6.43 \times 10^{-3} \text{ std } (m^3/s)$$

3.- Total water input (W):

W = water in air + steam + water in coal

$$= F \times \rho_{air} \times (\%\ H_2O\ w/w)_{air}/100 + S + \frac{r_c(\%H_2O)coal}{(1-\ \%H_2O)coal}$$

$$= 5.43 \times 10^{-3} \times \frac{28.91}{24.12 \times 10^{-3}} \times (0.0025) + 0 + \frac{1.580 \times 0.0082}{(1-0.0082)}$$

W = 0.029 (g/s)

4.- Wet gas flow:

Since calculating the wet gas flow from the orifice plate pressure drop requires knowing the molecular weight of the wet gas (see Appendix II) and this is not known until the water content of the gas is known, an iterative method is required. Assume first that all water entering the reactor is unreacted, and calculate the steam flow (W') at the conditions prevailing at the orifice plate.

Let v_s = specific volume of steam (m^3/g)

$= 2.898 \times 10^{-3}\ (m^3/g)$

at temperature and pressure measured at the orifice plate.

then:

$$W' = W \times v_s \quad (m^3/s)$$

$$= 0.029 \times 2.898 \times 10^{-3} = 0.084 \times 10^{-3}\ (m^3/s)$$

Calculate now molecular weight of gas ($M.W._{wet}$) on a wet basis. First calculate volume fraction of steam (s) in wet gas. Since at the orifice plate conditions the dry volume of gas (V') is given by:

$$V' = \frac{V \times T}{294} \times \frac{101320}{P}$$

$$= \frac{6.43 \times 10^{-3} \times 637}{294} \times \frac{101320}{102834} = 13.73 \times 10^{3}\ (m^3/s)$$

then

$$S = \frac{W'}{W' + V'} = \frac{0.084}{0.084 + 13.73} = 0.0061$$

$$M.W._{wet} = M.W._{dry}\,(1-s) + 18\,s \ (g/mol)$$

$$= 27.93\,(1-0.0061) + 18 \times 0.006$$

$$= 27.87\ (g/mol)$$

The flow (Q) measured by the orifice plate is given by (Appendix II)

$$Q = 0.084\sqrt{\frac{\Delta P \times T}{P.M._{wet} \times P}} \qquad (m^3/s)$$

$$= 0.084\sqrt{\frac{138.01 \times 637}{27.87 \times 102834}}$$

$$= 14.71 \times 10^{-3}\ (m^3/s)$$

$$Q \neq W' + V' = 13.81 \times 10^{-3} (m^3/s)$$

Since the measured value of the wet gas (Q) is different from the calculated one under the assumption that the water does not react use the new value of

$W' = Q'-V'$

and iterate until $Q = V' + W'$

in this case, $Q = 14.91 \times 10^{-3} (m^3/s)$

Therefore, the mass of water coming out of the reactor is

$$H_2O = \frac{Q-V'}{vs} = \frac{14.91 - 13.73}{2.898} = 0.407\ (g/s)$$

i.e water is produced in the reactor (water entering the reactor is $W = 0.029$ (g/s)).

5.- Overall mass balance:

Inputs

Coal = r_c = 1.580 (g/s)

Air = $24\ F \times \rho_{air} = 5.43 \times 10^{-3} \times \frac{28.91}{24.12 \times 10^{-3}}$ = 6.508 (g/s)

Water = W = 0.029 (g/s)

Purging N_2 = 0.435 (g/s

Total inputs = 8.552 (g/s)

Outputs

Dry gas = $V \times \rho_{dry\ gas} = \frac{6.43 \times 10^{-3} \times 27.93}{24.12 \times 10^{-3}}$ = 7.446 (g/s)

Water = H_2O = 0.407 (g/s)

Elutriated solids = r_c x cyclone catch

= 1.58 x 0.272 = 0.430 (g/s)

Tar = gas tar content x Vol. dry gas

= $3.307 \times 6.43 \times 10^{-3}$ = 0.021 (g/s)

Total outputs = 8.304 (g/s)

6.- Hydrogen Balance:

Inputs:

Water bound = W x 2/18 = 0.029/9 = 0.003 (g/s)

In dry coal = r_c x (% H_2 in coal w/w)/100

= 1.58 x 0.0445 = 0.070 (g/s)

Total inputs = 0.073 (g/s)

Outputs:

Water bound = H_2O /9 = 0.407/9 = 0.045 (g/s)

In gas = V x ${}_{gas}$ x (% H_2in gas w/w)/100

$$\doteq \frac{6.43 \times 10^{-3} \times 27.93 \times 0.0077}{24.12 \times 10^{-3}} = 0.057 \text{ (g/s)}$$

Total outputs = 0.102 (g/s)

7.- O_2 Balance:

$$\text{Inputs}\begin{cases} \text{Water bound} = W \times 16/18 = \dfrac{0.029 \times 16}{18} & = 0.026\ (\text{g/s}) \\[2ex] \text{In dry coal} = r_c(\%\ O_2 \text{ in coal w/w})/100 & \\ \qquad = 1.58 \times 0.0256 & = 0.040\ (\text{g/s}) \\[1ex] \text{In air} = F \times \rho_{air} \times (\%\ O_2 \text{in air (w/w)}/100 & \\ \qquad \equiv \dfrac{5.43 \times 10^{-3} \times 28.91 \times 0.2314}{24.12 \times 10^{-3}} & = 1.506\ (\text{g/s}) \end{cases}$$

Total inputs = 1.572 (g/s)

$$\text{Outputs}\begin{cases} \text{Water bound} = H_2O \times 16/10 = \dfrac{0.407 \times 16}{18} & = 0.362\ (\text{g/s}) \\[2ex] \text{In gas: } V \times \rho_{dry\ gas} \times (\%\ O_2 \text{ w/w})/100 & \\ \qquad \equiv \dfrac{6.43 \times 10^{-3} \times 27.93 \times 0.1793}{24.12 \times 10^{-3}} & = 1.335\ (\text{g/s}) \end{cases}$$

Total outputs = 1.697 (g/s)

8.- Carbon Balance:

Input: = r_c x (%C in coal w/w)/100 = 1.58 x 0.8031 = 1.269 (g/s)

$$\text{Output}\left\{\begin{array}{ll} \text{in gas} = V \times \rho_{gas} \times (\%\ C \text{ in gas, w/w})/100 & \\ \quad = \dfrac{6.43 \times 10^{-3} \times 27.93 \times 0.0911}{24.12 \times 10^{-3}} & = 0.678\ (g/s) \\ \text{In elutriated solids} = r_c \times (\text{carbon elutriation rate}) & \\ \quad = 1.58 \times 0.219 & = 0.346\ (g/s) \end{array}\right.$$

Total outputs = 1.024 (g/s)

9.- Ash Balances:

Input = r_c x (% Ash in coal w/w)/100 = 1.580 x 0.1081 = 0.171 (g/s)

Output = r_c x Ash elutriation rate

= = 1.580 x (0.272-0.219) = 0.085 (g/s)

TABLE VIII-1 - OVERALL MASS BALANCE

	Inputs					Orifice Plate Measurements				Dry Gas Flow		Outputs					Difference
Run No.	Dry coal (g/s)	Air (g/s)	Water (g/s)	Purging N_2 (g/s)	Total (g/s)	Δ P (mm H_2O)	P (Pa)	T (K)	Wet gas flow(T,P) (m^3/s)	Std (m^3/s)	at T,P (m^3/s)	Water (g/s)	Dry gas (g/s)	Solids (g/s)	Tar (g/s)	Total (g/s)	$\frac{\text{output-input}}{\text{input}}$ (%)
1	1.059	5.717	0.557	0.285	7.618	81.28	102168	527	$11.10\text{x}10^{-3}$	$5.21\text{x}10^{-3}$	$11.02\text{x}10^{-3}$	0.027	6.190	0.177	-	6.394	- 16.1
2	1.060	5.609	0.462	0.285	7.416	93.98	103150	638	12.53	5.18	11.04	0.512	5.847	0.108	-	6.467	- 12.8
3	1.060	3.632	0.475	0.285	5.452	36.83	102325	575	7.35	3.60	6.98	0.141	4.113	0.108	-	4.362	- 20.0
4	2.077	7.875	0.631	0.285	10.868	177.80	104233	703	17.27	7.40	17.21	0.019	8.720	0.347	-	9.086	- 16.4
5	0.983	5.022	0.621	0.285	6.911	22.86	104104	675.3	12.83	4.97	11.11	0.559	5.646	0.077	-	6.282	- 9.1
6	0.983	6.496	0.404	0.285	8.168	48.26	98680	760	8.67	6.49	17.23	0?	7.128	0.077	-	7.205	- 11.8
7	0.983	6.520	0.515	0.285	8.303	93.98	104934	719	12.28	6.37	15.04	0?	7.366	0.077	-	7.443	- 10.4
8	0.983	6.544	0.811	0.285	8.623	86.36	104934	716	12.30	5.94	13.97	0?	7.016	0.077	-	7.093	- 17.7
9	0.983	8.894	1.755	0.285	11.917	372.75	101728	815.8	27.90	7.46	20.62	1.962	9.371	0.077		11.410	- 4.3
10	0.271	3.296	0.379	0.285	4.231	26.25	103240	496.3	5.70	2.80	4.64	0.470	3.432	0.023	-	3.925	- 7.2
11	0.715	5.442	0.461	0.285	6.903	83.82	104080	588	11.12	4.97	9.68	0.538	5.848	0.062	-	6.448	- 6.6
12	0.798	4.183	0.551	0.285	5.817	50.55	104037	533	8.37	3.97	7.01	0.561	4.553	0.067	-	5.181	- 10.9
13	0.798	6.568	0.557	0.285	8.208	140.97	105690	647	15.19	6.19	13.06	0.723	7.127	0.067	-	7.917	- 3.5
14	1.185	7.803	0.944	0.285	10.217	218.44	104843	692.7	19.81	7.35	16.74	0.974	8.356	0.110	0.006	9.446	- 7.5
15	1.185	7.815	0.652	0.285	9.937	199.81	104572	676	18.57	7.21	16.06	0.813	8.298	0.110	0.006	9.227	- 7.1
16	1.273	5.681	0.425	0.285	7.664	159.17	105136	702.7	17.20	5.53	12.74	1.395	6.41	0.116	-	7.921	+ 3.4
17	1.273	7.264	0.429	0.285	9.251	232.75	106942	755.7	20.79	6.82	16.60	1.220	8.222	0.116	-	9.558	+ 3.3
18	1.273	7.575	1.230	0.285	10.363	355.6	107394	793.3	26.91	7.20	18.33	2.376	8.701	0.116	-	11.193	+ 8.0
19	1.673	8.390	0.730	0.285	11.078	266.70	106866	703	19.64	7.90	17.91	0.540	9.233	0.144	-	9.917	- 10.5
21	2.046	7.491	0.679	0.285	10.501	294.0	106942	768	23.97	7.73	19.02	1.423	8.913	0.186	-	10.522	+ 0.2
22	2.369	9.493	1.314	0.285	13.461	312.13	107044	724.7	25.12	9.48	22.11	0.913	10.710	0.220	-	11.843	- 12.0
23	0.397	4.111	0.707	0.285	5.500	40.01	102151	558.3	7.34	3.65	6.87	0.184	4.446	0.033	-	4.663	- 15.2
24	0.782	5.43	0.812	0.285	7.309	100.33	102913	623.3	13.12	5.14	10.73	0.841	5.640	0.073	-	6.555	- 10.3
25	1.223	7.623	1.295	0.285	10.426	198.12	105227	704.3	18.77	7.37	17.00	0.552	8.386	0.102	-	9.040	- 13.3
26	1.259	7.323	1.183	0.285	10.050	176.35	104253	674.6	17.65	7.42	16.55	0.359	8.112	0.115	-	8.586	- 14.6
27	1.516	7.995	0.504	0.285	10.300	306.25	105791	761.5	24.52	7.89	19.57	1.427	9.117	0.164	-	10.708	+ 4.0
28	1.516	8.079	1.046	0.285	10.926	355.60	107009	766.3	26.65	8.03	19.62	2.014	9.295	0.164	-	11.473	+ 5.0
29	1.516	8.091	1.252	0.285	11.144	355.60	106694	790.3	27.01	8.00	20.42	1.872	9.250	0.164	-	11.286	+ 1.3
30	1.516	8.103	1.669	0.285	11.573	406.4	107146	787.5	29.19	8.11	20.54	2.410	9.278	0.164	-	11.347	- 2.0
31	1.642	9.289	1.497	0.285	12.713	314.96	106672	758.2	23.79	9.71	24.70	0?	10.999	0.136	-	11.135	- 12.4
32	1.059	5.118	0.342	0.285	6.804	141.39	103669	665.3	16.06	4.96	10.97	1.682	5.755	0.122	0.002	7.561	+ 11.1
33	1.059	5.298	0.775	0.285	7.417	188.65	106028	682.0	18.98	5.31	11.77	2.323	6.009	0.122	0.002	8.456	+ 14.0
36	1.462	6.688	0.029	0.435	8.614	209.55	103607	706.5	19.68	6.34	14.92	1,479	7.529	0.398	0.021	9.429	+ 9.5
37	1.462	6.820	0.484	0.435	9.201	260.35	105622	735.5	22.54	6.49	15.58	2.080	7.641	0.398	0.021	10.140	+ 10.2
38	1.580	6.508	0.029	0.435	8.552	138.01	102834	637	14.91	6.43	13.73	0.407	7.445	0.430	0.021	8.303	- 2.9
39	1.462	6.808	0.031	0.435	8.736	188.81	104945	719.3	18.14	6.66	15.73	0.736	8.073	0.404	0.006	9.219	+ 5.5
40	1.462	6.868	0.389	0.435	9.154	226.91	104775	764.3	20.79	6.94	17.45	0.960	8.273	0.404	0.006	9.643	+ 5.3
41	1.968	6.856	0.044	0.435	9.303	190.50	104775	714.0	$18.70\text{x}10^{-3}$	$6.70\text{x}10^{-3}$	$15.74\text{x}10^{-3}$	0.914	7.703	0.274	0.008	8.899	- 4.3

TABLE VIII-2 - HYDROGEN MASS BALANCES

Run No.	Inputs			Outputs		
	Water bound (g/s)	In dry coal (g/s)	Total (g/s)	Water bound (g/s)	In dry gases (g/s)	Total (g/s)
1	0.062	0.045	0.107	0.03	0.029	0.032
2	0.051	0.045	0.096	0.057	0.038	0.095
3	0.053	0.045	0.098	0.016	0.030	0.046
4	0.070	0.088	0.158	0.002	0.054	0.561
5	0.069	0.042	0.111	0.062	0.043	0.105
6	0.036	0.042	0.078	-	0.071	0.071
7	0.057	0.042	0.099	-	0.046	0.046
8	0.090	0.042	0.132	-	0.032	0.032
9	0.195	0.042	0.237	0.218	0.008	0.226
10	0.042	0.012	0.054	0.052	0.003	0.055
11	0.051	0.030	0.081	0.060	0.026	0.086
12	0.061	0.034	0.095	0.062	0.028	0.090
13	0.062	0.034	0.096	0.080	0.043	0.123
14	0.105	0.050	0.155	0.108	0.061	0.169
15	0.072	0.050	0.122	0.090	0.051	0.141
16	0.047	0.054	0.101	0.155	0.033	0.188
17	0.048	0.054	0.102	0.136	0.030	0.166
18	0.137	0.054	0.191	0.264	0.037	0.301
19	0.081	0.071	0.152	0.060	0.051	0.111
21	0.075	0.087	0.162	0.158	0.063	0.221
22	0.146	0.101	0.247	0.101	0.094	0.195
23	0.079	0.017	0.096	0.020	0.010	0.030
24	0.090	0.033	0.123	0.093	0.046	0.139
25	0.144	0.052	0.196	0.061	0.069	0.130
26	0.131	0.054	0.185	0.040	0.079	0.119
27	0.056	0.064	0.120	0.159	0.058	0.217
28	0.116	0.064	0.180	0.224	0.065	0.289
29	0.139	0.064	0.203	0.208	0.053	0.261
30	0.185	0.064	0.249	0.268	0.076	0.344
31	0.166	0.070	0.236	0.0	0.099	0.099
32	0.038	0.045	0.083	0.187	0.031	0.218
33	0.086	0.045	0.131	0.258	0.050	0.308
36	0.003	0.065	0.068	0.164	0.037	0.201
37	0.054	0.065	0.119	0.231	0.043	0.274
38	0.003	0.070	0.073	0.045	0.057	0.102
39	0.003	0.065	0.068	0.082	0.035	0.117
40	0.043	0.065	0.108	0.107	0.047	0.154
41	0.005	0.087	0.092	0.102	0.061	0.163

TABLE VIII-3 - OXYGEN MASS BALANCES

Run No.	Inputs				Outputs		
	Water bound (g/s)	In dry coal (g/s)	In air (g/s)	Total (g/s)	Water bound (g/s)	In dry gas (g/s)	Total (g/s)
1	0.495	0.172	1.323	1.990	0.024	1.072	1.096
2	0.411	0.173	1.298	1.882	0.455	0.838	1.293
3	0.422	0.173	0.841	1.436	0.125	0.709	0.834
4	0.561	0.338	1.823	2.722	0.017	1.651	1.668
5	0.552	0.160	1.162	1.874	0.497	1.012	1.509
6	0.359	0.160	1.504	2.023	-	1.205	1.205
7	0.458	0.160	1.767	2.385	-	1.407	1.407
8	0.721	0.160	1.648	2.529	-	1.194	1.194
9	1.560	0.160	2.058	3.778	1.744	1.677	3.421
10	0.337	0.044	0.763	1.144	0.418	0.461	0.879
11	0.410	0.116	1.259	1.786	0.478	0.963	1.442
12	0.490	0.130	0.068	1.588	0.499	0.719	1.218
13	0.495	0.130	1.520	2.145	0.643	1.224	1.867
14	0.839	0.193	1.806	2.838	0.866	1.414	2.280
15	0.580	1.193	1.809	2.582	0.723	1.368	2.091
16	0.378	0.207	1.315	1.900	1.240	1.161	2.401
17	0.381	2.207	1.681	2.269	1.084	1.628	2.712
18	1.093	0.207	1.753	3.053	2.112	1.819	3.931
19	0.647	0.272	1.942	2.861	0.480	1.708	2.188
21	0.604	0.333	1.734	2.671	1.265	1.912	3.177
22	1.168	0.386	2.197	3.751	0.812	2.100	2.912
23	0.628	0.065	0.951	1.644	0.164	0.731	0.265
24	0.722	0.127	1.257	2.106	0.748	0.774	1.522
25	1.151	0.199	1.764	3.114	0.491	1.745	2.236
26	1.052	0.205	1.695	2.952	0.319	1.432	1.751
27	0.448	0.247	1.850	2.545	1.268	1.807	3.075
28	0.930	0.247	1.870	3.047	1.790	1.909	3.699
29	1.113	0.247	1.872	3.232	1.664	1.882	3.546
30	1.484	0.247	1.875	3.606	2.142	1.892	4.034
31	1.331	0.267	2.150	3.748	0.0	2.409	2.409
32	0.304	0.172	1.184	1.660	1.495	1.030	2.525
33	0.689	0.172	1.226	2.087	2.065	1.120	3.185
36	0.026	0.037	1.548	1.611	1.315	1.337	2.652
37	0.430	0.037	1.578	2.045	1.849	1.343	3.192
38	0.026	0.040	1.506	1.572	0.367	1.335	1.697
39	0.028	0.033	1.576	1.637	0.654	1.665	2.319
40	0.346	0.033	1.590	1.969	0.853	1.768	2.621
41	0.039	0.097	1.584	1.720	0.812	1.316	2.128

TABLE VIII-4 - CARBON AND ASH MASS BALANCES

Run No.	Carbon				Ash	
	Inputs	Outputs			Input	Output
	Coal (g/s)	Gas (g/s)	Solids (g/s)	Total (g/s)	Coal (g/s)	Solids (g/s)
1	0.707	0.469	0.110	0.579	0.155	0.067
2	0.708	0.430	0.059	0.489	0.155	0.049
3	0.708	0.332	0.059	0.391	0.155	0.049
4	1.387	0.752	0.216	0.968	0.303	0.131
5	0.657	0.495	0.023	0.518	0.121	0.054
6	0.657	0.646	0.023	0.669	0.121	0.054
7	0.657	0.684	0.023	0.707	0.121	0.054
8	0.657	0.550	0.023	0.573	0.121	0.054
9	0.657	0.672	0.023	0.695	0.121	0.054
10	0.181	0.188	0.008	0.196	0.026	0.014
11	0.478	0.450	0.031	0.481	0.068	0.030
12	0.533	0.350	0.025	0.375	0.076	0.042
13	0.533	0.601	0.025	0.626	0.076	0.042
14	0.791	0.688	0.064	0.752	0.113	0.046
15	0.791	0.667	0.064	0.731	0.113	0.046
16	0.850	0.619	0.035	0.654	0.121	0.081
17	0.850	0.770	0.035	0.805	0.121	0.081
18	0.850	0.817	0.035	0.852	0.121	0.081
19	1.117	0.825	0.072	0.897	0.159	0.714
21	1.367	0.962	0.055	1.017	0.195	0.130
22	1.582	1.023	0.128	1.151	0.226	0.091
23	0.265	0.310	0.013	0.323	0.035	0.020
24	0.522	0.425	0.022	0.447	0.069	0.051
25	0.817	0.839	0.039	0.878	0.107	0.063
26	0.841	0.761	0.049	0.810	0.110	0.065
27	1.013	0.903	0.067	0.970	0.133	0.097
28	1.013	0.907	0.067	0.974	0.133	0.097
29	1.013	0.883	0.067	0.950	0.133	0.097
30	1.013	0.875	0.067	0.942	0.133	0.097
31	1.097	1.164	0.053	1.217	0.144	0.084
32	0.707	0.525	0.122	0.647	0.111	0.076
33	0.707	0.537	0.122	0.659	0.111	0.076
36	1.174	0.650	0.320	0.970	0.158	0.077
37	1.174	0.649	0.320	0.969	0.158	0.077
38	1.269	0.678	0.346	1.024	0.171	0.084
39	1.156	0.775	0.250	1.025	0.182	0.154
40	1.156	0.818	0.250	1.068	0.182	0.154
41	1.541	0.689	0.197	0.884	0.213	0.007

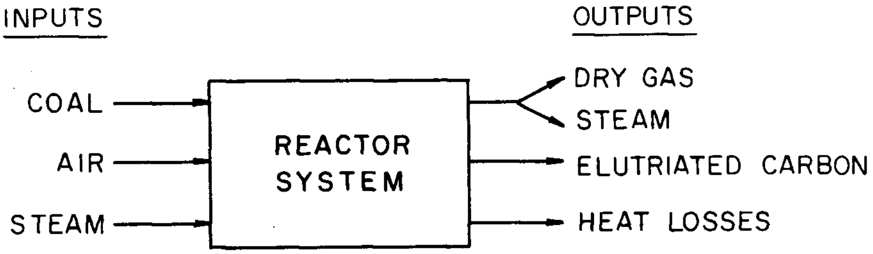
APPENDIX IX

ENERGY BALANCES

ENERGY BALANCES

GENERAL PROCEDURE

Consider the fluidized bed reactor as a "black box" in sketch below:


INPUTS
OUTPUTS
COAL
AIR
STEAM
REACTOR SYSTEM
DRY GAS
STEAM
ELUTRIATED CARBON
HEAT LOSSES

The enthalpy balance is carried out in the basis of one second of operation and with a reference temperature of 288.6 K (60^{o}F = North American standard reference temperature for combustion processes).

The heat content of inputs and outputs is calculated as below and the heat losses are calculated by difference.

Inputs:

1) Heat content of coal:

H_{coal} = Dry coal feed rate (g/s) x heat value of coal (KJ/g) (=) KJ/s

Heat value of coal obtained from analysis (Table 7)

Dry coal feed measured experimentally.

2) Heat content of air:

$$H_{air} = F_{air}\int_{T_{ref}}^{T_1} CpdT = F_{air}\,\bar{C}p_{air}(T_1 - T_{ref}) \times 4.185 \times 10^{-3} \text{ (=) KJ/s}$$

where:

F_{air} = molar air feed rate

$$= \frac{\text{Volumetric air feed rate (std m}^3\text{/s)}}{24.12 \times 10^{-3} \text{ (std. m}^3\text{/mol)}}$$

4.185×10^{-3} = conversion factor (=) KJ/cal

T_1 = air feed temperature (K), measured experimentally

$\overline{Cp}_{air}$ = mean molal heat capacity of air (cal/mol K)

$\overline{Cp}_{air}$ is calculated as the arithmetic mean of Cp_{air} evaluated at temperatures T_1 and T_{ref}, by the expression (32):

$$Cp_{air} = 6.8085 + 0.0008351\ T - 39323.15\ T^{-2}$$

This is a good approximation since $T_1 \sim T_{ref}$.

3) Heat content of steam:

$$H_{si} = S[\ \overline{Cp}_{water}\ (T_{b.p.} - T_{ref}) + \overline{Cp}_{steam}(T_2 - T_{b.p.}\ ) + \lambda_s] \text{ (=)KJ/s}$$

$$= S_i \times h_{si}$$

where:

S_i = mass flow rate of saturated steam (g/s), at 137.89 kPa (20 psia)

h_{si} = specific enthalpy of saturated steam (KJ/s)(T_{ref}= 288.6 K)

T_2 = 381.9 K = temperature of saturated steam at 137.89 kPa

$T_{b.p.} = T_{boiling\ point}$ = 373 K

The specific enthalpy of saturated steam h_{gi} is given in the steam tables (32), but here the reference temperature is 273 K rather than the reference temperature of 288.6 K used here. Therefore, the value given by the tables h_{gi} is related to h_{si} by:

$$h_{si} = h_{gi} - \overline{Cp}_{water}(288.6 - 273)$$

since in this temperature range

$$\overline{Cp}_{water} = 1 \text{ (cal/mol K)} = \frac{4.185 \times 10^{-3}}{18} = 0.23 \times 10^{-3} \text{ (KJ/gK)}$$

and $h_{gi} = 2688.98 \times 10^{-3}$ (KJ/g)

then:

$$H_{si} = S_i \times 2685.39 \times 10^{-3} \quad \text{(KJ/s)}$$

4) Heat content of purging nitrogen is assumed to be negligible

5) Total inputs = $H_{coal} + H_{air} + H_{steam}$.

Outputs:

1) Calorific value (combustion heat) of dry gas:

H_{cg} = Volume of dry gas(m^3/s) x calorific value of the unit volume of gas (KJ/m^3) (=) KJ/s

Volume of dry gas is obtained from the mass balances (Table VIII-1, Appendix VIII), and the calorific value of the unit volume of gas is calculated from the gas composition (See Chapter III, and Tables VII-1 and VII-2, Appendix VII).

2) Sensible heat of dry gas:

$$H_{sg} = G\int_{T_{ref}}^{T_3} Cp_g \, dT = G \times \overline{Cp}_g \, (T_3 - T_{ref}) \times 4.185 \times 10^{-3} \quad (=) \text{ KJ/s}$$

where:

G = molar dry gas flow

$$= \frac{\text{volume of dry gas}}{24.12 \times 10^{-3}} \text{ (moles/s)}$$

T_3= Reactor outlet gas temperature (K), measured experimentally.

4.185×10^{-3} = conversion factor (KJ/cal)

$\overline{Cp}_g$ = mean molal specific heat of the gas in the temperature range $T_{ref} - T_3$ (cal/mol k)

The mean molal specific heat of the gas is approximated as the arithmetic mean of the molal specific heats of the gas at the temperatures T_3 and T_{ref}. In turn, the specific heat of the gas at these temperatures is evaluated as the weighted molal average of the individual gases composing it; i.e.

$Cp_{gas}(T) = \sum_i Cp_i\ (\%\ gi)$

where Cpi = molar specific heat of component i at T and

(% gi) = molal fraction of component i in dry gas.

The specific heats of the individual components at temperature T are obtained from (32):

$Cp_{N_2} = 6.62 + 0.00081\ T$

$Cp_{CO} = 6.60 + 0.00120\ T$

$Cp_{CH_4} = 5.34 + 0.0115\ T$

$Cp_{CO_2} = 10.34 + 0.00274\ T - 195500\ T^{-2}$

$Cp_{N_2} = 6.5 + 0.00100\ T.$

3) Heat of Steam

The enthalpy of the steam is given by $H_{so} = S_o \times h_{so}$ (KJ/s)

where:

S_o = water flow out of the reactor (g/s), calculated from the mass balances (Table VIII-1, Appendix VIII). and

h_{so} = specific enthalpy of steam (KJ/g) at outlet of reactor (T_3, atmospheric pressure, $T_{ref} = 288.6$). h_{so} is obtained from superheated steam tables (32) after correcting for the difference in reference temperature i.e.

$H_{so} = S_o\ (h_{go} - 3.59 \times 10^{-3})$

where:

h_{go} = specific enthalpy of steam (KJ/g), from superheated steam tables.

4.- Heat value of elutriated carbon:

This is given by $H_c = e_c \times h_c$ (KJ/s) where

e_c = elutriated carbon (g/s)which is obtained experimentally. (Tables VII - 1 and VII-2, Appendix VII) and

h_c = heat of combustion of carbon = 32.773 (KJ/g)

5) Total outputs = $H_{cg} + H_{sg} + H_{so} + H_e$

Losses:

Heat losses are obtained by difference between the total energy input and the total energy output.

Sample Calculation, Run No. 38:

A sample calculation following the general procedure outlined above is presented here for run No. 38, the same for which a sample calculation of the mass balance was presented in Appendix VIII.

Energy Inputs:

1) Heat content of coal:

$$H_{coal} = 1.586 \times 31.34 = 50.305 \text{ (KJ/s)}$$

2) Heat content of air:

$$F_{air} = \frac{5.43 \times 10^{-3}}{24.12 \times 10^{-3}} = 0.225 \text{ (moles/s)}$$

T_1 = air temperature = 294.8 K

$$Cp_{air}(T_1) = 6.8035 + 0.0008351 \times 294.8 - 39325.15 \times (294.8)^{-2}$$

$$= 6.602 \text{ (cal/mol K)}$$

$$Cp_{air}(T_{ref}) = 6.8085 + 0.0008351 \times 288.6 - 39325.15 \times (288.6)^{-2}$$

$$= 6.577 \text{ (cal/mol K)}$$

$$\therefore \bar{C}p_{air} = \frac{6.602 + 6.577}{2} = 6.590 \text{ (cal/mol K)}$$

$H_{air} = 0.225 \times 6.590 \times (294.8 - 288.6) \times 4.185 \times 10^{-3}$
$= 0.038$ (KJ/s)

3) Heat content of steam:

$S_i = 0.0$ (g/s)

$H_{si} = 0.2685.39 \times 10^{-3} = 0$ (KJ/s)

4) Total inputs = 50.305 + 0.038 + 0.0 = 50.343 (KJ/s)

Energy Outputs:

1) Calorific value of dry gas:

Calorific value of unit volume of gas:

$h_{cg} = 0.068 \times 12.109 + 0.071 \times 11.997 + 0.02 \times 37.743 = 2.43$ MJ/m^3

$\therefore\ H_{cg} = 6.43 \times 10^{-3} \times 2.43 \times 10^3 = 15.625$ (KJ/s)

2) Sensible heat of dry gas:

$$G = \frac{6.43 \times 10^{-3}}{24.12 \times 10^{-3}} = 0.267 \text{ (mol/s)}$$

$T_3 = 1127.7$ K

Specific heat of individual dry gas components at temperature T_3:

$Cp_{H_2} = 6.62 + 0.00081 \times 1127.7 = 7.533$ (cal/mol K)

$Cp_{CO} = 6.60 + 0.00120 \times 1127.7 = 7.953$ (cal/mol K)

$C_{P_{CH_4}} = 5.34 + 0.0115 \times 1127.7 = 18.309$ (cal/mol K)

$Cp_{CO_2} = 10.34 + 0.00274 \times 1127.7 - 195500 \times (1127.7)^{-2}$
$= 13.276$(cal/mol K)

$Cp_{N_2} = 6.5 + 0.001 \times 1127.7 = 7.628$ (cal/mol K)

$\therefore\ Cp_{gas}\ (T_3) = 7.533 \times 0.068 + 7.953 \times 0.071 + 18.309 \times 0.02$
$+ 13.276 \times 0.121 + 7.628 \times 0.72 = 8.542$ (cal/mol K)

Similarly, Cp_{gas} (T_{ref}) = 7.083 (cal/mol K)

Then: $\overline{C}p_{gas} = \frac{7.083 + 8.542}{2} = 7.813$ (cal/mol K)

$\therefore H_{sg} = 0.267 \times 7.813 \; (1127.7 - 288.6) \times 4.185 \times 10^{-3}$

$= 7.314$ (KJ/s)

3) Heat of steam

$H_{so} = 0.407 \; (4280.76 \times 10^{-3} - 3.59 \times 10^{-3}) = 1.741$ (KJ/s)

4) Heat value of elutriated carbon:

$H_c = 0.346 \times 32.773 = 11.339$ (KJ/s)

5) Total energy output = 15.625 + 7.314 + 1.741 + 11.339

= 36.019 KJ/s

Losses:

Energy losses = 50.343 - 36.019 = 14.324 (KJ/s)

Percentage losses $= \frac{\text{losses}}{\text{input}} = \frac{14.324}{50.343} = 28.5\%$

TABLE IX-1

ENERGY BALANCES FOR TYPICAL GASIFICATION OF FORESTBURG AND SUKUNKA COAL

Run #	Air Inlet Temperature K	Reactor Outlet Temperature K	Energy Inputs (KJ/s)				Calorific Energy Outputs (KJ/s)					Energy Losses (KJ/s)	% Losses Input-Output / Input
			air	coal	steam	Total	Calorific Value of Dry Gas	Sensible Heat of Gas	Elutriated Carbon Heat Content	steam	Total		
12	294.5	907	0.024	20.91	0.760	21.694	7.70	3.199	0.819	2.116	13.834	7.860	36.2
14	291.3	1028.3	0.020	31.06	1.455	32.535	15.95	7.190	2.097	3.940	29.177	3.358	10.3
15	290.8	1026.3	0.016	31.06	0.671	31.747	14.42	7.025	2.097	3.285	26.827	4.92	15.5
16	296.6	1025.7	0.043	33.36	0.0	33.403	13.38	5.366	1.147	5.635	25.528	7.875	23.6
17	296.1	1124.3	0.052	33.36	0.0	33.412	11.73	7.704	1.147	5.209	25.790	7.622	22.8
18	294.2	1130.0	0.040	33.36	2.148	35.548	11.52	8.272	1.147	10.175	31.114	4.434	12.5
19	293.6	1047.3	0.040	43.85	0.448	44.338	15.30	7.987	2.360	2.208	27.855	16.483	37.2
21	294.6	1064.5	0.043	53.62	0.0	53.663	20.41	8.047	1.803	5.876	36.136	17.527	32.7
22	294.3	1026.7	0.052	62.09	1.413	63.555	24.27	9.345	4.195	3.690	41.500	22.055	34.7
23	294.8	997	0.024	10.41	1.531	11.965	3.14	3.437	0.426	0.731	7.734	4.231	35.4
25	294.3	1068.7	0.041	38.05	2.377	40.468	18.57	7.720	1.278	2.285	29.853	10.615	26.2
31	293.5	1075.4	0.043	43.04	2.551	45.634	26.31	10.303	1.737	0.00	38.350	7.284	16.0
36	295.1	1113.7	0.041	46.54	0.0	46.581	11.73	7.047	10.487	6.278	35.542	11.039	23.7
37	295.7	1070	0.046	46.54	1.222	47.808	12.46	6.807	10.487	8.616	38.37	9.438	19.7
38	294.8	1127.7	0.038	50.30	0.0	50.338	15.62	7.314	11.340	1.741	36.015	14.323	28.5
39	294.4	1133	0.038	46.65	0.0	46.688	11.66	7.692	8.193	3.157	30.702	15.986	34.2
40	294.9	1080.7	0.041	46.65	0.961	47.652	13.67	7.484	8.193	4.001	33.348	14.304	30.0

Runs 12-31 : Forestburg coal
Runs 36-40 : Sukunka coal.