FLUIDISED BED GASIFICATION AND PYROLYSIS OF WOODCHIPS

by

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SUMMARY

The work presented in this thesis includes experimental investigation using a basic fluidised bed to gasify woodchips and cold modelling studies to improve the fluid bed reactor dynamics incorporating bed internals, such as draft tubes and jets.

Low grade fuel gas was produced from woodchips as feedstock, in a 154 mm i/d fluidised bed as the main experimental part of the project using air as the gasifying medium. The influence of a number of process variables on the gasification process were studied including fuel feedrates, temperatures and bed heights, with respect to their effects on quality and quantity of the fuel gas produced. It was found that fuel gas of about 6 MJ/Nm^3 can be obtained with temperatures in excess of 700 $^\circ$C and with fuel feedrates in excess of 3.5 times stoichiometric. The process also benefitted from increasing the static bed heights of the fluidised bed, which was due to the better separation of the combustion and gasification zones.

The cold modelling studies coducted using a 2-D glass model employing a draft tube a nd jet system, and using a novel photographic technique produced more realistic data. This showed that both the systems in question produced induced recirculation rates which can be controlled by the process variables such as bed height, bed

and jet velocities. Further studies employing these systems for biomass conversion should prove that a better fuel gas quality and quantity can be achieved. In addition a variety of feedstocks can be utilised using the same reactor configuration.

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CHAPTER ONE
INTRODUCTION

1 INTRODUCTION

1.1 GENERAL

Energy is the convertible currency of technology. The "wheels" of technology are turned by the conversion of energy from one form to another, just as world trade is accomplished by the exchange of convertible currency. The rules governing the conversion of energy, from heat to kinetic energy and then to electricity, for example, are rigorous and well understood. It is a pity that the laws of economics are not as well-defined or reliable.

Although manyparts of the world have become more affluent during the last century or more, this has been associated with a large growth in energy consumption, especially of the fossil fuels, coal, oil and natural gas. In recent decades, coal has beens extensively replaced by oil and gas; and the two latter fuels accounted for almost all the growth in the primary energy use. Figure 1.1, (OLIVER et al.,1983 and DANIEL and FLAVIN, 1983), shows how the world as a whole has grown steadily more dependent on the energy supplies in general, and oil and natural gas in particular. The Figure refers to commercial energy use, and excludes the extensive use in many countries of biomass fuels, which are seldom recorded in official statistics, such as wood, animal and crop wastes.

FIGURE 1·1 : World Energy Use

TWyr$^{(a)}$

Legend:
- COAL
- OIL
- NATURAL GAS
- HYDROPOWER
- NUCLEAR POWER
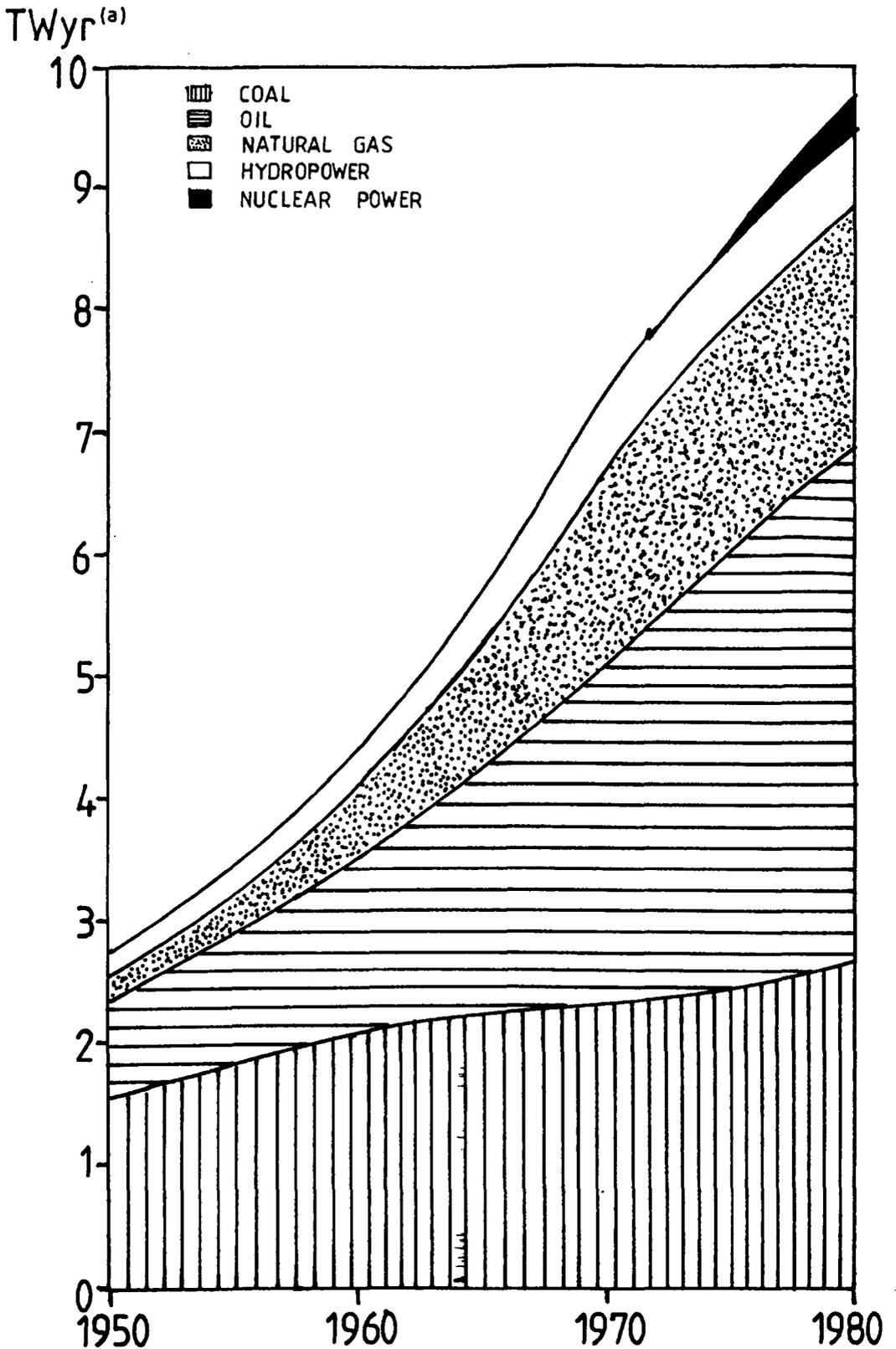
(a) TWyr $\simeq 1 \times 10^{4}$ ton of coal equivalent.

For most of the last decade, the world has been stranded at an energy crossroads. The shocks to the world caused by the oil price increases of the early and late seveties, and the realisation of the limited fossil fuel reserves, have set into motion complex reactions and adjustments that are still unfolding. Few people doubt that energy, lifeblood of civilisation, is on the verge of a major transition. It is less certain what new energy source or sources will take over the central role which fossil fuels now occupy on the energy scene.

Table 1.1, (HAEFELE, 1979), summarizes the global energy supply situation with respect to the energy carrier's production potential and overall resource availability. Total renewable resources in TWyr are infinite, while the annual production potential is 6TWyr . The constraints of renewable energy resources are determined by the environmental and economic aspects.

1.2 RENEWABLE ENERGY

The progress of the last several years marks a coming of the age for renewable energy. Of course, no energy can unfold overnight. Switching from wood to coal during the industrial revolution took most countries a century or more, while several decades were needed to

TABLE 1·1: GLOBAL ENERGY SUPPLY: Resources, production potentials and constraints

	Production (TWyr/yr)	Resource (TWyr)	Constraints
Renewables			
wood	2·5	∞	economy–environment
hydro	1–1·5	∞	economy–environment
total	6–(14)	∞	economy–(nature)
oil and gas	8–12 (?)	1000	economy–environment–resources
coal	10–14 (??)	2000 (?)	society–environment–economy
nuclear			
burner	12 for 2020	300	resources
breeders	≲17 by 2030	300,000	builduprates–resources
fusion	2–3 by 2030	300,000	technology– buildup rates
solar			
soft	1–2	∞	economy–land–infrastructure
hard	2–3 by 2030	∞	buildup rates–materials

1 TWyr ≃ 1 ×10^9 ton of coal equivalent

introduce oil and natural gas. The key to a viable renewable energy-based future is that the world find means to make the transition gradually, phasing in the new fuels before the old ones run out and simultaneously reshaping economics and societies to accept them. The most encouraging aspect of the progress made in the last decade is that it has cleared the way for general change. Energy conservation has provided breathing space while new technologies are developed which will allow a meshing of renewable and conventional energy sources during the decades of transition.

One misconception that seems to spring up again and again is that energy sources must come in large packages. Fifty to hundred years from now historians may well look back at the world's heavy reliance on one fuel as unhealthy anomaly born of the decades of low oil prices. In the future differences in climate, natural resources and economic systems will determine which energy sources will be used in which regions. Even within nations energy supplies will vary from region to region hence making use of several sources of energy.

Table 1.2, (HAEFELE 1979), shows the technical and realisable potential of renewable energy resources such as biomass, wind, hydro and solar.

TABLE 1·2 : Renewable Energy Resoures

Source	Potential		Constraint
	Technical	Realizable	
Forest and Fuel Farms	7·5	2·5	ecological climatological
Solar Panels Soil Storage Heat Pumps	5·0	1·0	economic technological
Hydropower	2·9	1·5	ecological social
Wind	3·0	1·0	economic
OTEC	1·0	0–1	ecological climatological technological
Geothermal	0·4	0·2	economic
Organic Wastes	0·1	0·1	balanced
Glacier Power	0·1	0	technological
Tidal	0·04	0	computational
Total	20	6·3 – 7·3 TW	

TW = 10^{12} W

1.3 BIOGAS

The utilisation of gaseous fuels for commercial and domestic purposes has many advantages over other fuels. There is no emission of particulates and sour gases into the atmosphere, no environmental pollution arising from the transportation and storage of the fuel. It is therefore not suprising that the world consumption of natural gas has trebled since 1962. The natural gas extraction is likely to be increased further, between 20 to 30 per cent during the next two decades (SEAY 1980), to deal with the rising demand. Unfortunately, the world's natural gas reserves are as unequally distributed as its oil reserves, since most natural gas deposits are found togather with oil deposits. Most of the increase in output will occur in just four regions, Mexico, the Soviet Union, the Middle East and North Africa. A few other developing nations have ample reserves, but most poor countries do not. Among industrial countries, natural gas will be a severly limited resource. On balance, natural gas will be a major energy resource for just a few nations, most countries will have to develop other gaseous fuels. The possible alternatives are :

1) Long distance pipeline gas from areas of large natural gas reserves.

2) Imported liquefied natural gas (LNG) and

liquefied peteroleum gas (LPG) shipped from places such as the Middle East.

3) Substituted natural gas (SNG) from coal gasification supplying both commercial and domestic markets.

4) Low and medium calorific value gas produced and consumed at industrial site made from biomass fuels such as peat, wood, municipal refuse and crop waste.

In short all the above alternatives will play an increasing role in maintaining the supply of gaseous fuels to the commercial and domestic markets, after the exhaustion of natural gas and oil. The fourth alternative, the production of low and medium calorific gas from biomass fuels, is the only viable prospect for the long-term.

1.4 THE ENERGY POTENTIAL OF BIOMASS

The total amount of solar energy falling on the land area of the countries of the European Community is of the order of 136,000 million tonnes of oil equivalent (Mtoe) per year (PALZ and CHARTIER 1980). Compared to this the estimated European Community energy consumption of somewhere around 1,000 Mtoe per year is small. Part of the solar energy is fixed in terrestrial and aquatic plants which can become sources of direct or indirect energy. The

tapping of the vast solar energy resource in this way to
supplement European energy supplies, which still depends on
37% of its total energy consumption from imported oil, is
an exciting prospect. However, only a small fraction of
the solar energy is fixed in this biomass as even high
yielding crops have photosynthetic efficiencies of only
a few percent. The use of biomass for fuel also competes
with food (even though the European Community has surplus
of some food supplies) and materials production and
because of this, only a small but significant proportion
of European energy requirements can reasonably be
expected to be met from biomass.

The question of what and how much biomass is
available, and of the technology to use it, is complex.
Biomass has the potential of producing approximately
90 Mtoe of useable energy within the European Community by
the year 2000. This is around 7% of the energy consumption
of the Community in 1985. The potential is given in Table
1.3, (STRUB 1983), which has been calculated as the net
energy produced, making assumptions as to the conversion
efficiency of raw materials to various products, including
heat by direct consumption, gas by thermal or biological
means, and liquid fuels in the form of alcohols (methanol
or ethanol) or pyrolytic oils.

TABLE 1·3: Total Net Energy Potential (after conversion)

RESOURCE	NET (TW) [a]
Wastes and residues	
1) Animal wastes	$16·4 \times 10^{-3}$
2) Crop residues	$17·8 \times 10^{-3}$
3) Forestry/wood residues	$11·3 \times 10^{-3}$
4) Municipal solid wastes	$12·7 \times 10^{-3}$
5) Sewage	$2·6 \times 10^{-3}$
Firewood	$9·4 \times 10^{-3}$
Energy forestry and crops	
Dry (Forestry and short rotation plantations)	$40·1 \times 10^{-3}$ [c]
Wet (catch crops, algae and aquatic plants)	$12·1 \times 10^{-3}$ [c]
Total	$122·4 \times 10^{-3}$ [b]

[a] $TW = 10^{12}$ W.

[b] 7% of the estimated consumption by the year 1985 (EC without Greece).

[c] This potential will be available in 15 years.

1.5 THERMOCHEMICAL CONVERSION OF BIOMASS

1.5.1 Introduction

Like the well-known food chain, the thermochemical conversion chain depends upon the orderly flow of material, in this case biomass, through properly selected and sized successive process steps, as shown in Figure 1.2, (MILES 1983).

1.5.2 Gasification

Thermochemical gasification via partial oxidation is an old art which has been developed to substantially transfer the combustion value of a solid feedstock to a gaseous energy carrier in the form of chemical energy. Gasification reactors fall into four main categories, each having their own distinct advantages and disadvantages:

(1) Fixed or moving beds

(2) Fluidised beds

(3) Entrained or fast fluidised beds

(4) Recirculating or circulating beds

In the studies undertaken in this work, fluidised bed systems were employed initially, as they achieve conditions of high heat transfer rates required for rapid pyrolysis and degradation of fuel particles of varying size and density. The use of an inert bed of sand as a

FIGURE 1·2 : The Biomass Thermochemical Conversion Chain
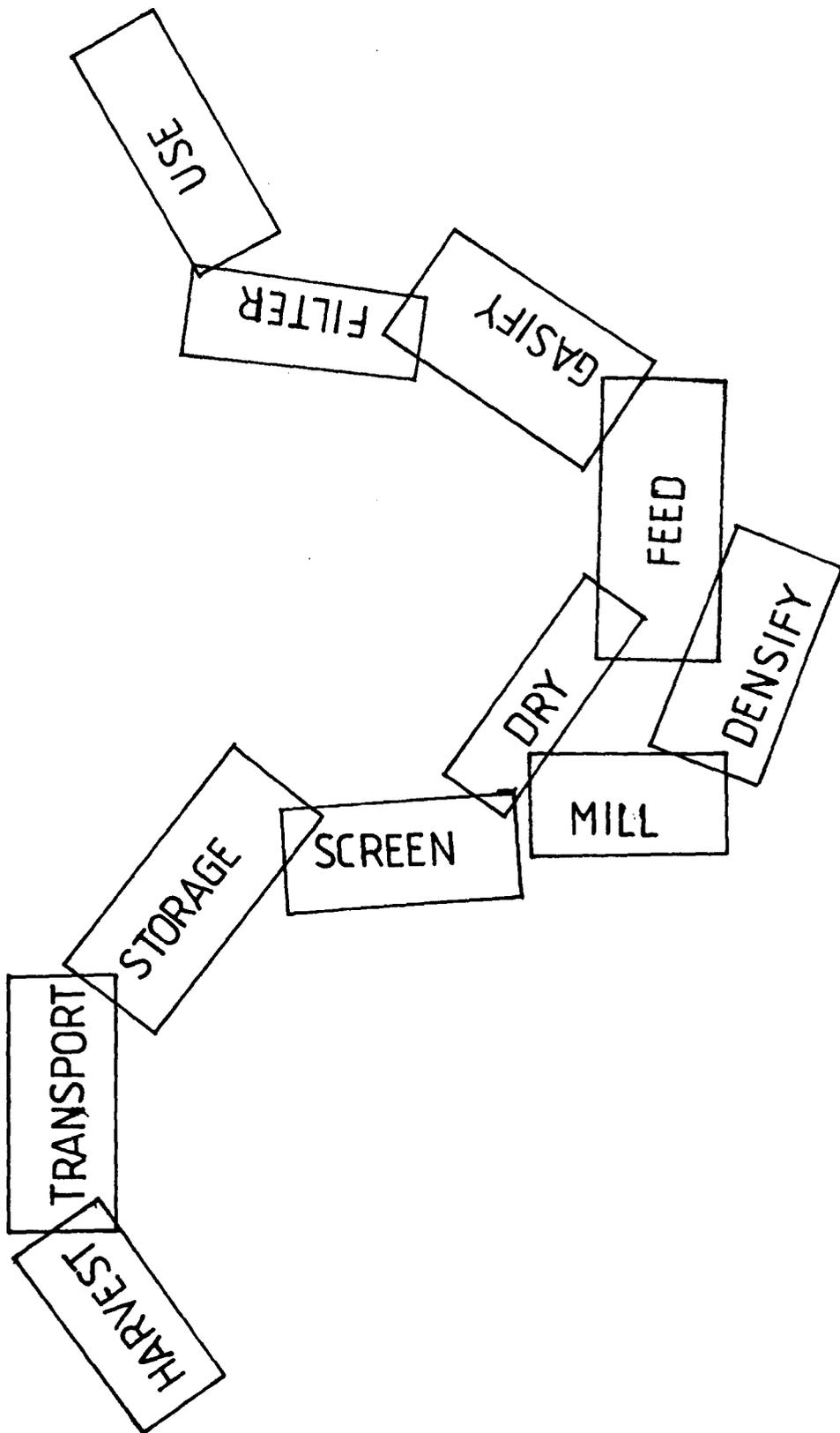
heat carrier has the effect of distributing heat from

regions of combustion to regions of reduction, giving an
overall uniform temperature profile throughout the bed
which prevents the formation of hot spots and subsequent
ash sintering leading to defluidisation and blockage of the
reactor. Later work was performed using recirculation of
bed particles via a jet to yield a greater residence time
for better char combustion.

1.6 OBJECTIVES OF THE PRESENT STUDY

The principle aims of the work presented in this
thesis were two-fold and are presented below:

(1) To investigate experimentally the
gasification/pyrolysis of wood in a 154 mm i/d fluidised
bed to produce a low calorific value fuel gas for the use
as an alternative to coal produced gases or natural gas.
The experiments were designed to study the effects of
varying the following main process parameters:

    (a) Temperature,

    (b) Fuel feed rates (fuel/air ratios), and

    (c) Reactor bed depth.

The results obtained from these studies are presented in
Chapter 3 along with relevant discussion.

(2) To investigate the effect of recirculation of
the bed material on the gasification/pyrolysis process

using a 2-dimensional cold model, incorporating draft tubes
and a jet. From the cold model the influence of the
following parameters on the solid recirculation rates were
determined:

     (a) Fluidising air velocities through the bed,

     (b) Air velocities through the jet, and

     (c) Bed heights.

Results obtained from the cold modelling are discussed in
Chapter 4.

CHAPTER TWO

LITERATURE SURVEY

2 LITERATURE SURVEY

2.1 INTRODUCTION

As the world's resources of naturally occuring oil
and gas are gradually exhausted, there is increasing
worldwide interest in other sources of energy. In
particular renewable sources such as biomass.

Any living materials, either plant or animal, and
the waste produced by them may be described as biomass.
Certain biomass may be converted by physical or chemical
processing into solids, liquids or gases. Which in turn
may be used as fuels to replace conventional coal, oil or
gas supplies. The fuels derived from biomass are called
biofuels. The processes which can be used to convert
biomass into biofuels are based on well proven
technologies, which have previously been developed for
processing more conventional materials such as coal.
Although the technologies themselves are, on the whole,
well understood new factors are introduced when processing
biomass, which require further research and development.

2.2 BIOMASS AS A FUEL

All fossil fuels originated from biomass. One cannot help but be impressed by the major physical and chemical differences from which they derived. Present-day reserves of coal, oil and natural gas have been subjected to extremely cost-free processing operations through the combined action of climate and geological forces over enormous spans of time. Fresh biomass has serious disadvantages as a fuel compared with the fossil fuels. The disadvantages of biomass are :

(1) that biofuels usually have only a modest thermal content compared with fossil fuels,

(2) they often have a high moisture content, which has the effects of inhibiting ready combustion, causing major energy loss on combustion, mainly as latent heat of steam, and also rendering the material putrifiable so that it cannot be readily stored,

(3) they usually have a low density and, in particular, a low bulk density, factors which increase the necessary size of equipment for handling, storage and burning, and

(4) the physical form is rarely homogeneous and free flowing, which militates against automatic feeding to a combustion plant.

The geological conversion processes which formed fossil fuels have increased the thermal content per unit weight relative to the fresh biomass. Virtually eliminating the moisture content, increased the density and bulk density, and converted the material to fluid (as in the case of oil and gas), or a readily handled solid, as in the case of coal.

The advantages of biomass fuels, such as they are, are not connected with suitability in storage and use. They constitute a continually renewable resource whose use leads to no long-term increase in the atmospheric carbon-dioxide. Sometimes they may be cheap and readily available. When this coincides with the material being in a sun dried condition, as it does in gathering firewood in a forest, when a farmer uses straw from his field to heat the farmhouse or when tea plantation workers use bush prunings that have dropped on the geound, then these may, indeed, be real advantages.

2.3 BIOMASS MATERIALS AND RESOURCES

The biomass which could be available as a source of energy in the UK may be divided conveniently into three groups, and the following materials have been considered in the Department of Energy's Biofuel Programme (1982) :

(1) SOLID FUELS

    (a) Domestic refuse

    (b) Industrial waste

    (c) Straw

    (d) Wood waste from timber processing

    (e) Logs and forestry thinnings

    (f) Chicken litter

    (g) Peat

(2) WET AGRICULTURAL WASTES AND VEGETABLE CROPS

    (a) Pig, poultry and dairy waste

    (b) Vegetable residues

    (c) Catch crops

    (d) Perennial crops

(3) FORESTRY

    (a) Wood grown specifically as an energy crop

The estimated quantities of biomass available in the UK is shown in Figure 2.1.

FIGURE 2·1 : Estimates of Quantities of Biomass Available.

ENERGY CONTENT OF BIOMASS		PROCESS	ENERGY CONTENT OF BIOFUEL
DOMESTIC REFUSE	6 Mtce		
INDUSTRIAL WASTE	6 Mtce		SOLID FUEL 15·8 Mtce
STRAW*	2 Mtce		
LOGS/THINNINGS*	1 Mtce		
WOOD WASTE/CHICKEN LITTER*	0·8 Mtce		
ANIMAL WASTES	3 Mtce		
VEGETABLE WASTES	0·6 Mtce	ANAEROBIC DIGESTION	BIOGAS 8·5 Mtce (or SNG)
CATCH CROPS°	5 Mtce		
PERENNIAL CROPS°	3 Mtce		
FORESTRY°	8 Mtce	THERMAL PROCESSING	METHANOL 4·4 Mtce

TOTAL ~35 Mtce   TOTAL ~ 29 Mtce

* Represents estimates of quantities that could be grown and harvested on land available now.

° Represents estimates of quantities that are available for energy use now.

2.4 DIRECT COMBUSTION

The direct combustion of biomass to produce heat should in theory be the most efficient route of utilisation, since with no conversion process involved, there are no conversion losses. On the other hand, the advantage from conversion processes, such as moisture reduction, ease of handling and increase of thermal value, are not available and the boiler or furnace designer must grapple with these problems. Biomass rarely comes in an acceptable form for burning, for example, straw is of low bulk density until compressed and requires chopping or grinding for automatic feeding. Although wood may be burnt as logs, such large pieces generally combust too slowly to be efficient in industrial equipment, boilers generally requiring chipped wood. These preliminary processes of chipping, chopping, grinding, densification or drying are associated with financial costs and energy expenditure. Where these problems have not been fully resolved, the combustion process becomes inefficient by wasting heat in volatilisation of moisture, heat losses from the equipment and demanding excessively large and expensive equipment in relation to heat yield.

Moisture content is the major problem in relation to the combustion of biomass as a fuel. It is commonly stated that biomass material, to combust, should not

contain more than 15% moisture by weight. In practice
though all that is required is that the biomass should
contain sufficient gross thermal value to volatilise the
moisture that is present, to raise it to the temperture of
the flue gases and to keep the combustible material
sufficiently hot to enable combustion to continue.
Although the latent heat of evaporation of the water is
often thought of as the major obstacle, and it is indeed
significant in the calculations, its effect is not as
serious as is often supposed.

Other preparations involved in the processing of
biomass to enable it to combust, are densification and size
reduction.

The choice between direct combustion and any of
the conversion processes is therefore made based upon a
combination of, the fuel moisture content, density, thermal
value and physical form, especially related to mechanical
handling. Ash content is also a parameter of some
importance in comparison to ashless fuels such as oil and
gas.

Once direct combustion has been selected, the
choice of the boiler is related to these same parameters,
though it is also dependent upon the scale of operation and
the end use of the energy.

2.5 ANAEROBIC DIGESTION PROCESS

This process has been known for several centuries and the work of some early chemists in Europe have made reference to it during the late 18th and early 19th centuries. It consists of a microbiological process during which the chemical structure of the waste material is broken down and mixture of gases is released. The gases comprise of carbon-dioxide (about 30%) and methane (about 70%), with traces of hydrogen sulphide. The waste materials are collected in specially designed tanks (digesters) which exclude air from the reactants (anaerobic) and contain the gases produced. The thermal content of the biogas depends upon the specific proportions of its constituents, and is generally described as intermediate, i.e. it is a medium CV gas (26 MJ/Nm^3) compared with SNG (36 MJ/Nm^3).

Anaerobic digestion is a technically simple process requiring relatively straight forward plant, although the biological processes involved are complex, and not well understood. It is especially suited to the treatment of organic materials with a high water content such as animal wastes, plant and vegetable materials. Although digesters are commercially available both in the United Kingdom and on the Continent as means of pollution control and waste disposal, relatively little work has been

directed so far to their optimisation for fuel production.

The fuel gas produced from anaerobic digestion can be used either directly for heating purposes, to generate electricity, or cleaned and upgraded to SNG. One advantage of anaerobic digestion is that the sludge remaining, after the waste has been processed, can be used as a fertiliser. Since it contains a similar amount of plant nutrient to the raw animal waste. It may also be used as an animal feed supplement, being rich in protein.

2.6 FERMENTATION PROCESS

The production of ethanol by fermentation is a well known technique, where the source of carbon in the reaction is sugars derived from complex carbohydrates. Brazil and USA have developed the technology based on their large surpluses of sugar cane, cassava and grain. The alcohol thus produced is used in a blend of 20% alcohol and 80% gasoline by volume, commonly known as Gasohol.

In the UK the potential feedstock for ethanol fermentation would be crops such as sugar beet and cereals. These, however, have a high value in the food market which makes them too expensive for conversion to ethanol.

Another possible route for the production of ethanol is through the use of agricultural and municipal

wastes as a source of cellulose. In the UK, however, the alternative uses of these wastes in direct combustion systems or anaerobic digesters seem to be economically more attractive than the production of ethanol.

2.7 THERMAL PROCESSING

2.7.1 The Objective And Nature Of Thermal Processing

It has already been mentioned that fossil fuels have been derived from biomass through a geological process which overcame the principal disadvantages of biomass as fuel. Conversion of biomass to coal involved a biochemical stage of transformation in which peat was formed under largely anaerobic conditions. Followed by a geological stage in which the peat was subjected to the pressure of superimposed layers of sediment or rock and an increase in temperature. The type of coal was determined by the nature of the original biomass, the degree of anaerobicity of its decay, the temperature and pressures reached during the geological stage and the age of the seam. During these processes the carbon content had been increased and the volatile matter diminished (FRANCIS, 1961). Although a

broadly similar pattern of events probably gave rise to oil deposits and natural gas, the differences, which led to much more hydrogen remaining, are not clearly understood. Attempts to reproduce these conditions in a laboratory have been largely unsuccessful. Therefore thermal processing of biomass may be seen as an attempt to substitute for the natural processes of conversion. Thermal processing has been quite widely applied to coal. i.e. to an already converted fossil fuel, and when this is done the objective is normally a change of state to yield either a liquid or a gaseous fuel.

In a now obsolete process for the production of methanol, acetic acid and acetone, wood was heated in the absence of air to a temperature of about 250 ^{o}C. The feedstock decomposes at this temperature to yield :

(1) wood gas, containing hydrogen, methane, carbon-dioxide and some minor components,

(2) pyroligneous acid, which is an aqueous distillate containing methanol, acetic acid and acetone,

(3) a wood tar, rich in phenolic substances, and

(4) charcoal, consisting of carbon with the original mineral or ash component of the wood.

The yield of the desired components is only 5-6% based upon the original weight of the wood. However, the process used to provide fuel incidentally to its main objective in the form of the gas, which was usually used to heat the

retorts, and charcoal as a combustible solid fuel. This constitutes a simple form of thermal processing although the treatment temperature was very low and the range of products excessively diverse. Processes of this type, conducted in the absence of air and with no other addition to the system, are known as pyrolysis. The nature of the products is markedly affected by the temperature employed and the rate of heating the feedstock.

In Australian work done by McCANN and SADDLER (1976), for example, carried out at 500-650 ^{o}C, a yield of a pyrolytic oil from straw was obtained, amounting to some 39% of the input energy of the feedstock. This was of poor quality, containing a significant amount of oxygen and hence low thermal content per unit weight, and the process was not considered to be economical. Nevertheless, this process serves to illustrate the principle and aims of thermal processing, which always involves converting the inconvenient, low CV and sometimes moist fuel into a dry convenient fuel, either an oil or gas. The advantage applies particularly to process versions in which the feedstock is converted with high efficiency to one form of fuel, because a multiplicity of products complicates marketing and distribution. In practice this usually necessitates one or more of the following steps :

(1) optimisation of operating parameters to maximise
    the percentage of the desired fuel in the primary

products,

(2) introduction of additional reactants into the primary reaction chamber, especially, air, oxygen, steam or hydrogen, and

(3) secondary processes to convert, modify or upgrade some or all of the primary products.

The advantages of the resulting fuels compared to the feedstocks relate to its greater ease of transport and distribution. Leading to its better ability to be metered in a controlled way for automatic feeding purposes, and hence increased effectiveness in electricity generation. To secure the full benefits of these various advantages, quite considerable modifications to primary reaction products are often necessary.

2.7.2 Pyrolysis Or Devolatilisation

This is the basic version of the process, as decribed briefly in the previous section, in which biomass is simply heated in the absence of any air or additional reactants. An amount of char is always formed. Carbonisation is the name given to the process when the amount of char is maximised, the basic reaction being a driving of water from typical biomass at the oxidation level of carbohydrate,

i.e.

$$C_{6n}(H_2O)_{5n} \quad \text{------->} \quad 6nC \quad + \quad 5nH_2O \qquad [2.1]$$
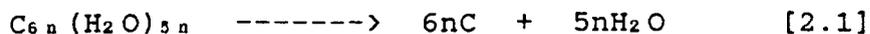
In practice, this reaction cannot be conducted
with high efficiency, being complicated by other reactions,
one of the most important being considered to be :

$$C \quad + \quad H_2O \quad \text{<------->} \quad CO \quad + \quad H_2 \qquad [2.2]$$

which is responsible for generating what are probably the
main primary constituents of pyrolysis gas and the main
components which are obtained in the upper range of
temperatures. Several secondary reactions also occur, such
as the following :

$$2CO \quad + \quad 2H_2 \quad \text{<------>} \quad CH_4 \quad + \quad CO_2 \qquad [2.3]$$

$$C \quad + \quad 2H_2 \quad \text{<------>} \quad CH_4 \qquad [2.4]$$

$$C \quad + \quad 2H_2O \quad \text{<------>} \quad CO_2 \quad + \quad 2H_2 \qquad [2.5]$$

so that methane and carbon-dioxide are also present in
various proportions in the pyrolysis gases, depending upon
conditions, upon the feedstock employed and its moisture
content.

A wide range of gas compositions have been
reported in the literature from various feedstocks, the
interesting changes of gas composition taking place between
500-1000 $^{\circ}$C. Experimental work with municipal waste
and with wood confirms that the volume percent of hydrogen
and carbon-monoxide in the gas increases steadily above 400
$^{\circ}$C until they become almost the sole gaseous products at

temperatures approaching 1000 ^{o}C, (LEWIS and ABLOW,
1976). High yield of gas is also favoured by high
temperatures with yields of oil, char and aqueous
distillate depressed. DOUGLAS et al. (1974) found that
with rapid heating and paper as feedstock, a gas yield on a
weight basis of over 50% of the charge was obtainable at
900 ^{o}C, but less than 10% at 400 ^{o}C, at which
temperature about 35% of the product was char. Work on
heating finely divided biomass with very rapid heating
rates suggests that even higher overall gas yields are
possible. Further work by KNIGHT (1979), using a sawdust
feedstock, found that the yield of oil remained fairly
constant, at around 16-17% of the feed, over the range
540-870 ^{o}C. The rate of heating, however, appears to be
quite significant in relation to the ratio between char and
liquid products. Slower heating rate tending to maximise
the formation of char. In view of the premium values
placed upon liquid fuels, especially for powering vehicle
engines, it may be that the main future role for simple
pyrolysis processes will be in the production of liquid
fuels. Since gas is more efficiently obtained from
processes using air, oxygen or other additional reactants.
However, the poor properties of the primary pyrolytic oil,
are further aggravated by its corrosiveness, high ash
content and a tendency to polymerise. Therefore
considerable modification after pyrolysis would be needed

to make an acceptable fuel for the commercial market.

2.7.3 Gasification

This is in principle a similar process to pyrolysis except, in this case the feedstock is heated in a controlled atmosphere of air, oxygen, steam or any other reactant or reactants, which allows a more efficient conversion. Detailed discussion on the gasification processes available will be conducted in the next section.

2.7.4 The Role Of Thermochemical Processing

Recent years has seen renewed interest in thermochemical processing of biomass. In principle almost any reasonably dry form of biomass can be thermochemically processed, at scales ranging from 0.25t/hr to 40t/hr, but for evaluation purposes the practical combinations can be grouped into two types of systems according to WHITE and PLASKETT (1981) :

(a) Centralised systems for production of chemicals and high grade fuels, and

(b) Local systems for producing low calorific value gas.

2.7.4.1 Centralised systems

The centralised systems would be essentially conventional chemical plants, but on a smaller scale. They are likely to operate continuously producing upwards of 500t/day of refined fuels and chemicals to tight specifications for sale to third parties.

2.7.4.2 Local systems

The local systems would be small scale, producing gas from 0.5-10t/hr of biomass. They would be technically simpler and easy to set up and operate. Since the gas would be used in the neighbourhood or even on site as fuel, it may well be loosely specified, and would be produced from whatever biomass source is locally available. The gas users would probably be the equipment owners and operators, using the gases to save on conventional fuels.

Local systems have the major advantage that the cost of raw materials and the distribution of the products are very much reduced. Indeed, at many locations the only cost might be the capital cost of the equipment.

Currently, the main impediments of the deployment of such fuel gas technology is that the technology is not yet ready to be commercially deployed. From an overall viewpoint, thermochemical processing of biomass seems to have quite a bright future. Production of gases is already economic and exploitation of the technology in this market

could provide the necessary market pull to stimulate the biomass supply system.

2.8 BIOMASS GASIFICATION

2.8.1 Introduction

Thermochemical gasification via partial oxidation is an art, over 150 years old, which has been developed to substantially transfer the combustion value of a solid (or liquid) feedstock to a gaseous energy carrier in the form of chemical energy. This is done because a gas offers many advantages as a heat carrier and, if required, the gas produced can be processed to produce liquid fuels. Sometimes, for example in the case of coal gasification, a secondary aim is to remove undesired components from the solid fuel, such as sulphur. However, in biomass gasification this plays no significant role. The main reason for biomass gasification is to produce usable fuel to supplement the depleting supply of fossil fuels. This could play an important role in the energy supplies of the regions in the world where biomass supplies are redundant and where fossil fuel supplies are economically

unobtainable.

It has been necessary to designate a whole chapter for the evaluation of the gasification processes in use in the past and present. To discuss the chemical and physical processes involved during biomass gasification, with particular reference to wood. Also to assess the possible future alternatives regarding biomass gasification processes.

2.8.2 History Of Gasification

The oldest way of gas production from carbonaceous materials is dry distillation, heating up the feedstock in an oxygen-free atmosphere. On this process principle the first coal-gas company started in London in 1812, to produce gas for lighting purposes (HASSAN, 1977).

A second method for producing gas was the (cyclic) water-gas process, developed in about 1880 and applied on an industrial scale from 1900 onwards (FISCHER and GWOSDZ, 1921).

In 1839 Bischof designed a vertical-shaft gasifier for coal and peat gasification, as shown in Figure 2.2 (MEUNIER, 1962). This type of gasifier was improved

further by many other workers, for example Siemens (1857),
and was used in Europe for more than a century (MEUNIER,
1962). The main problems associated with these gasifiers
were the production of tar and ash removal. In 1840
Ebelman suggested that the tar production could be overcome
by leading the product gas through a glowing coke bed.
Based on that principle several gasifiers with one or two
glowing zones, to decrease tar production were realised.

In the period from 1880 to 1920, specially in
Germany, improved designs were made for gasification of low
grade coals like lignite. From the same period some
examples are known of installations for the gasification
of vegetable materials. This was probably the first
attempt at biomass gasification.

FIGURE 2·2 : Counter-current gasifier Bischof (1839)
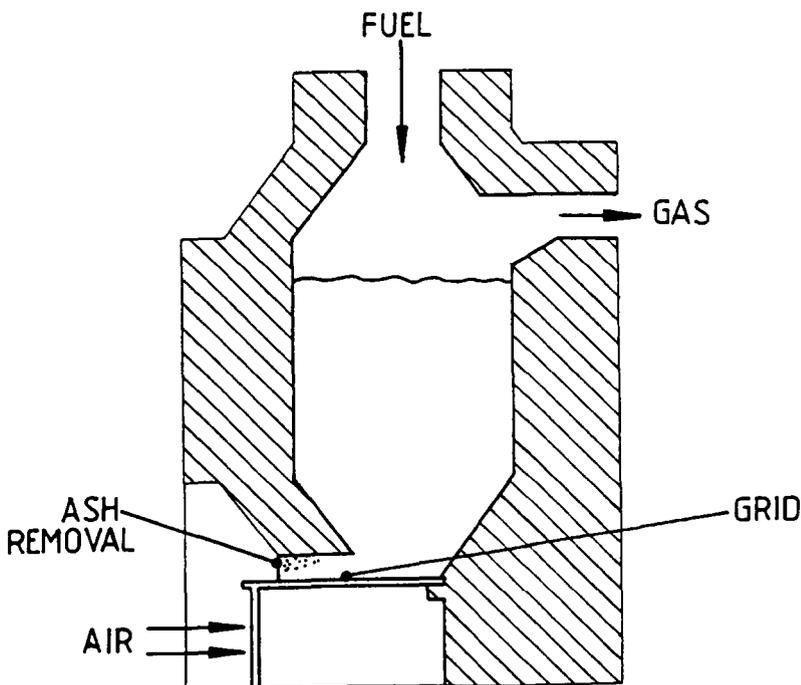
Other gasifiers have been designed specifically to cope with the problem of ash removal, leading to the famous rotating grid type. The first such gasifier was designed by Kerpely in 1904 (FISCHER and GWOSDZ). The powdered coal gasifier of Hirth and Maconnet (1905) was an early example of a new technique for eliminating the ash removal problems as well as the problem of the caking coal in a packed bed. This technique has led to the development of the fluidised bed gasifier by Schmalfeldt and Wintershall in 1940.

During the second world war small gasifiers were constructed for traction of cars and lorries, because of the shortage of liquid hydrocarbons in continental Europe. Most of them used anthracite or charcoals as feedstock but quite a lot used gasified peat or woodblocks. While early gasifiers operated mostly at atmospheric pressures, during and after the war pressurised gasifiers were developed, for example Lurgi (1930) and Koppers-Totzek (1948) (MEUNIER, 1962).

Oil and natural gas took over the predominant position of coal in energy supply, however, and consequently the role of gas production was greatly reduced. Exceptions can be found in regions of the world with special economic circumstances, for example South Africa, parts of USSR and in developing countries.

The energy crisis has however renewed the interest in gasification processes of coal and other potential solid

fuels, in particular biomass. In the next sections various routes of gasification will be discussed with a brief summary of the types of processes available.

2.8.3 The Various Routes To Solid Fuel Gasification

The chemical reactions involved in the gasification of solid fuel are all heterogeneous gas-solid reactions accompanied by relatively large energy changes Questions of material and heat transfer therefore play a prominent part in the design of gasification processes. It is an extremely difficult problem in process engineering to adjust the course of the chemical reactions especially in conjuction with all the other process parameters. It is obvious that numerous compromises are made, or may be imposed by the external constraints.

To achieve complete gasification it is necessary to transfer large quantities of heat to the feedstock in order to provide the heat of reaction. Therefore in the designing of gasifiers, the technique adopted depends on the method of heat transfer. There are basically two different heat transfer processes, autothermal and allothermal. In the autothermal process the heat of reaction is produced inside the gasification reactor itself. While in the allothermal process heat for the endothermic gasification reactions has to be transfered to

the reactor from an external source.

In autothermal processes the composition of the
gasifying medium is generally the means by which the
interplay between individual exothermic combustion
reactions and the endothermic gasification reactions, is
adjusted in such a way that the overall process takes place
in the required manner. Heat, in this process, is most
commonly supplied by the partial combustion of the solid
feedstock with either air, oxygen or a mixture of
air/oxygen/hydrogen. As a result the raw gas will contain
an appreciably higher content of carbon-monoxide and
carbon-dioxide, than is the case with allothermal
processes. The product gas is also diluted with nitrogen
when using air or a mixture including air as the gasifying
medium. On the other hand, autothermal processes possess
the great advantage of avoiding the losses associated with
heat transfer. The design and construction of this type of
gasifier is also simplified. Further, the amount of heat
needed for the reaction can be adjusted quickly and
accurately as required by regulating the oxygen content of
the gasifying medium. On account of these advantages, all
the gasification processes which are at present being
operated on large industrial scale or pilot scale are
without exception autothermal processes.

Allothermal processes, on the other hand, have
losses associated with heat transfer and complexity of

construction, as a consequence have not been extensively developed. One promising application of this process, being developed in Germany by JUNTGEN et al. (1974), is steam gasification of coal which employs process heat from high temperature nuclear reactors and the good heat transfer properties of fluidised beds to feed heat into the gasifier.

Although heat transfer is the essential step in solid fuel gasification processes. In the design of the gasifier, the movement of the solid fuel and the gasifying medium through the system are crucial to which the question of heat transfer has to be subordinated. It seemed useful, therefore, to classify gasification processes on the basis of the mode of conveyance chosen.

Counter-current flow of the solid feedstock and the gasifying medium, is used in most fixed-bed gasifiers. This leads to discrete zones in the bed at different temperatures caused by the exothermic and endothermic reactions dominating at different points. Fresh fuel, in these systems, is fed at the top of the reactor which is preheated by the upward moving hydrogen-containing product gas. As a result the feedstock is carbonised and hydrogenated, leaving the raw gas with relatively large amounts of tars, oils and gaseous hydrocarbons. The tars and oils are undesired products of gasification, and therefore reduce the gasification efficiencies. The char

produced in this carbonisation zone is largely consumed by the very hot steam in the upward flowing medium. The gasifying medium is heated up as passes through the bed of hot ash in the combustion zone with the raw gas being cooled down as it passes through the carbonising and preheating zones. So that in addition to very good conveyancing of material, an excellent heat utilisation is obtained. Figure 2.3 shows a schematic diagram of the counter-current gasification process. The most recent developments with the Lurgi fixed bed gasifier operating at super atmospheric pressures by British Gas at the WESTFIELD DEVELOPMENT CENTRE (1975), has shown that this process can handle most types of caking and non-caking coals.

If a current of air is passed through a bed of fine-grained material at increased blast pressure to achieve a high enough velocity, the solid particles separate from one another and are set in turbulent circulatory motion. This phenomena is known as fluidisation. It is true to say that the majority of the new generation of gasification processes involve fluidised bed reactors in single (SCHILLING et al., 1981), twin and multiple configurations (ANDERSON and TILLMAN, 1979), in which no discrete reaction zones are able to form as they do in a fixed bed, owing to the rapid internal mixing of the solid particles. The fluidised bed, therefore, attains a very homogeneous temperature distribution through its

whole volume. This results in the gasification reactions taking place in a much more uniform manner. In contrast to the behaviour of the solid, however, the composition of the gas varies continuously through the fluidised bed. Combustion taking place mainly in the lower part of the bed. While gasification and pyrolysis reactions taking place in the upper part. The fresh fuel fed is heated extremely rapidly to the reactor temperature on entering the fluidised bed, where pyrolysis takes place along with gasification. The raw gas leaves the reactor with large amount of sensible heat, therefore tar and oil contents are low due to cracking. Although to achieve the highest possible thermal efficiency the sensible heat content of the raw gas must be recovered. Inevitably, due to the grading effect of the gas stream, there is a significant proportion of the fine char particles elutriated from the bed which represents a carbon loss to the system if the dust is not reintroduced to the gasifier.

The fluidised bed process can handle both caking and non-caking coals of various size ranges but cannot efficiently handle fines. On the other hand, entrained flow reactors can only handle finely ground fuel in order to obtain a high degree of conversion due to short residence times of the fuel particles in the reactor. Since there is a need to reach high reaction temperatures the oxygen consumption of this type of gasifier is

FIGURE 2·3

Schematic diagram of a counter-current (fixed bed gasification process

| Wood | → | WOOD [drying zone] | → | Gas |

$CO, H_2, CH_4, C_xH_x,$
$H_2O, CO_2, (N_2$ if
oxygen supply is air)

↓

To purification and/or
combustion

[reaction zone]

[combustion zone] → Ash

Oxygen supply air or O_2 →
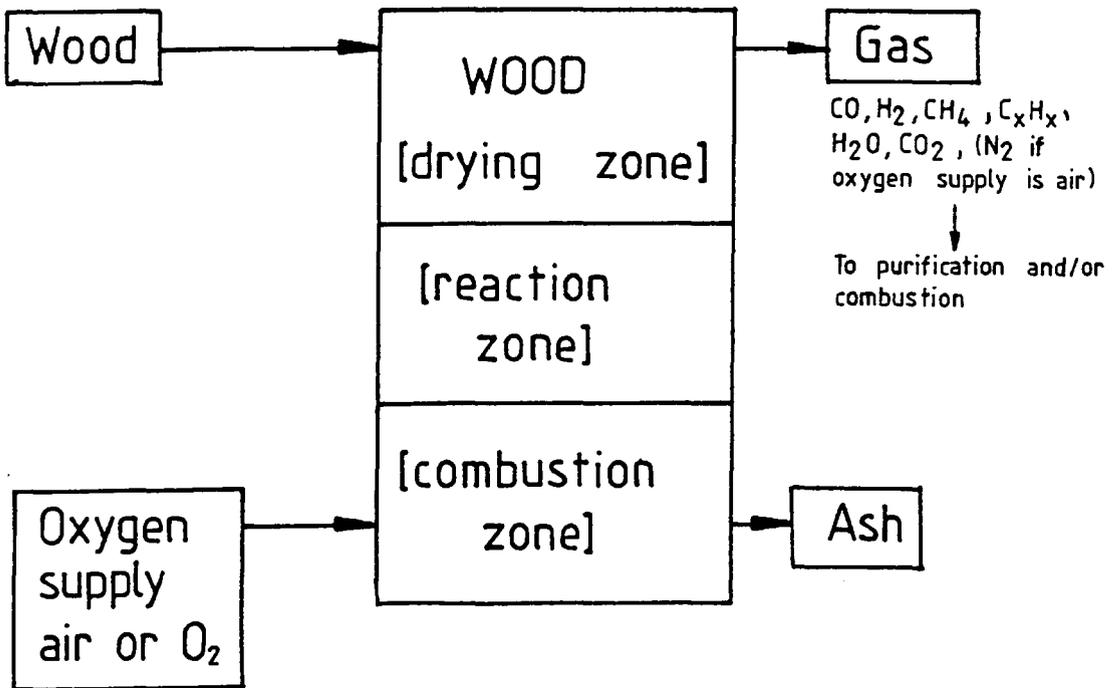
relatively high compared with other types. To overcome unfavourable heat and material transfer conditions from the the co-current flow of fuel and gasifying medium some processes have a reactor section constructed as a vortex chamber with tangential injection, where the difference in inertia between the solid and the gas means that the two phases move relative to one another in the reaction zone and material transfer is improved.

Gasifier types can further be subdivided by the types of gas they produce. This is determined principally by the composition of the gasifying medium and the type of fuel being processed. Commonly gasifying media used are air, oxygen, air/oxygen, oxygen/steam, hydrogen and steam. These directly influence the heating value of the product gas. An air/steam blown gasifier will produce a low CV product gas of less than 8 MJ/Nm^3 comprising of a mixture of hydrogen, carbon-monoxide, carbon-dioxide, nitrogen and small amounts of methane. The nitrogen from the air and carbon-dioixde from combustion could account for 60-70% of the product gas by volume. Resulting in the dilution of the combustible components, carbon-monoxide, hydrogen and methane, thereby reducing the CV. An oxygen or an oxygen/steam blown gasifier would yield a medium CV gas of 10-16 MJ/Nm^3, which does not have the adverse dilution effects of nitrogen and comprises mainly of hydrogen, carbon-monoxide, carbon-dioxide and small quantities of

methane. This gas is the main feedstock for further
processing to give a gas of pipeline or SNG quality by
catalytic methanation. The methanation process requires
the hydrogen/carbon-monoxide ratio to be approximately 3:1,
this therefore requires the conversion of the
carbon-monoxide by the shift reaction to take place (SEGUN,
1975). The end result of this process is a methane rich
(>98%) fuel gas of CV approaching 37 MJ/Nm^3, which is
suitable as SNG.

In all the gasification processes in use and under
development the aim is to produce the highest quality gas
possible within the limitations imposed by the operating
conditions, i.e., fuel quality, gasification medium and
temperature. The gasifier type is then chosen to satisfy
the needs of a particular application of the produced gas.
The general advantages and disadvantages of the gasifier
types are given in Figures 2.4 (a), (b), (c) and (d). A
large number of the new gasification processes, for
gasification of biomass, under development employ fluidised
beds, with the production of a wide range of fuel gases
from low to high CV. Further discussion of particular
processes involved in the gasification of wood will be
presented in a later section.

FIGURE 2·4 (a)

Counter-current moving bed gasifier

FEED INPUT

PRODUCT GAS OUTPUT

PYROLYSIS ZONE

REDUCTION ZONE

OXIDATION ZONE

GASIFYING MEDIUM INPUT

ASH

TEMPERATURE

300   900   1500

$T/K \longrightarrow$

SOLIDS CONVERSION

0   50   100

$\% \longrightarrow$
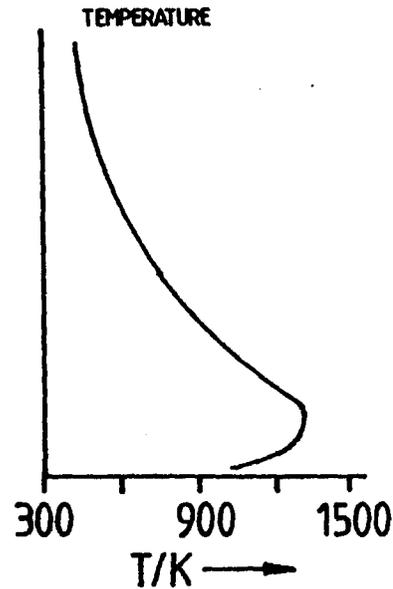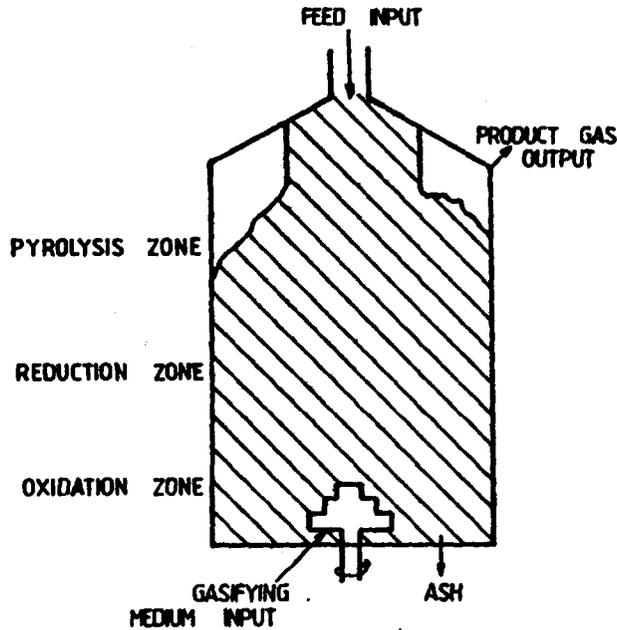
Advantages

1) Simple

2) Low exit gas temperature

Disadvantages

1) High tar content in product gas

2) Channelling and bridging possible

3) Lump or block feed only

FIGURE 2·4 (b)
Co-current moving bed gasifier

GASIFYING MEDIUM
INPUT

FEED INPUT

PYROLYSIS ZONE
OXIDATION ZONE

REDUCTION ZONE

PRODUCT GAS
OUTPUT

ASH

TEMPERATURE

300   900   1500
T/K ⟶

SOLIDS CONVERSION

0   50   100
% ⟶
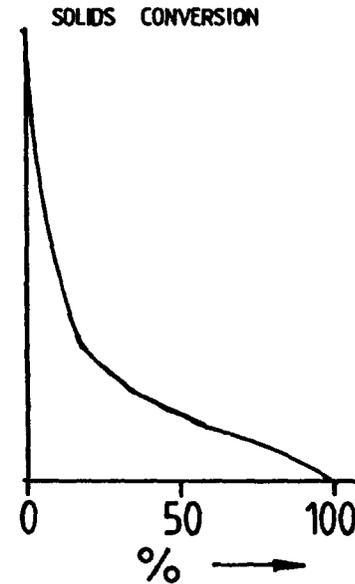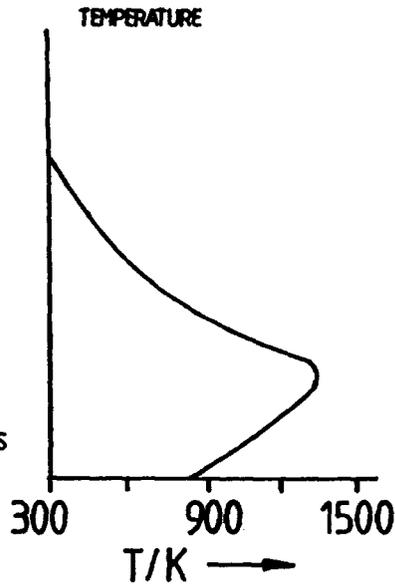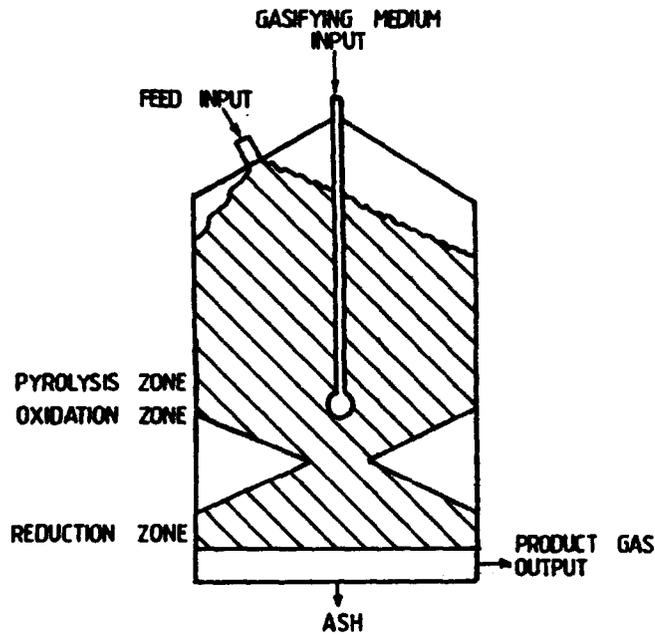
Advantages

1) Nearly no · tar in product
2) Simple

Disadvantages

1) High gas outlet temperature
2) Bridging possible
3) Lump or block feed
4) Limited ash content of feed
5) Difficult to scale up

FIGURE 2·4 (c)
Cross-current moving bed gasifier

FEED INPUT

GASIFYING MEDIUM INPUT

PYROLYSIS

OXIDATION REDUCTION

PRODUCT GAS OUTPUT

ASH

TEMPERATURE

300   900   1500

T/K ⟶

SOLIDS CONVERSION

0   50   100

% ⟶
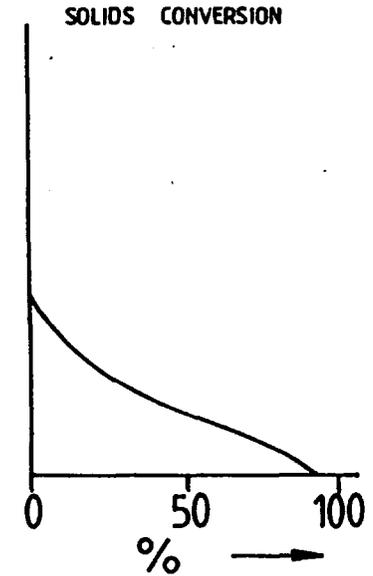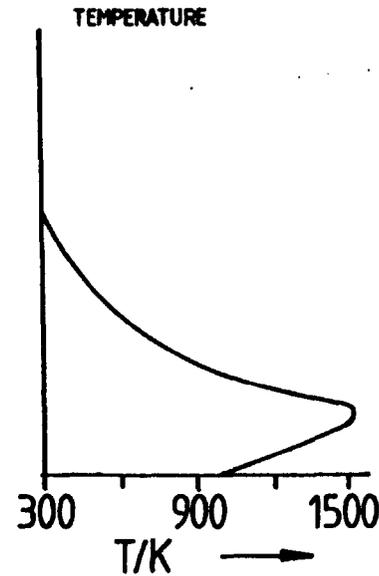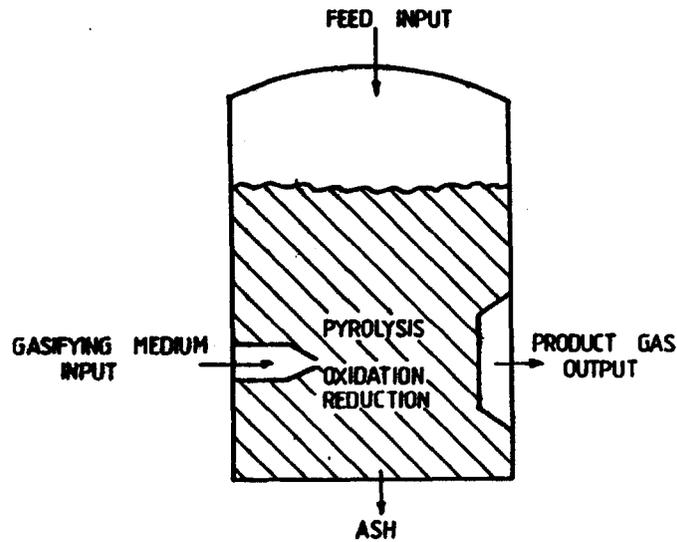
Advantages

1) Simple

2) Low capital cost

3) Light

4) Less sensitive to plugging

Disadvantages

1) Wood char only

FIGURE 2·4 (d)

Fluidised bed gasifier

PRODUCT GAS OUTPUT

→ ASH

PYROLYSIS, OXIDATION
& REDUCTION ZONES

GASIFYING MEDIUM INPUT

TEMPERATURE

300   900   1500

T/K ⟶

SOLIDS CONVERSION

0   50   100

% ⟶
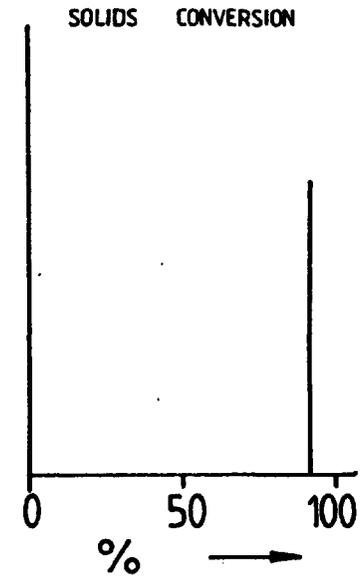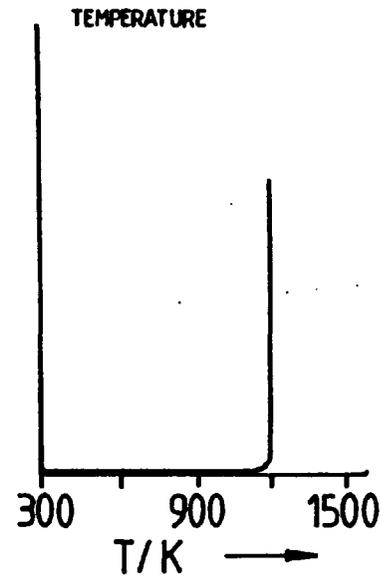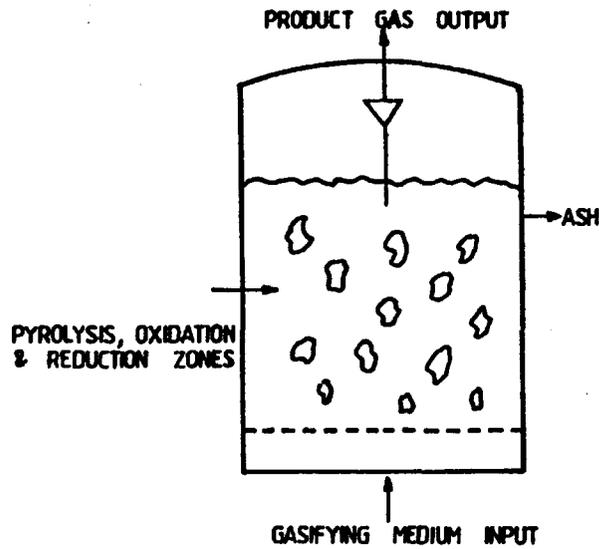
Advantages

1) Can accept wide range of particle sizes

2) High ash fuels can be accepted

Disadvantages

1) High outlet temperature

2) Tars and fine particles in product gas

3) Difficult to control and operate

2.8.4 Chemical And Physical Processes Invloved During Biomass Gasification

The gasification of wood or any other biomass material proceeds by a sequence of steps :

(a) DRYING during which the temperature rises from ambient to about 100 $^\circ$C.

(b) PYROLYSIS or DEVOLATILISATION which starts at a temperature of about 180 $^\circ$C and continues up to much higher temperatures (i.e. 1000 $^\circ$C).

(c) GASIFICATION of residual char with reactive gases such as oxygen, steam, carbon-dioxide and hydrogen. The reaction with oxygen starts at about 400 $^\circ$C, while temperatures of about 700 $^\circ$C are necessary for the other reactions.

(d) SECONDARY REACTIONS of the pyrolysis and gasification products either in the gas phase or on contact materials (eg. bed material, fly-ash, charcoal, etc.)

In some types of gasifiers, as discussed previously, drying, pyrolysis, gasification and oxidation occur at distinct zones. In others there is a partial or almost complete overlap of these successive steps.

The individual processes of drying, pyrolysis, gasification and the secondary reactions will be discussed

in the following sub-sections.

A typical product distribution for gasification of wood or any other biomass fuel is represented in Figure 2.5. It can be seen that most of the gaseous products are initially produced in the pyrolysis step (about 65%). Therefore the pyrolysis reaction is important to determine the product gas composition. In reality, depending on the reactor configuration and reactor temperature, these pyrolysis gases will undergo secondary reactions. The final gas compotion can thus significantly differ from the primary pyrolysis gas composition.

FIGURE 2·5

Product distribution for gasification of wood or any other biomass
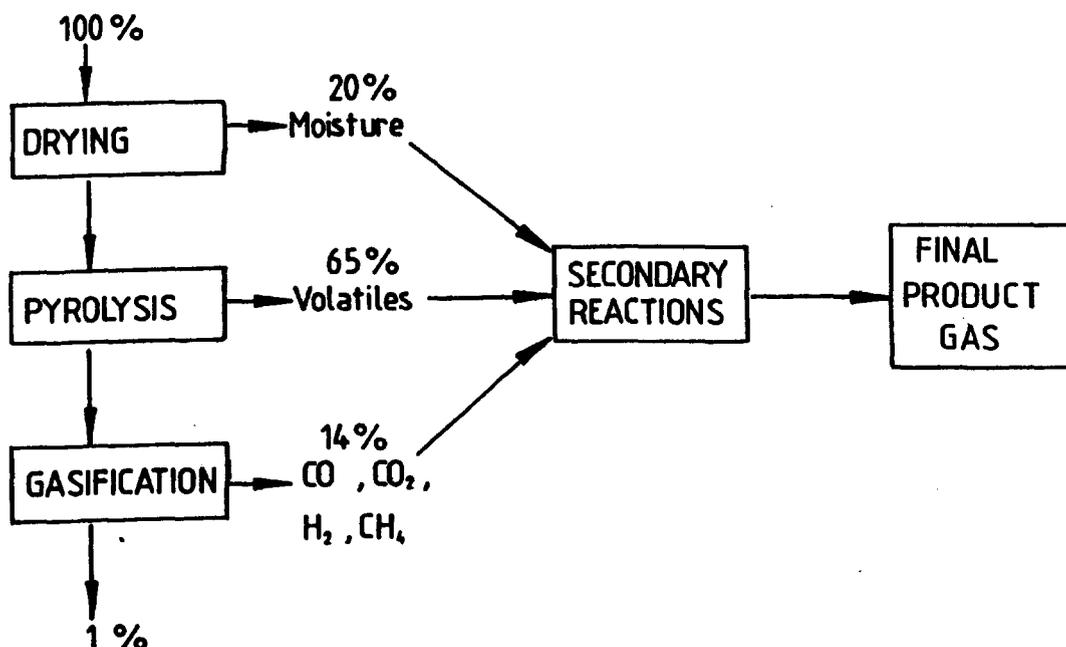
2.8.4.1 Drying

Drying is the first step involved in the gasification process. As the particle of biomass is heated to 100 °C, drying takes place by the evaporation of the "free water". Several types of biomass, such as green wood, bark or sludge, contain 50% or more moisture. For these materials drying is an important and capacity limiting phenomenon. The gasifier should be designed to provide a sufficient residence time for evaporation of the moisture. In some cases provisions should even be made for an additional heat input to accelerate the drying, for example preheating the blast.

In this work woodchips with a low (less than 15% by weight) moisture content were used. Moreover heating rate in a fluidised bed is extremely rapid due to the fast heat transfer throughout the bed and the large temperature difference between the fresh feedstock on the one hand, and the bed material on the other. It is therefore assumed that the drying process is instantaneous, and the kinetics of evaporation are therefore ignored

2.8.4.2 Pyrolysis or devolatilisation

For biomass the pyrolysis or devolatilisation is of paramount importance. Since the ratio of the volatile matter of the original dry mass is substantial (typically 80% or more for wood and other similar biomass materials).

The rate of pyrolysis is controlled by heat transfer and depends on the "volatile matter content" of the fuel. The volatile matter content of a few selected fuels is given in Table 2.1 (SCHOETERS, 1983). The volatile matter content of a fuel is determined according to a standard method and varies with the rate of heating, final temperature, particle size, sample preparation, etc. During pyrolysis gases and vapours are released which consist mainly of hydrocarbons, water vapour and other light gases such as carbon-monoxide, carbon-dioxide and hydrogen. This is accompanied by a large decrease in mass and volume. Both endothermic and exothermic reactions occur, and physical and thermodynamic properties vary continuously.

$$\text{Solid Fuel} \xrightarrow[\text{Drying}]{\text{Pyrolysis}} \text{Gas + Tar + } H_2O \text{ + Char} \qquad [2.6]$$
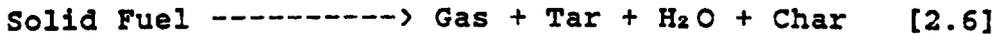
TABLE 2·1

Volatile matter content of selected solid fuels

FUEL	VOLATILE MATTER	FIXED CARBON	ASH
Lignite	43·0	46·6	10·4
Woods			
Douglas fir	86·2	13·7	0·1
White fir	84·4	15·1	0·5
Ponderosa pine	87·0	12·8	0·2
Western Hemlock	84·8	15·0	0·2
Cedar	77·0	21·0	2·0
Barks			
Douglas fir	70·6	27·2	2·2
Ponderosa pine	73·4	25·9	0·7
Cedar	86·7	13·1	0·2
Municipal refuse	65·9	9·1	25·0

Source : Schoeters (1983)

2.8.4.3 Gasification

After having passed through drying and pyrolysis, the residual hot char is further reacted, gasified, with mixtures cotaining oxygenand steam in a series of chemical reactions to yield combustible gases as outlined below :

(a) Gasification with oxygen or air (partial combustion).

$$C + 1/2O_2 \text{ ------------>} CO \qquad [2.7]$$
$$dH = -123.1 \text{ KJ/mol}$$
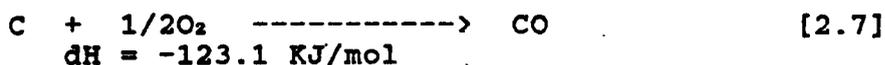
(b) Combustion with oxygen.

$$C + O_2 \text{ ------------>} CO_2 \qquad [2.8]$$
$$dH = -405.9 \text{ KJ/mol}$$
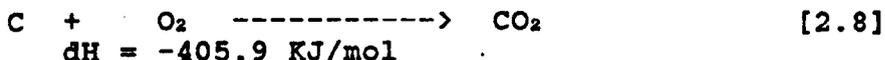
(c) Gasification with steam (water gas reaction)

$$C + H_2O \text{ <----------->} CO + H_2 \qquad [2.9]$$
$$dH = 118.9 \text{ KJ/mol}$$
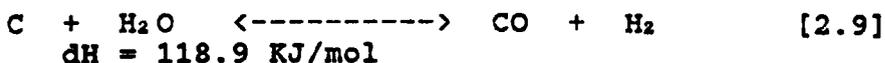
(d) Gasification with CO (Boudouard reaction)

$$C + CO_2 \text{ <---------->} 2CO \qquad [2.10]$$
$$dH = 159.7 \text{ KJ/mol}$$

(e) Gasification with hydrogen (hydrogasification)

$$C + 2H_2 \text{ <---------->} CH_4 \qquad [2.11]$$
$$dH = 87.4 \text{ KJ/mol}$$
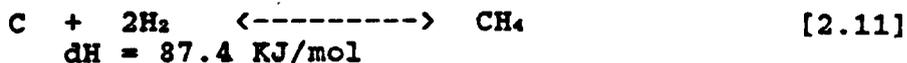
(f) Water-gas shift reaction

$$CO + H_2O \text{ <---------->} H_2 + CO_2 \qquad [2.12]$$
$$dH = -40.9 \text{ KJ/mol}$$

(g) Methanation

$$CO + 3H_2 \longleftrightarrow CH_4 + H_2O \qquad [2.13]$$
$$dH = -206.3 \ KJ/mol$$
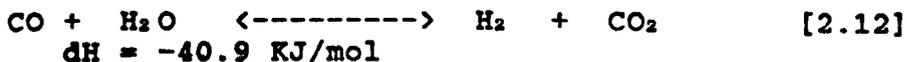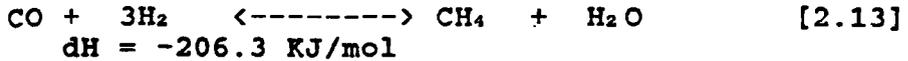
(dH at 298K, 1.013bar, H C = 12.5 KJ/mol)

In a gasifier operating autothermally, the exothermic oxidation reactions (a) and (b) which liberate heat are balanced by the endothermic gasification reactions (c) and (d). Energy is also consumed in the raising of the gasifying medium temperature to the reactor operating temperature and by thermal losses through the reactor walls. To achieve stable operating temperatures, steam addition to the gasifying medium is employed which also improves the quality of the gas produced through the formation of carbon-monoxide and hydrogen. The producer gas process utilises a medium of air or oxygen saturated with steam at 50 ^{o}C.

The reactions (a) to (g) give the stoichiometry and energy change of the reactions which take place during gasification, but provide no information on the completeness of the reaction or the way in which it proceeds. Therefore any discussion on the gasification chemistry must include the thermodynamics and kinetics of the processes invloved.

2.8.5 Thermodynamics Of Gasification Processes

The chemical reactions (a) to (g) all tend to a state of equilibrium. This means that after a sufficient reaction time, depending on the temperature and pressure, a fixed relation exists between the amounts of starting materials and the end products. The gasification reactions (c) to (g) can proceed from left to right or from right to left as indicated by the arrows. If the concentration of the gaseous species is expressed as its partial pressure, xp, in the gas mixture and the rate constants of the forward and reverse reactions are given by k_p and k_p' respectively. Then the equilibrium constants k_p for reactions (c) to (g) are given in Table 2.2.

The most important factor affecting the equilibrium gas composition is temperature. The derivation of the relationships which show this dependency are well documented in physical chemistry text books. The end product of the mathematical proof is the well known relationship of :

$$\log k_p = \frac{-dH}{2.303\ RT} + \text{Constant} \qquad [2.14]$$

A plot of log k_p against 1/T gives a straight line with the gradient of the slope equal to -dH/2.303R. This shows that the equilibrium constant will increase as temperature increases, which effectively pushes the equilibrium in

favour of reaction products for endothermic reactions, but in the direction of the reactants in the case of an exothermic reaction. The dependency can be evaluated numerically to give the equilibrium constant (k_P) at any temperature so long as the heat of reaction (dH) is known, or dH can be evaluated if k_P is known.

TABLE 2·2 : Equilibrium Constants
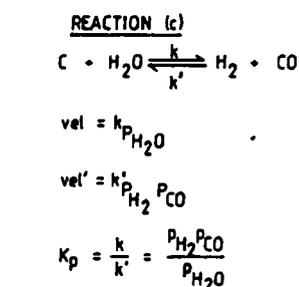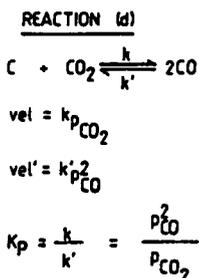
REACTION (c)

$$C + H_2O \xrightleftharpoons[k']{k} H_2 + CO$$

$$vel = k_{P_{H_2O}}$$

$$vel' = k'_{P_{H_2} P_{CO}}$$

$$K_p = \frac{k}{k'} = \frac{P_{H_2} P_{CO}}{P_{H_2O}}$$

In terms of mole fraction)

$$K_p = \frac{x_{H_2} x_{CO}}{x_{H_2O}} \cdot P$$

REACTION (d)

$$C + CO_2 \xrightleftharpoons[k']{k} 2CO$$

$$vel = k_{P_{CO_2}}$$

$$vel' = k'_{P_{CO}^2}$$

$$K_p = \frac{k}{k'} = \frac{P_{CO}^2}{P_{CO_2}}$$

In terms of mole fraction

$$K_p = \frac{x_{CO}^2}{x_{CO_2}} \cdot P$$

REACTION (e)

$$C + 2H_2 \xrightleftharpoons[k']{k} CH_4$$

$$vel = k_{P_{H_2}^2}$$

$$vel' = k'_{P_{CH_4}}$$

$$K_p = \frac{k}{k'} = \frac{P_{H_2}^2}{P_{CH_4}}$$

In terms of mole fraction

$$K_p = \frac{x_{CO}^2}{x_{CH_4}} \cdot P$$

REACTION (f)

$$CO + H_2O \xrightleftharpoons[k']{k} H_2 + CO_2$$

$$vel = k_{P_{CO} P_{H_2O}}$$

$$vel' = k'_{P_{H_2} P_{CO_2}}$$

$$K_p = \frac{k}{k'} = \frac{P_{H_2} P_{CO_2}}{P_{CO} P_{H_2O}}$$
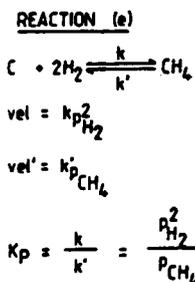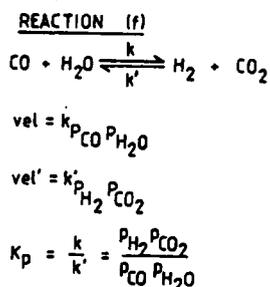
In terms of mole fraction

$$K_p = \frac{x_{H_2} x_{CO_2}}{x_{CO} x_{H_2O}} \cdot$$

REACTION (g)

$$CO + 3H_2 \xrightleftharpoons[k']{k} CH_4 + H_2O$$

$$vel = k_{P_{CO} P_{H_2}^3}$$

$$vel' = k'_{P_{CH_4} P_{H_2O}}$$

$$K_p = \frac{k}{k'} = \frac{P_{CH_4} P_{H_2O}}{P_{CO} P_{H_2}^3}$$
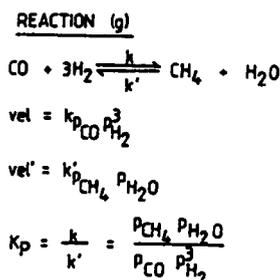
In terms of mole fraction

$$K_p = \frac{x_{CH_4} x_{H_2O}}{x_{CO} x_{H_2}^3} \cdot \frac{1}{P^2}$$

In terms of mole fraction, where

$$x_A = \frac{P_A}{P}$$

and where   P = Total Pressure

An indication of the possibility of a reaction to achieve equilibrium at any given temperature can be gained from the free energy change which has taken place, and is given by the following equation :

$$dH = -RT(\ln k_P)$$ [2.15]

A large negative value of dG indicates that at the given temperature the chemical position which will substantially favour the formation of the products and the time taken will be short. On the other hand, a small negative dG value indicates a slow approach to equilibrium and a final position which may favour reactant formation.

In a gasification system of carbon, steam, carbon-dioxide, carbon-monoxide, hydrogen and methane there are many reactions struggling to reach their own thermodynamic equilibrium positions. Some are reliant on other reactions for their reactant species, making a complicated chemical system where prediction of the equilibrium compositions through calculative procedures extremely difficult. Further complications are imposed by the physical nature of the gas-solid contacting which is prevalent in the fluidised bed reactor. There are a number of factors which contribute to the failure to achieve ideal equilibrium :

    (i)   The residence time of the gas in the bed in

            contact with the solid carbon, may not be

sufficient for equilibrium to be attained.

(ii) Resistances to diffusion of gaseous reactant

species to the carbon particle surfaces by

diffusing products and bulk gas.

(iii) The resistances to the mass transfer posed by gas

by passing through the bed in the form of

bubbles.

All of these factors combined mean that thermodynamic

equilibrium, as predicted from k_p data, will not prevail

and that only an approach to predictable, ideal equilibrium

compositions will be achieved. The degree of approach will

depend on the strength of the effects outlined above.

2.9 WOOD AS A FUEL

2.9.1 Available Wood Resource

The life cycles of plant depend on photosynthesis,

the process by which plant matter is produced from the

energy of the sun. According to CHEREMISINOFF (1980),

roughly 6% of the earths land is cultivated, however, only

about 1% of this agricultural photosynthesis is actually

utilised by man to produce food. The remainder represents

a vast renewable source of fuel or feedstock that could be

converted to fuel. Forests represent a major portion of this biomass resource. Despite drastic deforestation programmes in many regions of the world, all developing countries still have some wood resources remaining. In most countries they represent the largest single renewable energy resource. These wood resources can be divided between those which exist within the major forest areas and the smaller category of dispersed woodlands. Accurate estimates of total national wood resources are generally unavailable and those that do exist, nearly always refer to the major forest areas only. The many problems associated with defining and measuring forest areas and productivities are the primary cause of the lack of good resource data. These problems are further aggravated by the rapid rate at which conditions have been changing due to deforestation.

HALL et al. (1982) provided an estimate of the wood resources existing in the developing countries, by estimating the forest area remaining in selected countries on a per capita basis as shown in Table 2.3.

It must be remembered, however, that even though very large amounts of wood may be produced annually in some countries, not all of this is actually available in practical terms for energy purposes. Disregarding ownership considerations for the moment, the fact that many forest areas are very remote, tends to make the extraction of wood purely for energy purposes uneconomic at the

present.

Considering the wood resource in the developed
countries. ZERBE (1977) identified that in the United
States, 19Mt (oven-dry) of forest industry process
woodwaste and bark were being unused. In addition to this,
120Mt (od) is left at the logging sites. Further, he
identified, a new resource of 100Mt (od) of
'non-commercial' timber. This consisted of rough and
rotten wood and wood of insatisfactory species, which was
largely found in the unmanaged timber stand of the United
States. Therefore the available wood resource in the
United States is equivalent to 775.60 Mtce/yr (od).

TABLE 2·3

Wood resources in selected developing countries

COUNTRY	POPULATION (10⁶)	FOREST REMAINING (10⁶ hectares)	FOREST PER CAPITA (hectares)	SUSTAINABLE WOOD YIELD (Mtce/yr)
Zaire	1	17	12·1	19·43
Brazil	116	350	3·0	11·91
Argentina	27	60	2·2	4·19
Burma	35	38	1·1	3·75
Tanzania	18	30	1·7	3·00
Thailand	48	20	0·4	1·26
Nepal	14	4	0·3	0·72
Mali	7	6	0·9	0·40
Upper Volta	7	4	0·5	0·34
India	694	75	0·1	0·24
Kenya	16	1	0·06	0·07
Bangladesh	89	3	0·03	0·07

Canada, like the United States, has a very large timber resource, much of which is considered to under utilised. LOVE and OVEREND (1978) evaluated that over 32% of the land area in Canada is forest, and with maximum productivity would yield 400 Mt/yr (od), with an energy content of 273.39 Mtce/yr (8 TJ). This is equivalent to the total Canadian basic energy demand in 1974.

In the EC woody forest products are used primarily for construction timber, paper pulp or composites. However, about 8% of the available woody resource is used directly for fuel (PALZ and CHATIER 1980). Wood processing wastes are also used to some extent as fuel sources within this industry. Despite this usage, STRUB (1985), evaluated the net energy potential (after conversion) of wood resources to be as follows :

	NET	TW ($x\ 10^{-3}$)
(1) Forestry/wood wastes	11.30 Mtce	11.3
(2) Firewood	9.44 Mtce	9.4
(3) Energy Forestry & Crops	52.34 Mtce(a)	52.2
TOTAL	73.08 Mtce/yr	72.9 (b)

(a) Potential available in 15 years
(b) 4.2% of the energy consumption of the EC in 1985.

In the UK 0.4 Mtce of wood waste is available as a fuel source, according to a report by the DEPARTMENT OF ENERGY (1982). 1 Mtce/yr of logs, thinnings and forestry

waste are also available. In their report they reached a conclusion suggesting that despite the limited land area of the UK, if the current pattern of land use for agriculture remained unchanged, about 6% of the total land area will be available to grow wood for conversion to biofuels or direct combustion. This would represent approximately 8 Mtce of wood could be available each year.

Total Wood Resource Available

		Net
(1)	America	775.60 Mtce/yr
(2)	Canada	272.39 Mtce/yr
(3)	EEC	73.08 Mtce/yr
(4)	Developing Countries	45.36 Mtce/yr
	TOTAL	1166.43 Mtce/yr

The possible wood resource available for utilisation is approximately 1166.43 Mtce/yr. which represents 12% of the total world energy usage in 1980. This woody biomass resource available could make a contribution in the energy supply market, even though it will not replace one of the major fossil fuel supplies (Oil 26.5%, Coal 52.5%, Natural Gas 19.0% of the total world energy usage in 1980). Nevertheless it can make a significant contribution in supplementing the fossil fuel supply and slowing down their rate of depletion.

2.9.2 Wood Chemistry And Utilisation

In characterising and correlating reactivity data for pyrolysis and gasification, it is necessary to have some idea of the chemical structure of the reactant material. Wood is a very heterogeneous compound, varying by tree species and wood material, for example, bark, heart-wood, sapwood, needles, etc. It has, therefore, got the same complexities in classification as coal. In determining the energy value of wood materials, the physical structure, for example cell walls, are considered to be less important than the chemical composition. Thus the chemistry of the wood materials determines its combustion and conversion characteristics.

Wood materials can be separated into three fractions, according to GRABOSKI and BAIN (1979), extractables, cell wall components and ash. The extractables, generally representing 4% to 20% of wood, consist of materials derived from the living call. The cell wall components, representing the bulk of the wood, are principally the lignin fraction and the total carbohydrate fraction (cellulose and hemicellulose), termed holocellulose. Lignin, the cementing agent for the cellulose fibres, is a complex polymer of phenylpropane.

Cellulose is a polymer formed from d(t)-glucose. While the hemicellulose polymer is based on the hexose and pentose sugars. In woods, the cell wall fraction generally consists of lignin/cellulose in the ratio 43/57. Residues of the total wood, such as bark and sawdust, have differing compositions.

TILLMAN (1977), measured the heating value of different tree species, the amounts of cellulose, hemicellulose and lignin content, as given in Table 2.4. From the data he obtained the following general formula for determining the approximate heating value of the wood on a dry weight, ash-free basis :

$$HHV = 17.51C + 26.70(1-C) \qquad [2.17]$$
where HHV = higher heating value in KJ/KG
C = fraction of wood consisting of holocellulose

The composition of the wood therefore determines both its heat content and how it releases useful energy.

Wood combustion and conversion processes begin with pyrolysis, as first chemical process. SHAFIZADEH (1977) stated that during pyrolysis of wood, holocellulose principally promotes the release of volatiles, with the production of an intermediate tar fraction containing levoglucosan in the initial depolymerisation. While lignin, which also releases volatiles, primarily promotes the formation of char. The volatile pyrolysis products further burn in the gas phase with flaming combustion.

TABLE 2·4

Different tree species

TREE SPECIES	% CELLULOSE	% LIGNIN	% HEMI-CELLULOSE	HEATING VALUE MJ/K(
Beech	45·2	22·1	32·7	19·67
White Birch	44·5	18·9	36·6	19·38
Red Maple	44·8	24·0	31·2	19·54
Eastern White Cedar	48·9	30·7	20·4	19·54
Eastern Hemlock	45·2	32·5	22·3	20·67
Jack Pine	45·0	28·6	26·4	20·77
White Spruce	48·5	27·1	21·4	20·68

While the charred residue burns at a relatively slower rate by surface oxidation or glowing combustion.   This is shown schematically in Figure 2.6.

FIGURE 2·6

COMPETING REACTIONS IN THE PYROLYSIS AND COMBUSTION OF CELLULOSE

PYROLYSIS	COMBUSTION
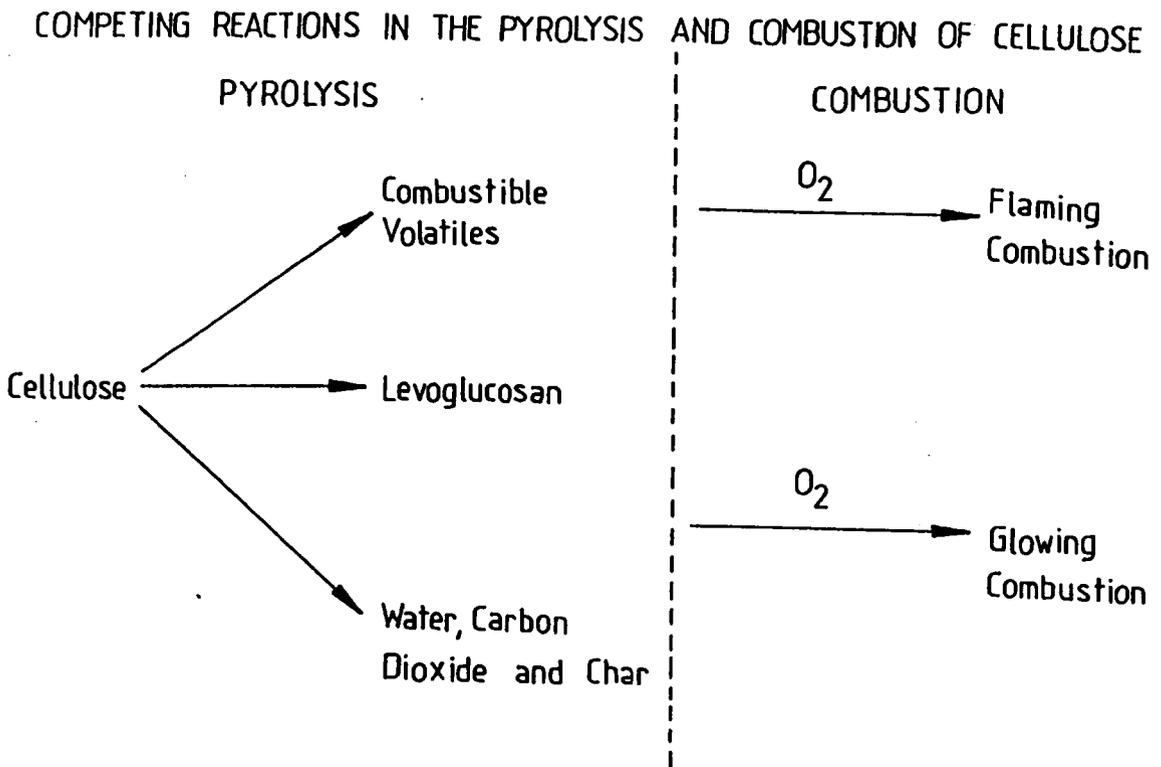
Pyrolysis or thermal degradation of cellulose and hemicellulose to flammable volatile products and chars involves a series of highly complex reactions and a variety of intermediate products, which have been extensively investigated by many researchers (SHAFIZADEH 1982, BRADBURY et al. 1979, SUSOTT et al. 1975, THURNER and MANN 1981). If we start with the basic concept of wood pyrolysing into three components, gas, tar and char, as stated by SHAFIZADEH and CHIN (1977), shown schematically below in Figure 2.7.

FIGURE 2.7

Mechanism of wood pyrolysis suggested

by SHAFIZADEH and CHINN (1977).
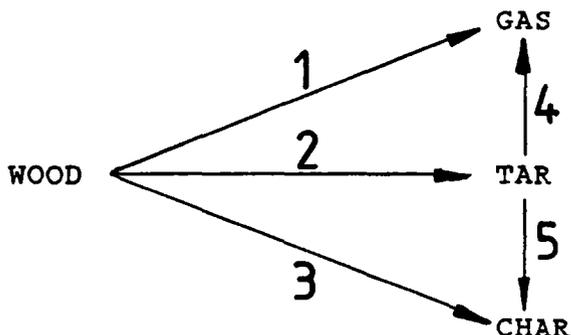
The reactions 1,2 and 3 are called primary
reactions since the tar decomposes into gas and char
according to the parallel secondary reactions 4 and 5.
Each product in above Figure represents a sum of the
numerous components which are lumped together to simplify
the analysis. The composition of each product, especially
the distribution between the gas and the tar, depends,
among other things, on the conditions under which the
products are collected. In principle, the reaction rate
constants of these five reactions can be determined by
measuring the amount of each product as a function of time.
When the tar is removed from the reaction zone the
secondary reactions are avoided and the reaction constants
of the primary reactions can be determined directly from
these measurements. Detailed descriptions of the pyrolysis
products and reactions are beyond the scope of this work.

Wood gasification is a major modification of the fundamental pyrolysis process. As with pyrolysis, initially the product streams are gaseous volatiles and char. However, in gasification, the reactions are carried out at sufficiently high temperatures to limit the production of condensible tars and oils. Further, some of the char is partially oxidised to provide the heat to drive the endothermic gasification reactions. The remaining char is reacted with water (steam) to produce additional gaseous compounds.

Wood combustion and conversion processes are given in Figure 2.8. The effects of the moisture on the wood pyrolysis, which is the initial stage in both the combustion and conversion processes, influences the relative properties of the char and volatiles produced, and reduces the rate of pyrolysis. This can be explained by the fact that solid-phase reactions, of heating, drying and pyrolysis, are influenced by the fuel composition and the fuel core's internal heating rate. Moisture affects these reactions by increasing the thermal conductivity, decreasing the flame temperature and increasing the heat capacity of the fuel particle. Consequently, reducing the heating rate to a point where char formation is favoured. This in turn reduces the formation of volatiles, the rate of pyrolysis and thus the rate of combustion or conversion.

FIGURE 2·8
WOOD ENERGY RECOVERY SYSTEMS
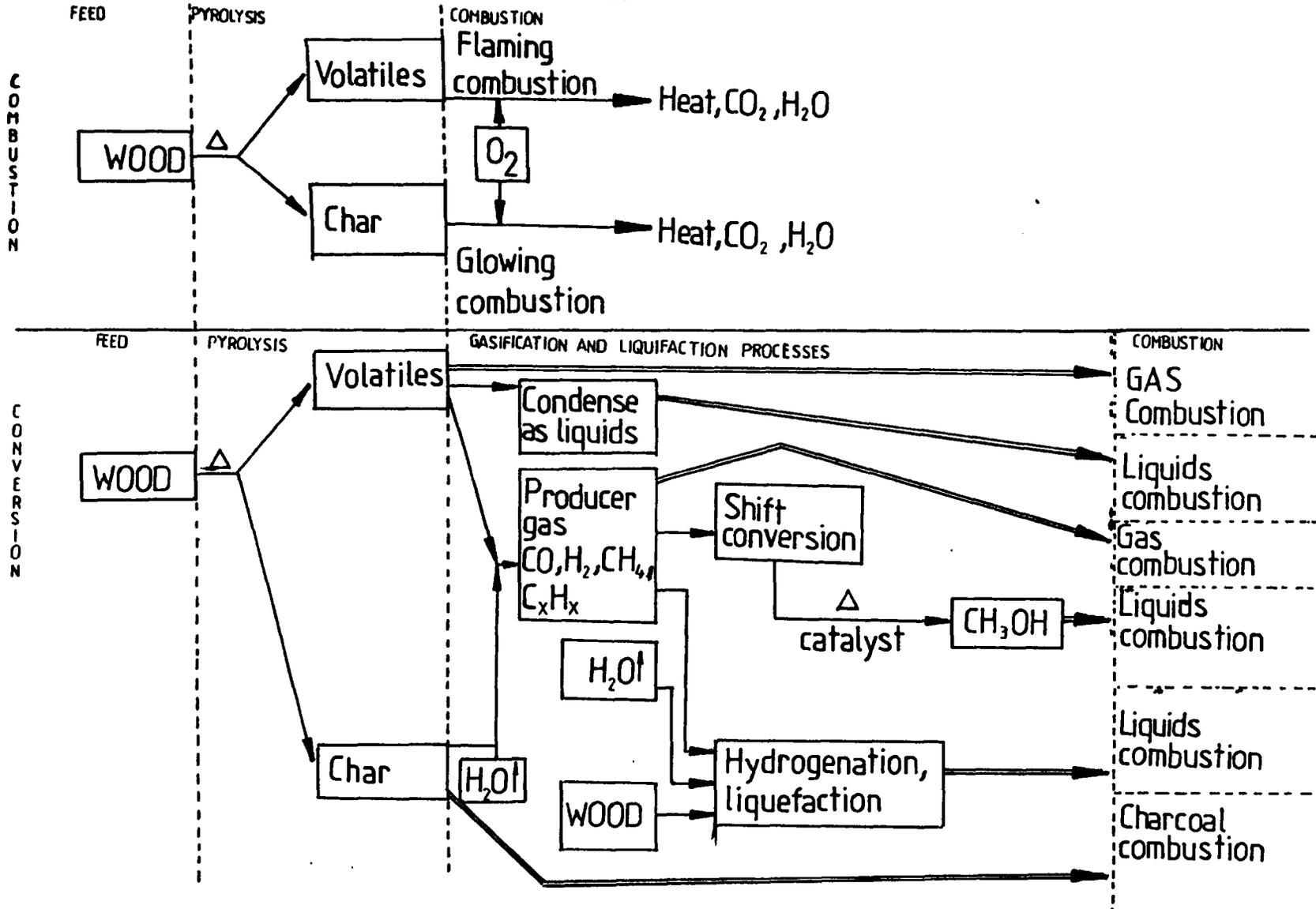
2.9.3 Physical Properties Of Wood

In addition to the chemical properties of wood, the other major physical data necessary for predicting the thermal response under pyrolysis, gasification and combustion processes are thermal conductivity, heat capacity, true density and diffusion coefficients.

2.9.3.1 Thermal Conductivity

The thermal conductivity is defined in general terms as a proportionality factor which relates heat flow through a material to a temperature difference across a specified distance in that material. Mathematically, thermal conductivity is defined by Fourier's Law of heat conduction, given here for unidimensional heat flux in the x-direction in the regular coordinates :

$$q = - k \frac{dt}{dx} \qquad [2.18]$$

For wood, the thermal conductivity is a function of temperature and spatial direction, and also the major constituents, including moisture, cellulose, hemicellulose and lignin.

2.9.3.2 Heat Capacity

Heat capacity is normally defined in terms of the enthalpy content of a material and represents the relative

ability of a material to store enthalpy. Enthalpy is a
function of temperature and pressure :

$$H = H(T,P) \qquad [2.19]$$

and

$$dH = (dH/dT)_P dT + (dH/dP)_T dP \qquad [2.20]$$

For solids and liquids (dH/dP) is very small and therefore

$$dH = (dH/dT)_P dT \qquad [2.21]$$

where (dH/dT) is called heat capacity at constant pressure,
Cp. Several Cp equations have been developed to predict
the heat capacities of wood at temperatures up to 100 °C
BEALL (1968) presented the following moisture dependent
equation :

$$Cp = 0.259 + (9.75 \times 10^{-3})M + (6.05 \times 10^{-6})T \qquad [2.22]$$
$$+ (1.30 \times 10^{-9})MT$$

where M = % moisture content (up to 27%)
       T = Temperature (°C)

Wood typically has a heat capacity of 0.45 - 0.67 for
20 - 40 °C.

2.9.3.3 Density

The density of the material is important when
considering energy contents of fuels on a volumetric basis,
such as for transporting, solids handling, etc. There are
three ways of reporting solid material density; bulk

density; apparent particle density and skeletal density.
The density values differ in the way in which the material
volume is calculated. The bulk density volume basis
includes the actual volume of the solid, the pore volume,
and the void volume between solid particles. Apparent
particle density includes solid volume and pore volume.
Skeletal density, or true density, includes only solid
volume.

2.9.4 Beneficiation Of Wood Resources

Beneficiation is defined as the treatment of some
parent material, in this case wood, so as to improve the
physical and/or chemical properties of the material. The
main aim of beneficiation processes for wood products is to
produce from the parent biomass, a material that is a
better quality feedstock for gasification, pyrolysis or
combustion, that has a higher volume energy density or
higher specific surface area, and that has a higher gross
heating values. There are three main beneficiation
processes for wood resources.

2.9.4.1 Comminution

Commminution is a process in which the particle
size of a parent material is reduced to a desired range
either by shredding, cutting, grinding or pulverisation.

DRONFIELD et al. (1978) identified fiberization as the process of comminution used for wood products. The main processes used are, compression and cutting. Although impaction and attrition are undoubtedly important in high speed cutting operations.

2.9.4.2 Drying

In the drying process physically bound water is driven off (the removal of chemically bound water is not included). The removal of water reduces the weight of the material that must be transported or handled in the processing plant. Thereby lowering the operating costs. In addition, the removal of water generally produces a feedstock of better quality. In any processing step, equipment and operating costs must be considered in evaluating the usefulness of the process.

In wood, drying normally means the removal of water from the solid to reduce the moisture content to an acceptablly low value. This is usually accomplished by thermal drying, as opposed to mechanical drying. Since most fresh woods contain a considerable amount of water (40% - 60% by weight), a significant saving can be made if waste heat from another processing step, such as the combustion step, is used to supply the thermal energy needed for drying.

2.9.4.3 Densification

In densification, the apparent particle and the bulk density of the material are increased so as to lower transportation costs or processing equipment size by reducing the volume of the material handled.

An in depth study of densification has been conducted by REED and BRYANT (1978). They presented five methods of densification for biomass in commercial operation and several other processes in the development stage. The five processes are characterised by the size of the densified feedstock required. The densification process takes advantage of the physical properties of two of the major components of wood or any other biomass material, ie. cellulose and lignin. Cellulose is stable to 250 ^{o}C while lignin begins to soften at temperatures as low as 100 ^{o}C. The densification process is carried out at temperatures that ensure that the cellulose material remains stable but that soften the lignin fraction, making it act as a 'self bonding' agent that gives the densified fuel its mechanical strength. Water content must be controlled in the range from 10% to 25% to maximise pressure requirements for densification.

2.9.5 Wood Gasification Processes

2.9.5.1 Wood Gasification processes In Operation

The processing of wood for merely fuel gas production is still in the development stage, but there are some commercial processes under development. Details of some of the better documented processes in literature are presented in Tables 2.5 and 2.6, and it would serve well the purpose of this chapter if a brief description of the processes were included.

The Solar Energy Research Institute's (SERI) 28 KG/hr, down-draught oxygen gasifier has been now tested for several years. In this system about a 1/2 to 2/3 of the total oxygen is fed through a diffusor into the freeboard zone at the top of the bed. The rest is fed through three groups of 4 oxygen lances down-stream in the reactor. Tars from the flash pyrolysis of the incoming particles and back diffusing tars from the bed are burnt above the bed. This provides heat for the rapid pyrolysis of the fresh particles falling into the reactor. The temperature in the pyrolysis oxidation zone is 900-1000 oC. Down-stream, the temperature profile will depend on the traction of oxygen fed through the lower lances. If that fraction is low, the temperature will continuously decrease because of the dominating effect of the endothermal reactions. If the

TABLE 2·5

Wood gasification systems in operation

PROCESS TYPE/NAME	GASIFYING MEDIUM	REACTOR TYPE	SCALE	MAJOR PRODUCT	REFERENCE
[1] SERI	Oxygen	Down-draft	28 Kg/hr	Gas	REED (1981)
[2] SFW-FUNK	Air/oxygen	Up-draft (pressurised)	520 Kg/hr	Gas	HUMMELSIEP (1981)
[3] CREUSOT-LOIRE	Oxygen	Fluidised bed	100 Kg/hr	Gas	GEHRMAN (1981)
[4] MINO	Oxygen/steam	Fluidised bed (pressurised)	10 Kg/hr	Gas	CHRYSOSTOME & LEMASLE (1982)
[5] LURGI	Oxygen/steam	Circulating fast fluidised bed	—	Gas	LINDNER & LINAKI (1982)
[6] BATTELLE COLUMBUS	Steam	Double fluidised bed	100 Kg/hr	Gas	FELDMAN (1981)

TABLE 2·6

	PROCESS TYPE	WOOD MOISTURE	CALORIFIC VALUE	VOLUME %						REACTOR TEMPERATURE
				H_2	CO	CH_4	C_nH_m	CO_2	N_2	
[1]	Gasification/ pyrolysis	10% wt.	Medium CV	28·7	43·3	2·4	—	12·7	2·4	900–1000 °C
[2]	Gasification	—	Medium CV	27·0	32·0	5·0	—	21·0	15·0	∿ 1000 °C
[3]	Gasification	15–40% wt.	Medium CV	18·7	35·5	10·0	—	30·3	1·8	850 °C
[4]	Gasification	10 % wt.	Medium CV	32·5	20·0	17·0	—	30·2	0·3	750 °C
[5]	Gasification	20 % wt.	Medium CV	37·8	36·0	0·9	—	24·9	0·5	900–1000 °C
[6]	Gasification	—	Medium CV	13·3	49·2	14·0	8·4	15·7	—	820 °C

fraction is large, there could be a possibility of temperature increasing as solid conversion/consumption is increased. SERI has been successful in producing a low tar (less than 300 mg/Nm^3), low methane synthesis gas from pellets and chips, a charcteristic diameter of 0.6cm. Since the pyrolysis time roughly increases with the square of the particle diameter, there probably will be an upper limit for the wood chip dimension if the synthesis gas must be low in methanated tars. This is because once the pyrolysis process is not completed in the top section of the gasifier, the methane content in the gas may sharply increase due to insufficient conversion of pyrolysis methane evolved in the relatively colder reduction zone (700-900 $^\circ$C). For this reason scale-up of the SERI process could provide some problems.

The SFW-Funk up-draft gasifier was developed originally to utilise municipal solid wastes. It was later used to gasify 520 KG wood/hr. The only problem of this process is that the tar and methane content in the synthesis gas are large. This was solved by passing the fuel gas through a wet scrubber and an electrofilter, after which the dust and tar content were reported to be under 10 mg/Nm^3.

The atmospheric fluidised bed was employed in the Creusot-Loire process to gasify wood. The process, initially developed by Winkler for coal gasification, is

relatively simple and flexible, which, if operated at
800-1000 ^{o}C., produces a low tar but still methane rich
synthesis gas. The process is basically autothermal as the
heat necessary for the gasification reactions is provided
by combusting part of the wood. As a consequence, the
carbon-dioxide content of the product gas is relatively
high.

The Mino process is a two staged gasification
process, where the first stage is a pressurised, 1-3 Mpa
fluidised bed in which the fuel is pyrolysed and gasified
with oxygen and steam at 700-900 ^{o}C. The gas leaving
the gasifier, after passing through a cyclone, enters a
high temperature filter where the last traces of dust are
removed. This is necessary to avoid plugging in the
secondary stage, where additional oxygen is added. The
temperature rises and tar and methane are catalytically
converted to carbon-monoxide and hydrogen. The main
advantages of this process are that ; low oxygen
consumption as the catalysts make the conversion possible
at lower temperature ; tar conversion over the catalyst is
high and pressurised operation reduces power consumption
and the volume of the vessels required.

The circulating fluidised bed was originally
developed by Lurgi for calcination and coal combustion,
application to wood gasification, is however, new.
Compared with a classical fluidised bed this reactor is

operated at much higher superficial gas velocities, although these are still lower than those realised in an entrained flow reactor (with particles of similar size and specific density). In this flow regime slip velocities between particles and gas are relatively high and as a result it would be possible to prevent ash sintering up to higher temperatures than in classical fluidised beds. Lurgi is presently investigating operation up to 1100 $^{\circ}$C, and if this turns out to be successful, it will most probably produce a virtually tar free gas in one step.

The double fluidised bed was initially developed for the catalytic cracking of mineral oils. A modified version of this has been developed by Battelle Columbus for steam gasification of wood. In which wood is fed to a steam fluidised bed, where pyrolysis and endothermic steam gasification takes place at temperatures of 800-850 $^{\circ}$C. An inert fluidising carrier (sand or alumina) and unconverted char are transported to a second fluidised bed where this char is exothermically oxidised at temperatures of 900-1000 $^{\circ}$C. The return flow of the hot carrier to the gasifier provides the heat for the endothermic gasification processes. This has been strongly advocated for several years by BAILIE (1981), BOWLING and WATERS (1968), and many others. The main advantages of this process are that the capital costs are relatively low since no oxygen plant is required, low operating expenses because

no oxygen is used and the product gas has low carbon-dioxide content since combustion and gasification processes are separated. The disadvantage might be the complexity required for the circulation of the inert solids between the two fluidised beds.

2.9.5.2 Hybrid Wood Gasification systems

The hybrid wood gasification systems, as shown in Table 2.7 are defined as process routes where something else is continuously fed into the reactor system, in addition to wood and the gasifying medium, for example alternative fuels (coal, hydrogen, methane, etc.) or alternative heat source (electricity or solar energy for example). The aim of such processes is always the same, to produce more synthesis gas per unit mass of wood or any other biomass fuel. All these processes are still in their early development or conceptual stages and therefore no further discussions are made here.

TABLE 2·7

Hybrid Medium CV Wood Gasification Routes

PROCESS	FEATURE	STATUS	REFERENCE
1) Flash pyrolysis and steam gasification in solar furnace (solar heat hybrid route).	Heating rate ≃1000 W/m Temp. rise =140000 °C/s t less than 1sec dp less than 1mm.	Demostrated at laboratory scale only.	ANTAL (1980)
2) Hyrogen hybrid route.	Oxygen from electrolysis of water; hydrogen from water added to the product gas.	Concept; probably nowhere competitive.	NITSCHKE (1981) MARSHALL (1979)
3) Electricity hybrid route.	Increased methanol wood ratio by adding electricity.	Very site specific but integral pilot plant exists.	NOVELERG (1982) BRECHERET (1981, 1982)
4) Methane hybrid route.	Methane reformed and added to balance hydrogen deficient product gas.	Site specific but relatively attractive; planned.	ROCK (1982) MARSHALL (1979)

2.9.6 Future Role Of Wood Gas

Fuel gas production from wood and its use is more
cumbersome and inconvenient than the use of oil and gas,
and arguably even than coal. The main attraction will be
the saving in the costs over conventional fuels. Although
there may be circumstances where fuel gases offer a
distinct advantage, such as in applications needing a low
sulphur content gas. In the future the role of wood fuel
gases will probably be, direct use via neighbourhood grid
system, in power generation, as fuel for use in boilers and
as feedstock for methanol production. These future roles
of wood gas will be discussed briefly in the following
sub-sections.

2.9.6.1 Fuel Gas

Medium CV gas produced from wood gasification
supplied in private pipelines for nieghbourhood use needs
only a modest increases in oil and natural gas for it to
become attractive to energy users. For the production of
methane from wood to become attractive to suppliers of gas
via the grid, the production costs of the pipeline gas will
need to rise quite significantly. This might be very
possible in the future, since as the stocks of natural gas
are depleted and costly to exploit, and when SNG is
produced from coal, the price of methane to the energy user
will be significantly increased. Low CV gas from wood at

local systems is already attractive in economic terms, and
the deployment problems are likely to be only relatively
minor. With the recent interest in alternative fuels these
problems will be ironed out in the not too distant future.

2.9.6.2 Power Generation

Use of wood gas in engines for electricity
generation is probably its most demanding application.
There is no saving in capital through fule gas, only
savings in diesel fuel costs. Since generation of
electricity for the grid is highly efficient in
industrialised countries, it is going to be difficult for
power generation from fuel gas to compete. In the UK,
generation at 1000KW scale is barely economical even if the
gas is free, as is the case of fuel gas from sewage sludge
digesters. Power generation using wood gas is only likely
to be attractive in isolated communities where there is no
grid electricity, and where diesel fuel is expensive and
wood is cheap.

2.9.6.3 Boiler Fuel

Wood gas will burn quite satisfactorily in
boilers,and in principle can replace oil and natural gas in
existing equipment. Medium CV gas can be used as a direct
substitute for natural gas, but use of low CV gas
containing less than 7 MJ/Nm^3 leads to derating of the

boiler according to BAILIE (1980). The extent of derating can be slightly compensated if the sensible heat generated in the gasification step is utilised. HODAM et al. (1982) have concluded in their report that for small boilers (less than 10 GJ/hr), the use of fuel gas from gasification of wood might have an edge, and on large scale operations it has a real advantage.

2.9.6.4 Methanol Production

The production of methanol by further processing of medium CV gas is another possibility for the use of wood gas. Methanol can be used as a supplement to transport fuels and its production costs from wood at the present is quite close to its selling price. Since the price of gasoline is greater than methanol, it is economically viable at the moment to blend 3% methanol, from wood, with gasoline. This can be achieved according to ADER et al. (1982) without any changes to either the distribution system or any other modification to the conventional engines.

2.10 SUMMARY OF THE LITERATURE SURVEY

Despite the uncertainty over future energy prices, wood with a capable resource of 1100 Mtce/yr, representing 12% of the total world energy usage in 1980, will have an important role to play in the future energy scene as part of the biomass utilisation strategy. This resource will be especially significant since the reserves of the traditional convenient fuels, such as oil and natural gas are constantly being consumed at an alarming rate.

The survey of literature presented in this chapter serves to highlight the number of different ways in which biomass can be exploited, and the various forms of biomass feedstocks available for utilisation. The only problems associated with a wide spread usage of biomass are that whether it is technologically feasible and that whether it is economically viable. At present, practical biomass prospects which are economically viable are relatively small in scale and localised in application, though large scale energy plantation type schemes of various sorts are envisaged. If biomass is to be exploited in quantity in this way, than a number of broad questions will arise. These concern the overall picture of world energy consumption, changing patterns of usage arising from increase in the cost of 'conventional' fuels and changing technology, doubts about the continuity of supply of the

previously dominant fuels, and their environmental effects.
Several paths of alternative energy supply are being
explored and, in parts adopted, of which biomass energy is
but one. Though the way things will go, will not become
apparent for a number of years, major changes in energy
supply and use are inevitable. Even the nuclear solution
to supply a large part of the world electrical energy needs
economically, and at a socially acceptable cost, cannot be
a complete answer as electricity is not a flexible
transport fuel. The need for gas, liquid and even solid
fuels will remain. Due to the diversity of potential
solutions, each having merit in a particular context of
energy supply or exploitation of a resource. The most
probable scenario will be one of great diversity including
a high degree of local variation in approach, and one in
which biomass will play a significant role, if not a
dominant role.

Wood in the form of wood residues or wastes has a
significant potential as a feedstock for conversion
technologies. The thermochemical processing of wood in
fluidised beds is only in its development stage, even
though some commercial plants are available, they are in a
relatively small scale. The potential of fluidised beds
in conversion technologies is already becoming clearer, and
with this improved perspective comes better defined
research and development requirements. Although many

problems have been solved, solutions often create new problems and thus research and developments are increasing rather than decreasing.

The First European Workshop on Thermochemical Processing of Biomass, which was held at the University of Aston in Birmingham, April 1983, concluded by saying, quote:

> *"The interest in biomass as a renewable energy and feedstock resource continues unabated. While R & D programmes world wide are now culminating in commercial systems, this should not be seen as the end of the activity in this area, but rather a justification of both the work to date and particularly the continuing need for a sound R & D programme in the future"*

CHAPTER THREE
EXPERIMENTAL

3 EXPERIMENTAL

3.1 INTRODUCTION

The work presented in this chapter is concerned with an experimental investigation into the production of a low CV fuel gas from woodchips using a fluidised bed reactor. A fluisdised bed was chosen because it offers a greater adaptiblity to handle various types and sizes of fuel feedstocks. Additionally it was the intention of this study to show that improvements to the basic fluidised bed dynamics are possible. This inspired the initiation of the the cold modelling study discussed in the next chapter.

The gasification process was one of partial combustion and pyrolysis, since air was used as the gasification medium and heat transfer within the bed was aided by hot sand particles. This process has two major advantages in that it avoids losses associated with heat transfer, and the design and construction of the equipment is simpler. The use of air without expensive oxygen or steam additions means that operating costs are kept to a minimum which is particularly important in commercial exploitation of low grade fuels. It is, therefore,

important to gain knowledge of the expected gas quality and quantity using small scale apparatus together with the identification of any problems which may be encountered.

3.2 EXPERIMENTAL EQUIPMENT

This section of the chapter is concerned with the description of the various component parts which comprised the experimental fluidised bed unit. The operation of the experimental rig is descibed in detail along with other techniques employed to gather experimental data. Schematic diagram of the equipment is shown in Figure 3.1. The apparatus is also shown, in its operational form in Plate 1.

The gasifier apparatus consisted of a fluidised bed reactor with freeboard and disengaging sections. Built around the central unit were the fuel feeding systems, the preheater section, for start-up and the cyclone for dust collection. The electrical and gas (air and nitrogen) supplies to the rig were controlled from a wall mounted panel, as shown in Plate 2. For convenience and safety all the piping and wiring was contained in an overhead supply conduit. Product gas from the rig was vented to the

FIGURE 3·1 (a)

DIAGRAM OF THE FLUIDISED BED REACTOR RIG
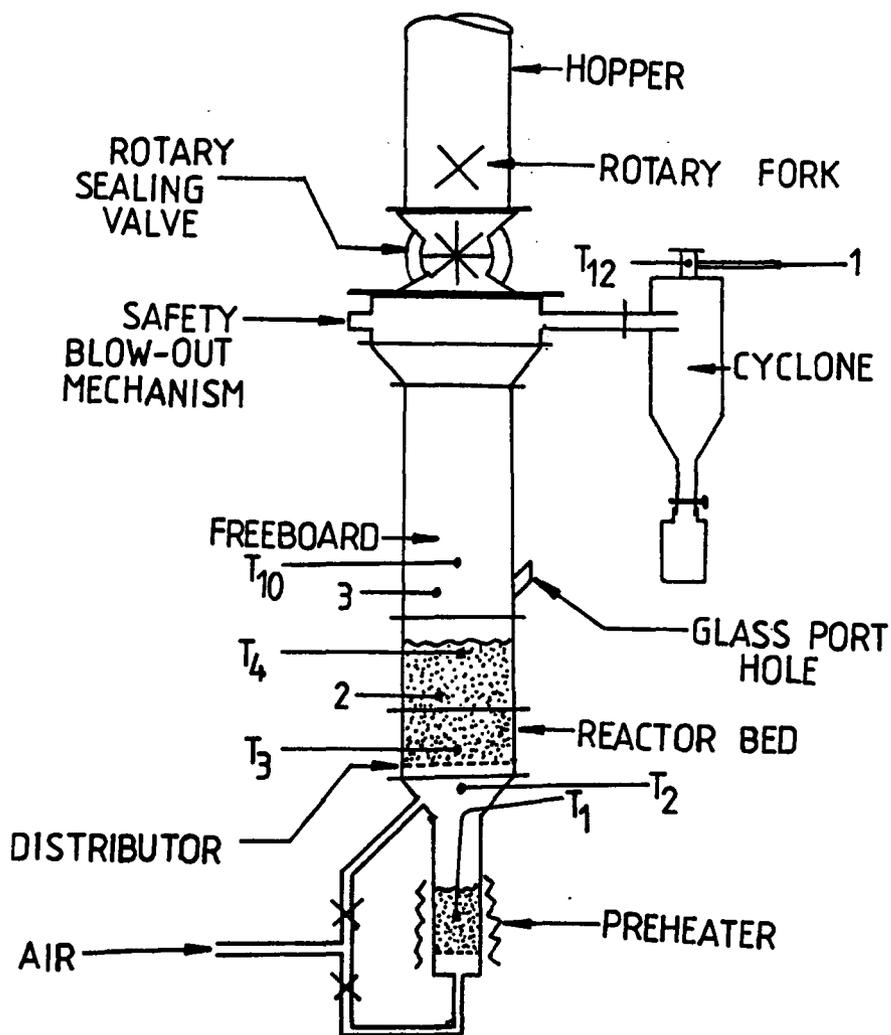
T_1	PREHEAT BED TEMPERATURE
T_2	PREHEAT GAS TEMPERATURE
T_3	BOTTOM OF BED TEMPERATURE
T_4	TOP OF BED TEMPERATURE
T_{10}	FREEBOARD TEMPERATURE
T_{12}	EXIT GAS TEMPERATURE

1	EXIT GAS SAMPLE PORT
2	BED GAS SAMPLE PORT
3	FREEBOARD GAS SAMPLE PORT

PLATE 1

Fluidised Bed Reactor Rig In Its Operational Form

ROTARY SEALING VALVE

SAFETY BLOW OUT MECHANISM

CYCLONE

GLASS PORT HOLE

FREEBOARD

REACTOR BED
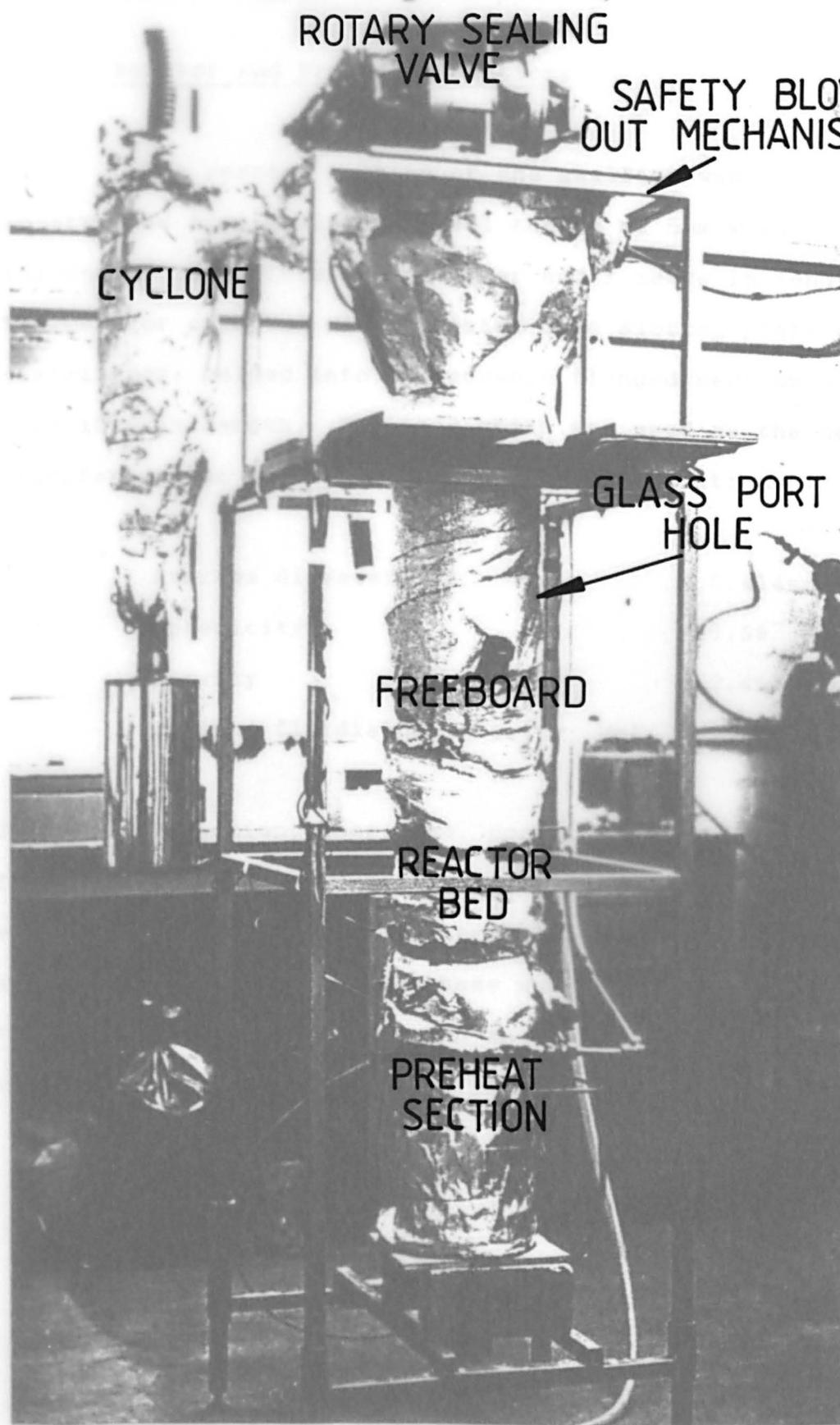
PREHEAT SECTION

PLATE 1

Fluidised Bed Reactor Rig In Its Operational Form
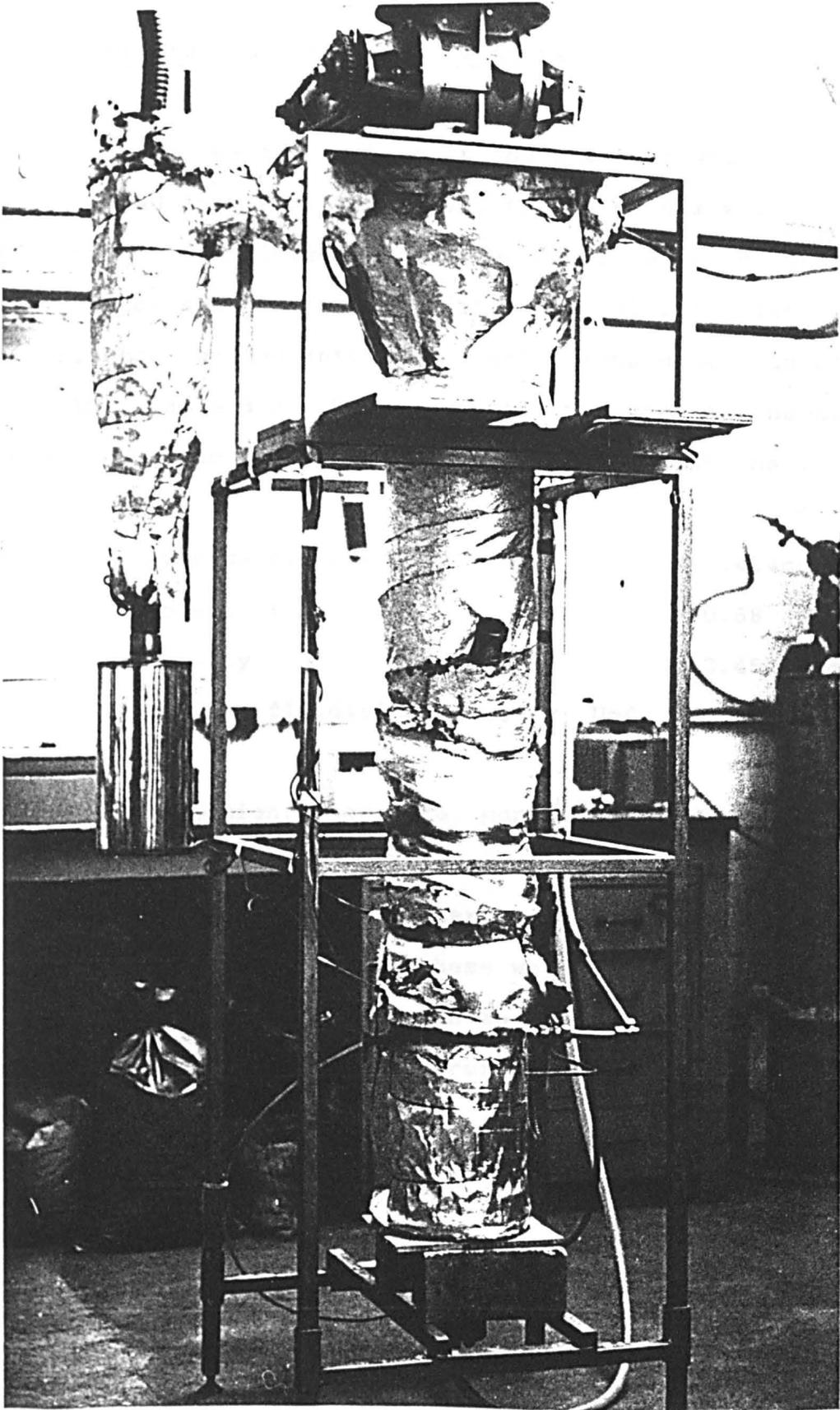
atmosphere after dilution through an extraction system.

3.2.1 Reactor And Preheater Sections

The reactor section of the gasifier was constructed from stainless steel tubing of 5mm wall thickness, 154mm internal diameter and 1 metre in length. The reactor sand bed was supported on a slotted plate gas distributor, welded into a removable flanged section of the tube 10cm in length. The sand which was used as the heat transfer medium was of the following specifications :

Average diameter, dp   = 0.414mm

Sphericity   = 0.68

Density   = 2.452Kg/m^3

Minimum fluidising velocity, Umf = 25cm/s

The reactor section, therefore, contained both the fluidised bed and the freeboard sections which were built with ports for insertion of thermocouples and sampling probes, dimensions at which these were placed is shown in Figure 3.2(b). The bed section and freeboard region were probed for gas samples to determine how the concentration

PLATE 2
Wall Mounted Control Panel

BACK PRESSURE
MONITORING

PREHEATER
AIR

TEMPERATURE
MEASUREMENT

FEED
CONTROL

PREHEATER
CONTROLS

NITROGEN
CONTROL
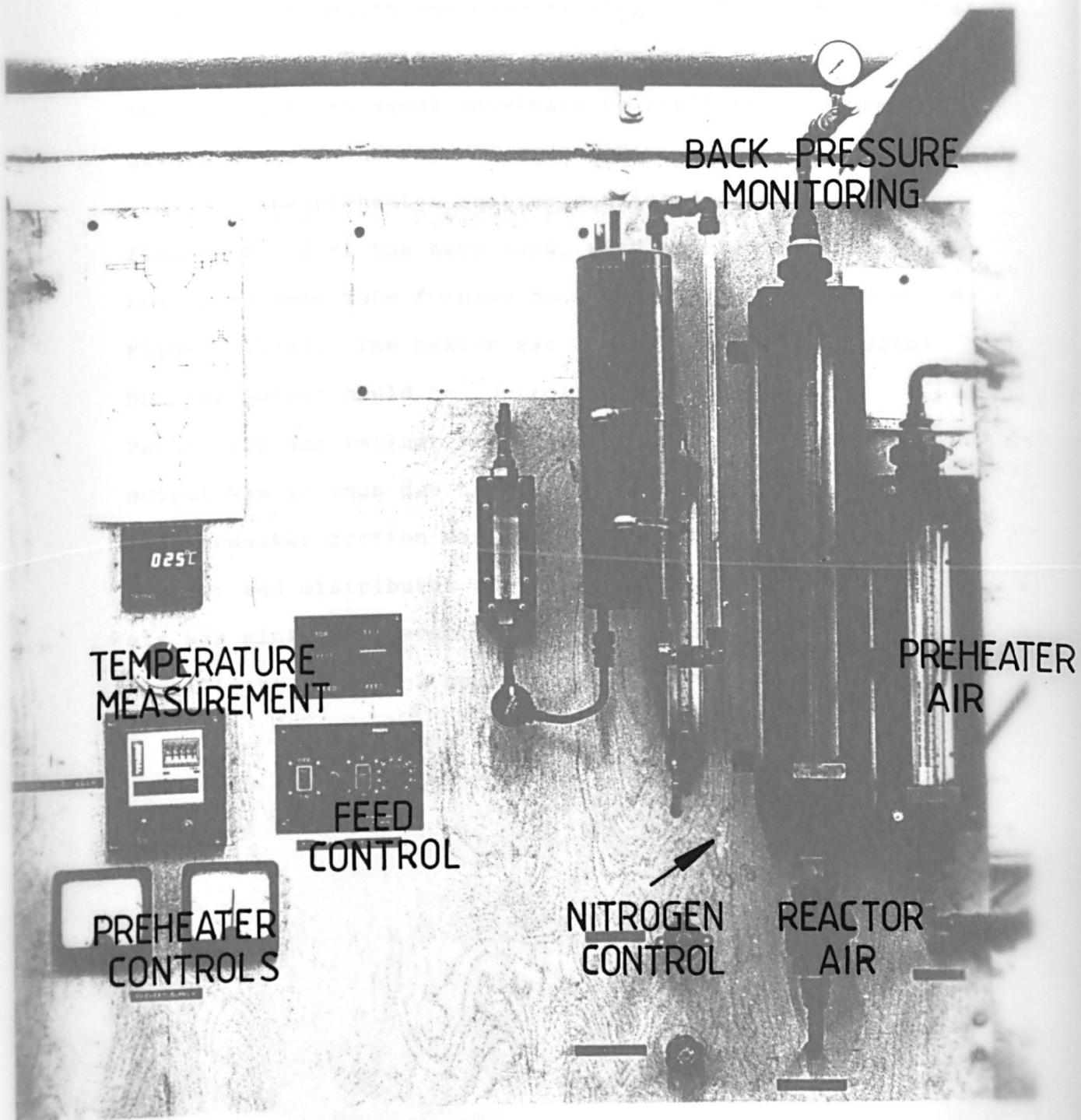
REACTOR
AIR

PLATE 2
Wall Mounted Control Panel
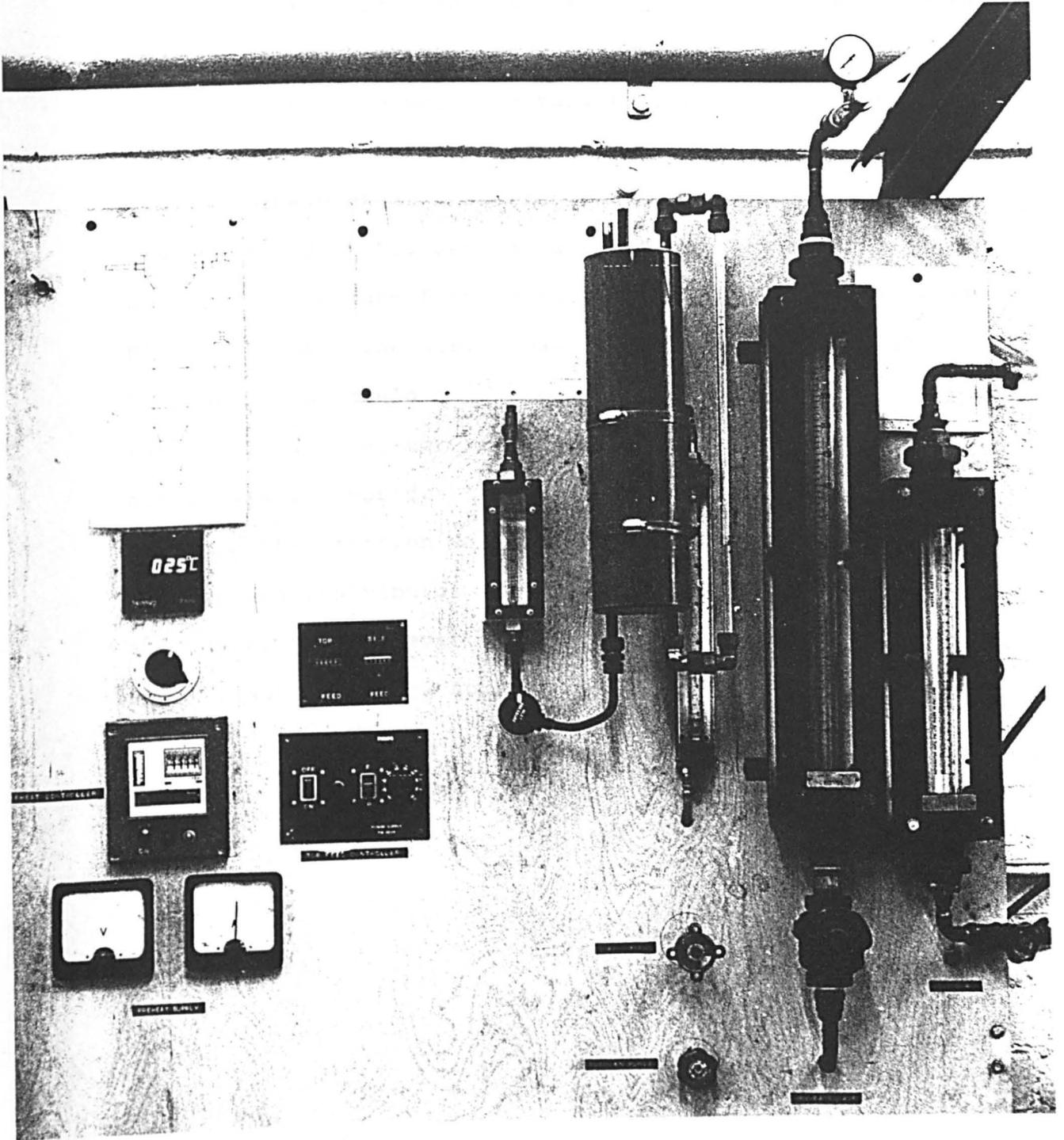
of gaseous species changed throughout the reactor section. Temperatures at various parts of the apparatus were monitored using chromel-alumel thermocouples linked to a multichannel switch and read directly from a Jenway digital display unit. Temperatures were recorded for all the thermocouples at equal intervals to yield temperature/time graphs for all the runs.

The preheater section consisted of a small fluidised bed of the same sand, with an electrically heated ceramic tube furnace mounted externally as shown in Figure 3.1(a). The heater was rated at 3 KW (240 volts) but the output could be controlled manually using a Variac (20 amp rating) voltage regulator. The maximum output was 10 amps due to the limit of the electric cables. The preheater section was located directly below the reactor bed distributor so that heat loss from preheated air was minimised, even though the heating period took a significant time (1-2 hours).

FIGURE 3·1 (b)

Locations of sample probes and thermocouple

3.2.2 Feeder And Cyclone

The feeding system adopted was that the woodchips were dropped through the freeboard section onto the fluidised bed from a 150mm internal diameter, cylindrical shaped, clear perspex storage hopper. The main problem with this feeding system was bridging in the hopper caused by the woodchip particles sticking together. This problem was overcome by incorporating a rotary fork in the hopper, which also regulated the flow of woodchips. The main regulator for the feeding mechanism was a rotary sealing valve with eight pockets, which gave a greater control of the feedrate. The drive for the fork was taken from the shaft of the rotary valve via sprocket and chain with the valve being driven by a 1/4h.p. DC. variable speed electric motor. The feedrate was varied by altering the voltage to the motor by a thyristor control unit. The woodchip particles were dropped through the freeboard section against the flow of the hot product gases, and were, therefore, preheated and pyrolysed to an extent before entering the hot reactor bed. To prevent leakage of the product gases through the hopper, the fuel was stored under a nitrogen blanket. this was necessary since the action of the rotary valve exchanged pockets of gas between the reactor and the hopper. Even though there was a slight leakage.

It was necessary to remove the elutriated particles of char and ash during the operation of the gasifier to obtain data which was essential when evaluating the overall material balance calculations. To this end a cyclone was included in the rig construction which was designed to operate most effectively at around 1m/s gas entry velocity. The dust collection vessel was connected to the cyclone via a gate valve so that it could be emptied while the gasifier was in operation. A port situated at the exit from the cyclone permitted the insertion of capillaries so that gas samples could be taken during a run for subsequent analysis.

3.2.3 Fluidising Air Velocity

Fluidising air velocity of 0.0716 m/s (15 °C) was chosen since it represented approximately 3 times minimum fluidising velocity at 750 °C. This was appropriate since this air velocity adequately fluidised the bed and kept elutriation to a minimum. Figure 3.2 shows the relationship between U_o/U_{mf} and temperature.

FIGURE 3·2 : Relationship between U_0/U_{mf} and temperature
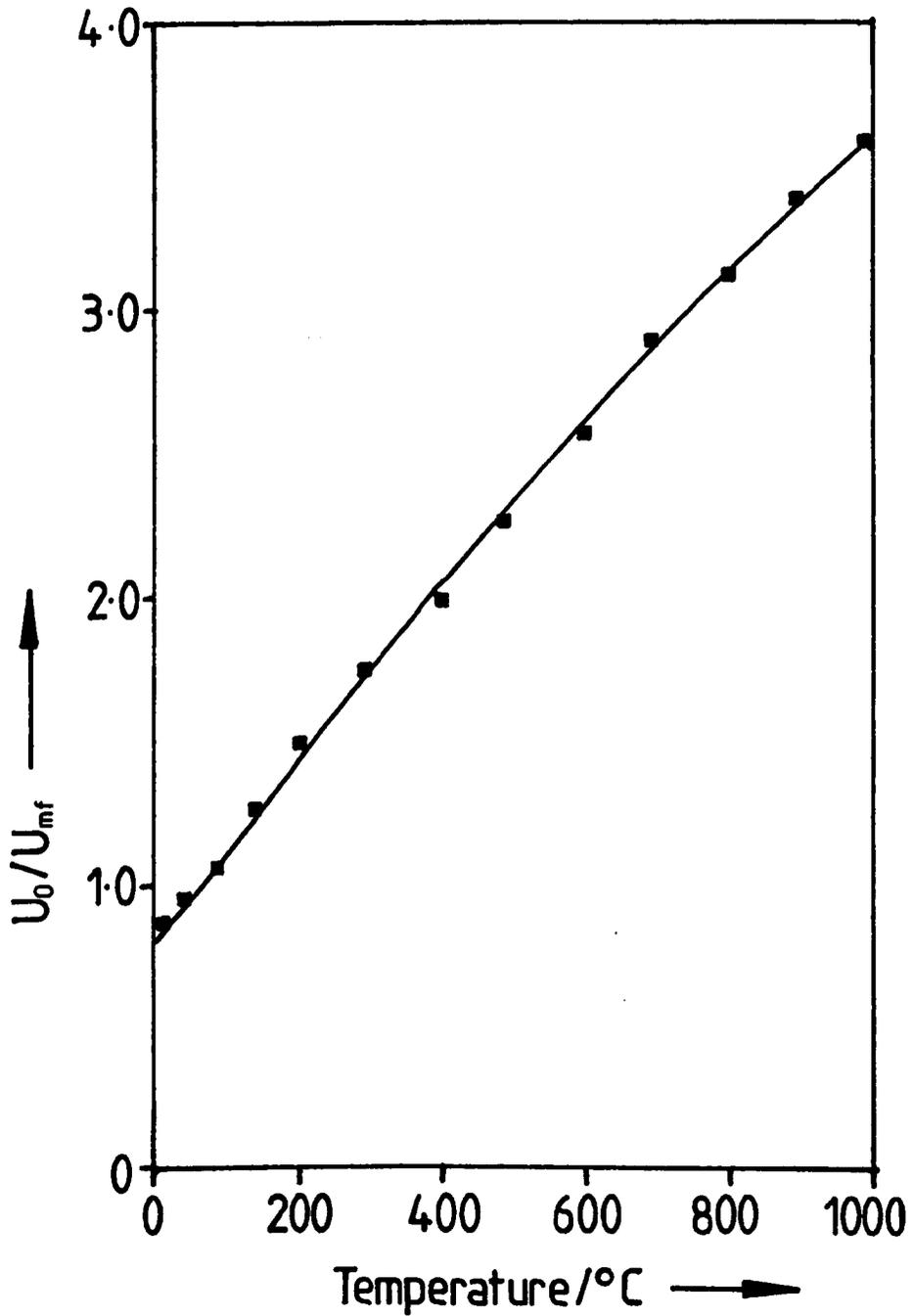
U_0 - Fluidising Bed Velocity

U_{mf} - Minimum Fluidising Velocity

3.3 EXPERIMENTAL PROCEDURE

At the start of each test the preheater was switched on after establishing a flow of air through both the fluidised beds. It was necessary for a small amount of air flow through the reactor bed to keep the hot air from the preheater section from burning the rubber tubing connecting the reactor bed by the effects of back pressure. The fluidising air velocity was usually about 7cm/s at 15 °C but to encourage fluidisation of the preheat section of the bed at the earliest possible moment the flow was switched in pulsing fashion from 7 to 18 cm/s. This procedure enhanced heat transfer from the external elecrically heated furnace to the preheater sand bed. Once fluidised, the sand in the bed rapidly attained a uniform temperature profile from top to bottom, thereby transfering heat more efficiently to the fluidising air. The temperature of the preheater bed was maintained at about 750 °C during the warm-up period. The hot air from the preheat section passed through the reactor bed which absorbed the heat along with the enclosing steel tubing, which was insulated with kaolinite blanket and reflective foil wrapping. The reactor bed was heated in this manner until the temperature reached 400-420 °C. At this stage small batches of fuel were introduced to the

reactor bed, so that the heat released by combustion increased the temperature to that which was required in the gasification tests, ie. 600-800 °C. The duration of the warm-up period depended upon the depth of the reactor sand bed (5-20 cm), from 1/2 hour to over 2 hours.

When continuous fuel feeding had been established and the desired reaction temperature was obtained, the hot air from the preheater section was altogether removed, by diverting the air around the preheat bed directly into the plenum chamber below the reactor bed. If further removal of heat was required to regulate the reactor bed temperature, this was done by removing small sections of the insulation around the fluidised bed reactor. This kept the operating temperature to within the required limits under investigation.

The samples needed for evaluating gas concentration along the bed required various sample probes. The sample probes were at 2 cm, 8 cm, 14 cm, 24 cm and 40 cm above the distributor plate as shown in Figure 3.1(b). These represented the bottom, top and freeboard area of the fluidised bed reactor. The syringes used to withdraw the gas samples were fitted with extra long stainless steel capillaries to enable a gas sample to be taken from the centre of the bed. The results from these sample probes enabled the bed gas concentration profiles to be

produced. In other tests, gas samples were extracted directly from the gasifier bed at several positions, above the distributor, in the reactor bed and also from the freeboard. This gave an overall picture of the gas profile along the length of the reactor bed. Gas samples were sucked through stainless steel capillaries (1mm i/d, 200mm long) by syringes (capacity 60ml), which were capped under positive pressure and set aside for analysis by gas chromatography at a later time. The use of thin walled stainless steel capillary tubing in the probe design permitted gas quenching before entering the syringe by convection currents of air. A small overlapping sleeve at the end of the probe contained silica fibres which acted as a filter preventing small char or ash particles from entering the capillary causing blockage.

On successful completion of an experimental run, the fuel feed was stopped and the reactor was purged with a nitrogen flow. To aid rapid cooling, sections of the kaowool insulation were removed. Flushing with nitrogen stopped carbon combustion occuring in the bed so that an examination of the contents of the reactor bed sand would reveal the char loading at the moment when air was replaced by nitrogen, assuming that elutriation during cooling was negligible. With woodchips there was negligible char

accumulation, it was therefore not necessary to purge every experimental run with nitrogen.

Gas samples taken during experimental runs were analysed using gas chromatography of both thermal conductivity and f.i.d. types. In the case of analysis for inorganics, 1ml samples were introduced into the tube columns of the Pye series 204 chromatograph. The column packing and other details are shown in the table below :

	COLUMN 1	COLUMN 2
Packing	Molecular sieve 5A 60 - 80 Mesh	Silica Gel 60 - 80 Mesh
Temperature	105 $^\circ$C	105 $^\circ$C
Carrier gas	Argon	Argon
Column length	2.44 metres	2.44 metres

The concentrations of hydrogen, oxygen, nitrogen and carbon-monoxide were found using the molecular sieve with carbon-dioxide found using the silica gel column, both using a thermal conductivity type detector.

The hydrocarbon components of the product gas mixture were found by analysis on a Pye Model R chromatograph with a sampling loop and pneumatically operated sampling valve giving extremely good repeatability of sample volumes. The chromatographic details are shown in the table below :

Column Packing	n-octane on paracil C support 80 - 100 Mesh
Oven Temperature	25 °C
Column Length	2.44 metres
Carrier Gas	nitrogen

The elution sequence of the hydrocarbons studied was as follows :

CH_4 - C_2H_4 +C_2H_6 - C_2H_2 - C_3H_8 - C_3H_6 - iso. C_4H_{10}

- nC_4H_{10} - nC_4H_8 - nC_5H_{12} - nC_5H_{10}

The calibrations of the gas chromatographs were performed using standard gases supplied in 4 litre cans by Phase Separations and were accurate to $\pm$ 5% of the stated values.

3.4 DISCUSSION OF EXPERIMENTAL RESULTS

3.4.1 Introduction

The reactor rig described earlier in the chapter was employed to investigate the experimental pyrolysis and gasification of woodchips. The conditions of operation and the composition of the woodchips are shown in Table 3.1.

TABLE 3·1

OPERATING PARAMETER	VALUE
Fuel composition by weight	carbon 47·9% hydrogen 6·0% oxygen 33·0% nitrogen 1·7% ash 0·4% moisture 11·0%
Fuel calorific value (gross)	19·11 MJ/Kg
Stoichiometric fuel feedrate (15°C)	16 g/min
Bed sand particle size	$\bar{d}_p = 0.53\,mm$
Approximate minimum fluidising velocity	$U_{mf} = 0.083\ m/s$
Fluidising air velocity (15°C)	0·0716 m/s (15°C) 0·25 m/s (750°C) (~3 × U_{mf})

The principal aims of the experimental tests on woodchips were as follows :

1. To determine the quality and composition of the product fuel gas.

2. To quantify the calorific output and the process efficiency.

3. To assess the loss to the the process through the elutriation of combustible fines.

4. To observe the operating characteristics of the fluidised bed while processing woodchips.

During the experimental tests the fuel/air ratio was varied by maintaining the fluidising air at a constant flow (approximately 3 times U_{mf}) and varying the feedrate of fuel to the reactor. The product gas quality and yield were obtained using the following calculation techniques :

3.4.1.1 Calculation Of Gas Yields

If the molar input of nitrogen to the product gas flow is assumed to be negligible (since it would contribute typically less than 0.001% to the flow), then the flow rate of dry product gas can be calculated using the following equation :

$$V_T = \frac{0.79}{y_{N_2}} \cdot V_A \qquad [3.1]$$

where y_{N2} = volume fraction of N2 in the product
gas from chromatographic analysis

V_A = inlet air flow rate 15 $^\circ$C

From the measured concentrations and the computed prduct gas flow rate the yield of each combustible component of the gas mixture can be calculated using the following equation :

$$Y_X = \frac{V_T}{M_F} \cdot y_X \qquad [3.2]$$

where M_F = fuel feedrate used (KG/min)

y_X = mol. fraction of gas component X (dry basis)

Y_X = yield of component X (Nm^3/KG)

3.4.1.2 Calculation Of Calorific Values

To quantify the heating potential of the fuel gases generated from the experimental tests the following formula was used :

$$CV = \text{Sum of } (y_X Q_X) \qquad [3.3]$$

where y_X = mol fraction (or volume %) of component X (hydrogen, oxygen etc.) in the dry product gas.

Q_X = gross or net heating potential of the gas component X (MJ/m^3)

In the cases where the fuel gas is required to be consumed hot, for example at close proximity to the reactor, then the CV of the gas will be complemented by its sensible

heat, hence we get :

$$CV = \text{Sum of } (y_x Q_x) + SH(T) \qquad [3.4]$$

where $SH(T)$ = sensible heat of the gas including moisture (MJ/Nm^3)

The CV's in this case would be net values since any latent heat from steam condensation will not be included. The net value of CV for any fuel is probably the more representative figure since combustion products inevitably leave the combustion zone at elevated temperatures carrying with them the latent heat of water (combustion product and fuel moisture) together with its sensible heat.

3.4.1.3 Calculation Of The Efficiencies

To describe the performance of a gasification process the following efficiencies are normally computed (SCHILLING et al., 1981) :

(a) GASIFICATION EFFICIENCY

This is the ratio of the chemically bound heat in the dry product gas to the amount of chemically bound heat in the fuel, ie. :

$$n_G = \frac{CV_{(gas)}}{CV_{(fuel)}} \cdot 100 \qquad [3.5]$$

(b) THERMAL EFFICIENCY

This is the ratio of the amount of useful chemically bound and sensible heat in the products obtained by gasification to the amount of chemcally bound and sensible heat in the fuel and the gasifying medium, ie. :

$$n_T = \frac{CV_{(gas)} + SH_{(gas)} + SH_{(steam)}}{CV_{(fuel)} + SH_{(fuel)} + SH_{(air)}} \, 100 \quad [3.6]$$

where $CV_{(gas)}$ and $CV_{(fuel)}$ are net values.

This value of efficiency for a hot gas would be more correct if the contribution of tar vapours of the total gas CV were considered. This is particularly appropriate when fuels with high volatile contents are being processed as in this work with woodchips.

3.4.1.4 Calculation Of Mass Balances

By performing mass balance on material input and output from the reactor we can compile an inventory of each elemental component of the system. The approach of the mass balance calculations to total accountability of individual components is called the mass balance efficiency and is given by the following equation :

$$n_{mx} = \frac{m_{in}}{m_{out}} \, . \, 100 \quad [3.7]$$

where X is a component of the mass input, carbon, hydrogen, oxygen and ash.

3.4.2 Temperature/Time History

Temperature/time history data were taken, as shown in Figures 3.3 to 3.6, to show for typical tests, for varying bed depths (5, 10, 15 and 20cm static bed depths), how the reactor bed, freeboard and cyclone temperatures varied with time. Also shown are the times when gas samples were taken, and also indicated is the warming up period and the running period. The Figures show that a typical test lasts approximately 3 hours, with haly the time required for the warming up period. This limits the running period since an average of 4 to 5 samples taken during the test requires a further 3-4 hours for chromatographic analysis. Since it was necessary to analyse the product gas on the same day as the test, because of the inevitable leakage of the product gas from the syringes, the running period was restricted to a maximum of 3 hours.

In the following sections graphical output obtained from the experimental tests, to investigate the effect of the operating conditions, will be presented and discussed. The tabulated results for the quality and yields of the product gas can be found in the appropriate Appendices.

FIGURE 3·3

Temperature/time history, bed depth = 5cm,
feedrate = 2·8 × S

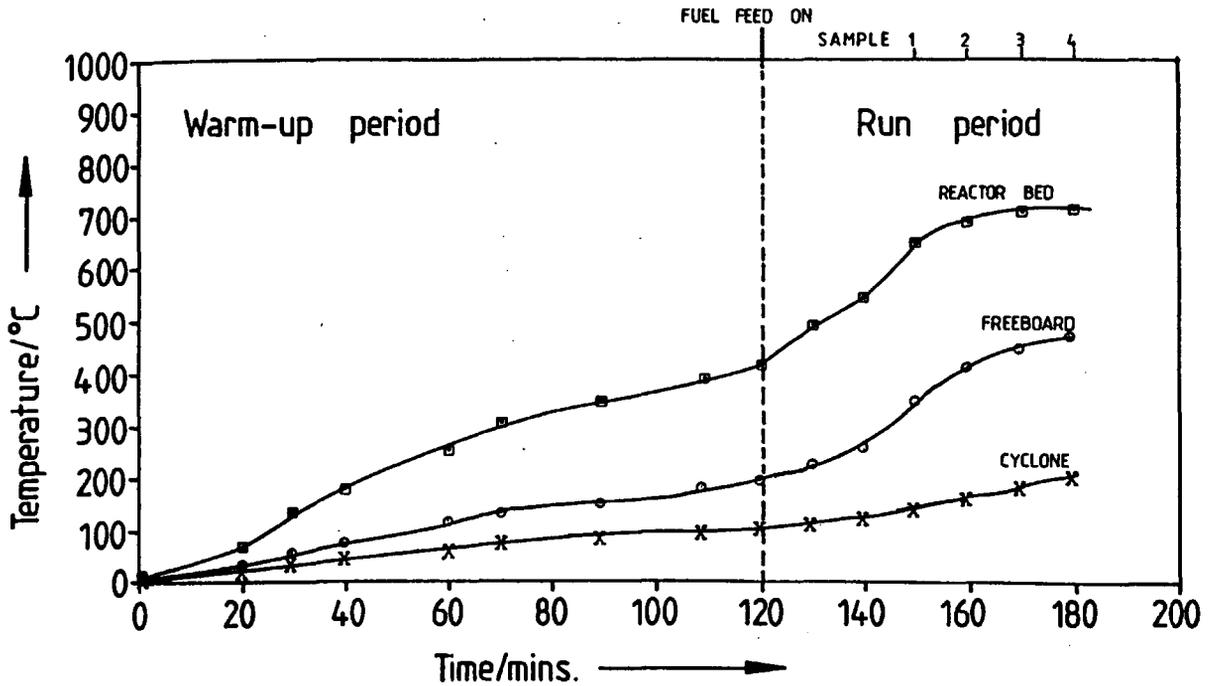
FIGURE 3·4

Temperature/time history, bed depth = 10 cm,
feedrate = 2·9 × S
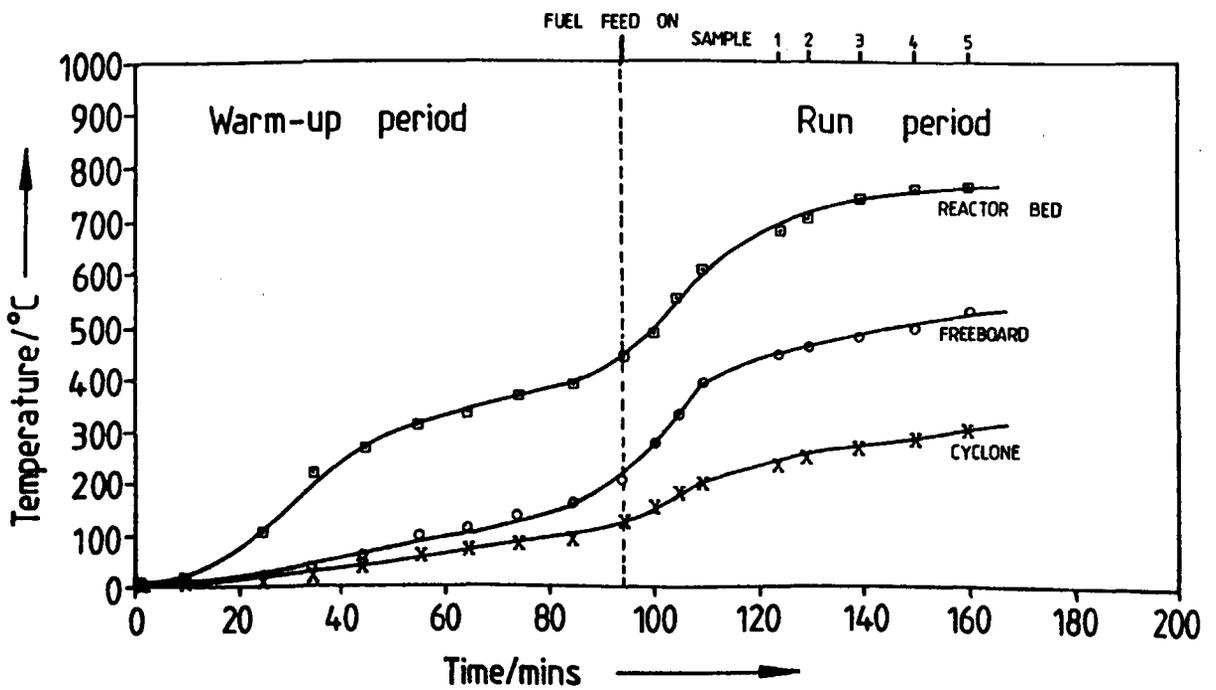
FIGURE 3·5

Temperature/time history, bed depth = 15 cm,
feedrate = 2·9 × S
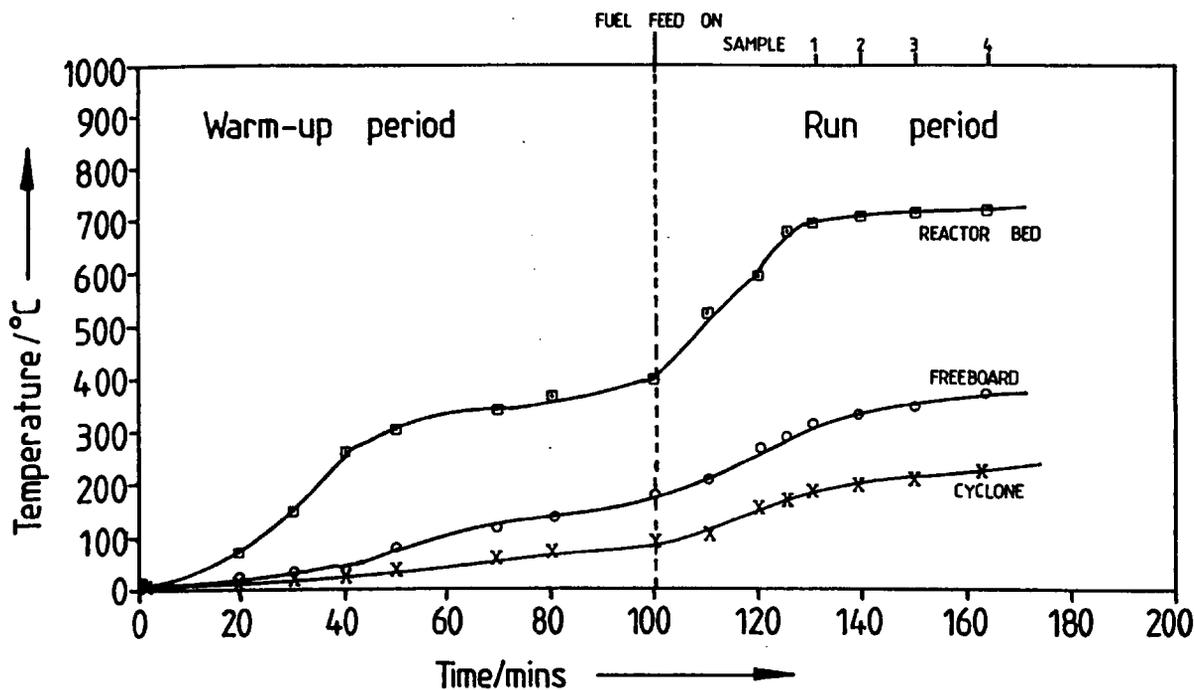
FIGURE 3·6

Temperature/time history, bed depth = 20 cm,
feedrate = 2·6 × S
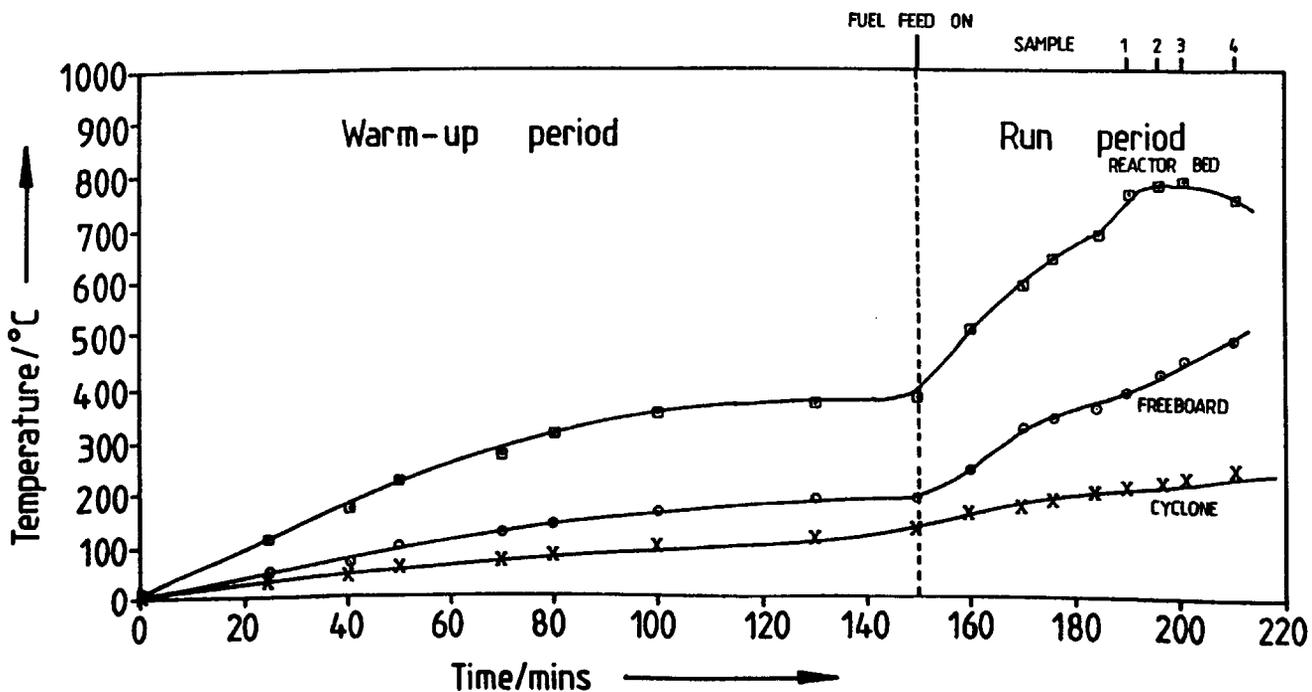
3.4.3 Significance Of Reactor Bed Gas Profiles

Any discussion of the results regarding the effect of the process variables on the quality and quantity of producty fuel gas can only benefit from a clear understanding of the chemical and physical processes occuring in the fluidised bed reactor. It is therefore, the object of this section to attempt to describe these processes from the experimental results obtained.

Information regarding the concentration of gaseous species at several distances above the distributor plate have been obtained in the form of a concentration profile through the bed. Samples were extracted by probing the bed under several differing conditions of fuel feedrate, in conjuction with product gas samples taken at the cyclone exit pipe. The results presented here are shown in graphical representations in Figures 3.7 to 3.15. The corresponding Tables (Tables 1 to 9) are presented in Appendix 1. These results also include experiments which were conducted to gather information on the effect of bed depth on concentration profiles of gaseous species.

The processes occuring in the bed are chemically and physically of a complex nature. Due to the fluidisation phenomenon we can say that the bed is made up of two phases, bubble and emulsion (or dense phase). Flow of gas through the bed exceeding that required to suspend

the bed particles takes the form of bubbles which rise from the distributor holes, growing in size until they burst at the bed surface throwing solids into the freeboard both vertically and laterally. The bubble movement, therefore, causes solid recirculation within the bed. As the bubble rises, gas is interchanged with the surrounding emulsion phase where the bulk of the solid-gas reactions take place.

FIGURE 3·7 : BED PROFILE (RUN 1)

Bed depth = 10 cm

Fuel feedrate = 2·0 times stoichiometric

FIGURE 3·8 : BED PROFILE (RUN 2)

Bed depth = 10 cm

Fuel feedrate = 3·1 times stoichiometric
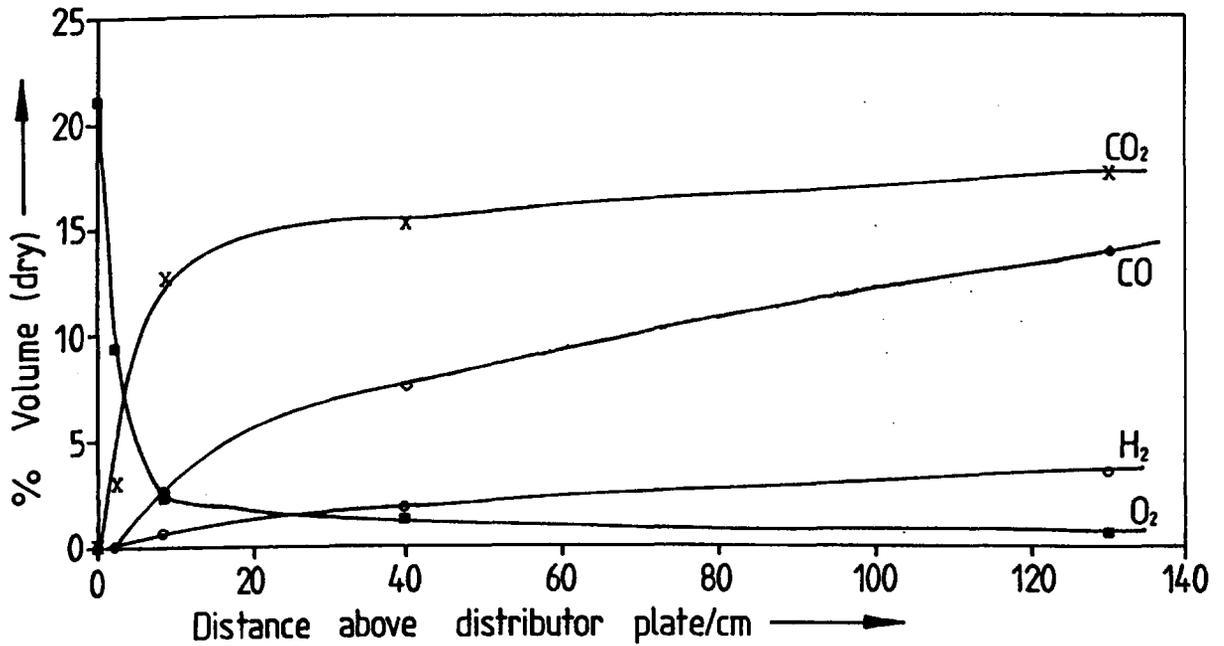
FIGURE 3·9 : BED PROFILE (RUN 3)

Bed depth = 10cm

Fuel feedrate = 3·1 times stoichiometric
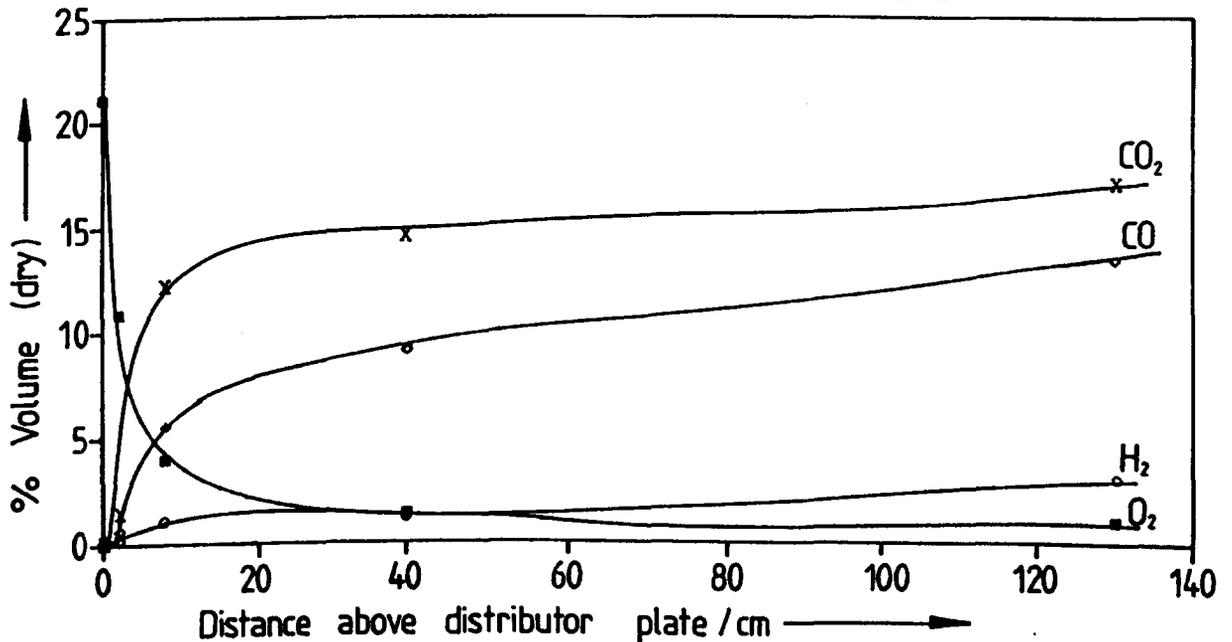
FIGURE 3·10 : BED PROFILE (RUN 4)

Bed depth = 10 cm

Fuel feedrate = 4·1 times stoichiometric
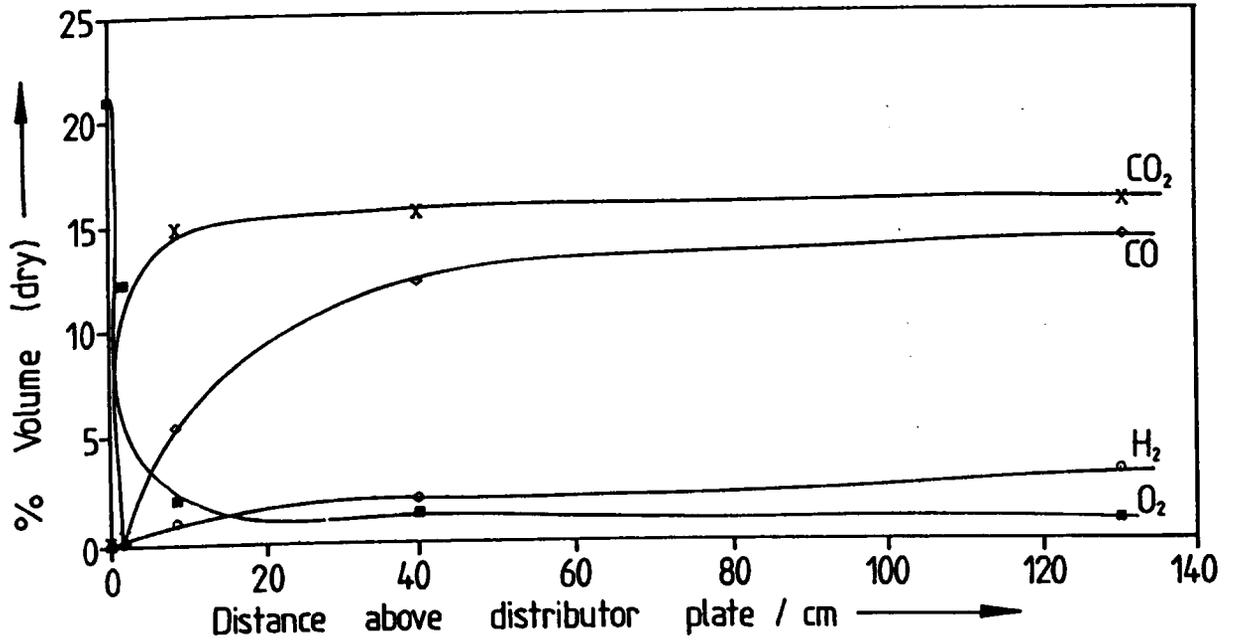
FIGURE 3·11 : BED PROFILE (RUN 5)

Bed depth = 10 cm

Fuel feedrate = 2·3 times stoichiometric
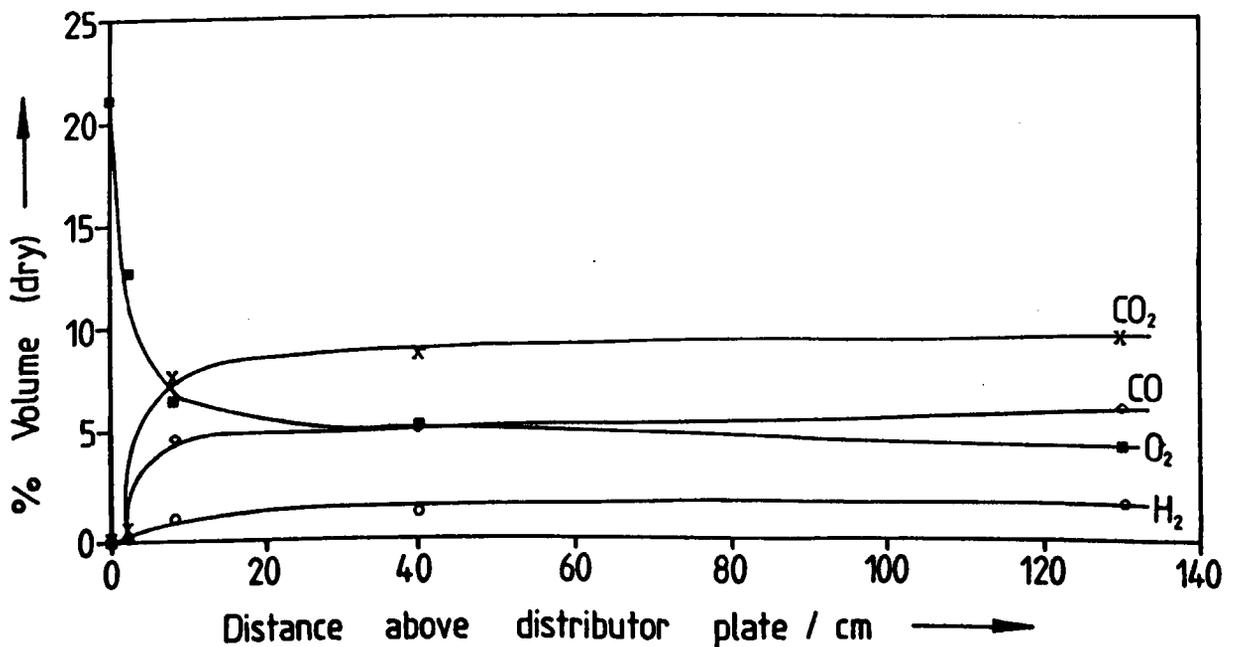
FIGURE 3·12 : BED PROFILE (RUN 6)

Bed depth = 10 cm

Fuel feedrate = 3·9 times stoichiometric
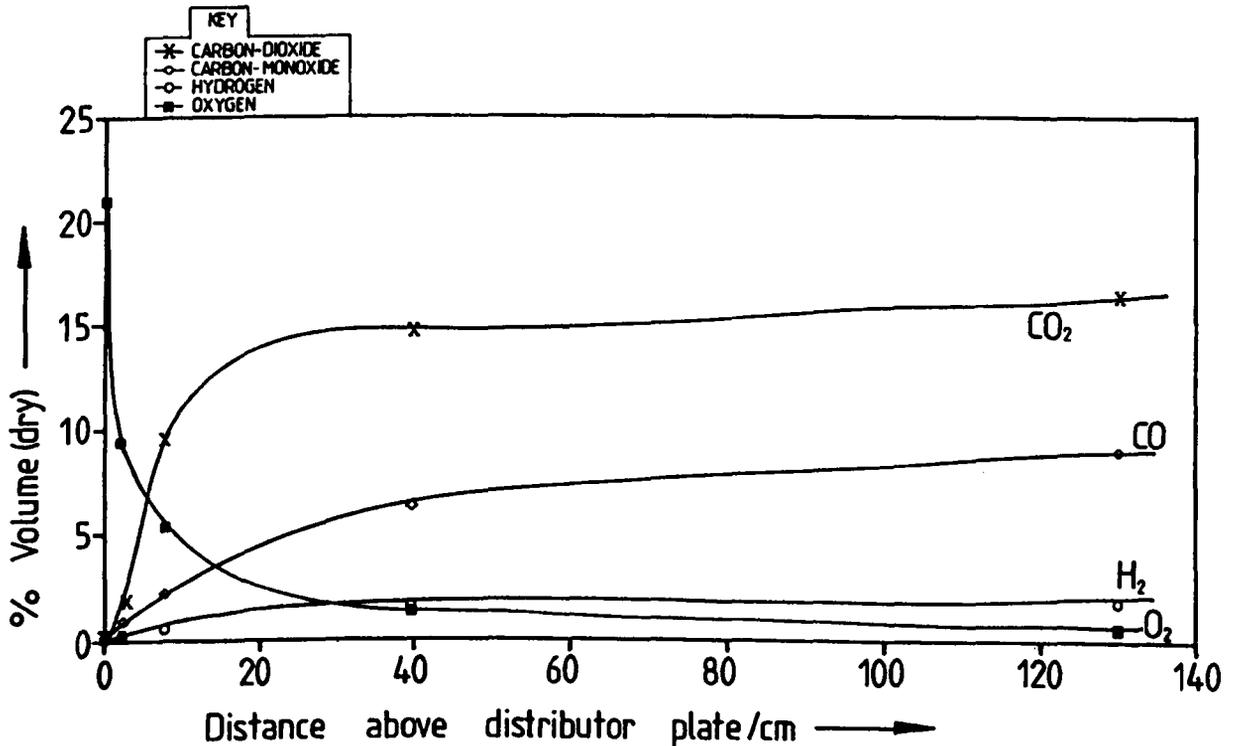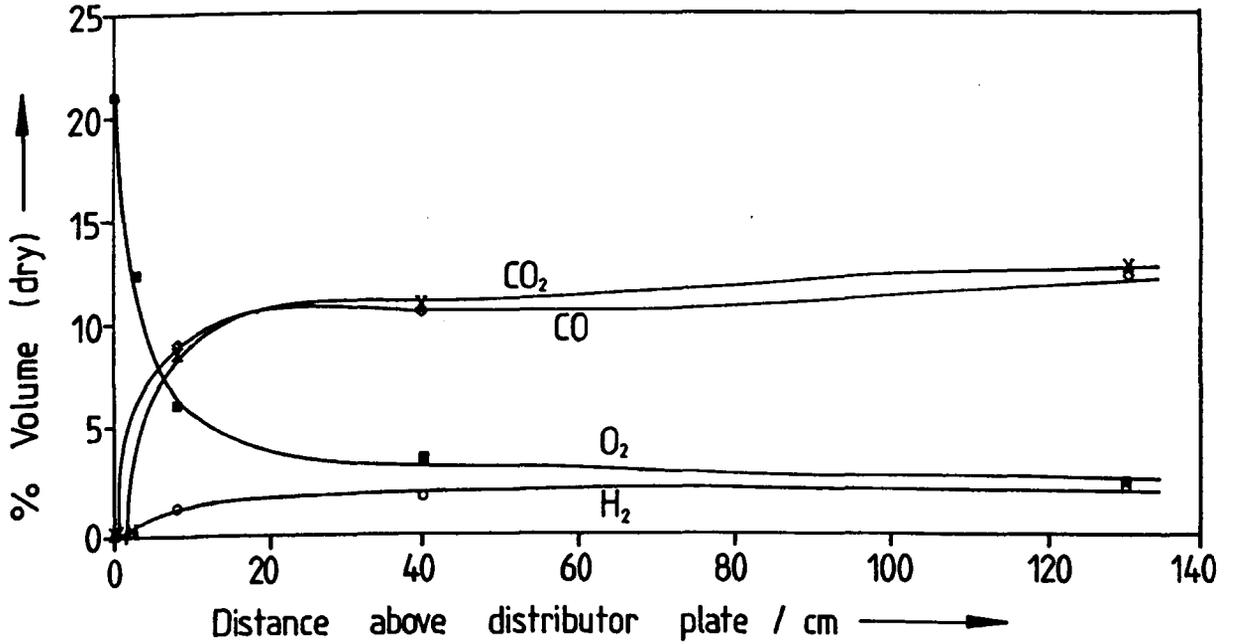

FIGURE 3·13 : BED PROFILE (RUN 7)

Bed depth = 10 cm

Fuel feedrate = 2·6 times stoichiometric
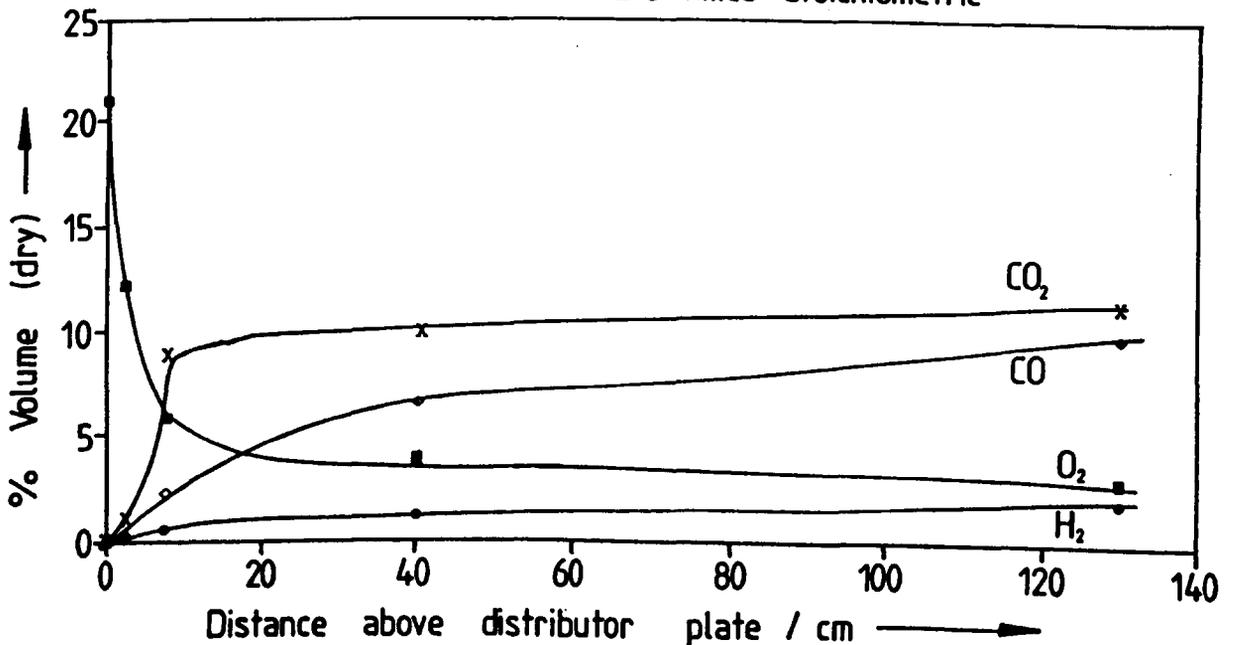
FIGURE 3·14 : BED   PROFILE   (RUN 8)

Bed   depth = 15 cm

Fuel   feedrate = 2·4   times   stoichiometric
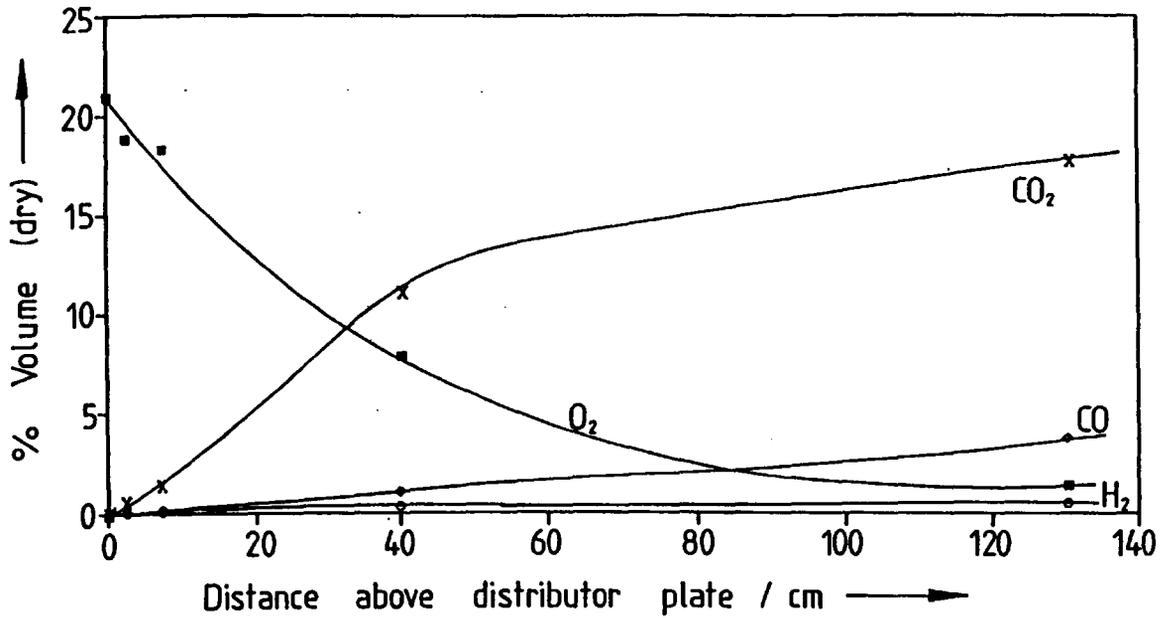
FIGURE 3·15:BED   PROFILE   (RUN 9)

Bed   depth = 5 cm

Fuel   feedrate = 3·6   times   stoichiometric
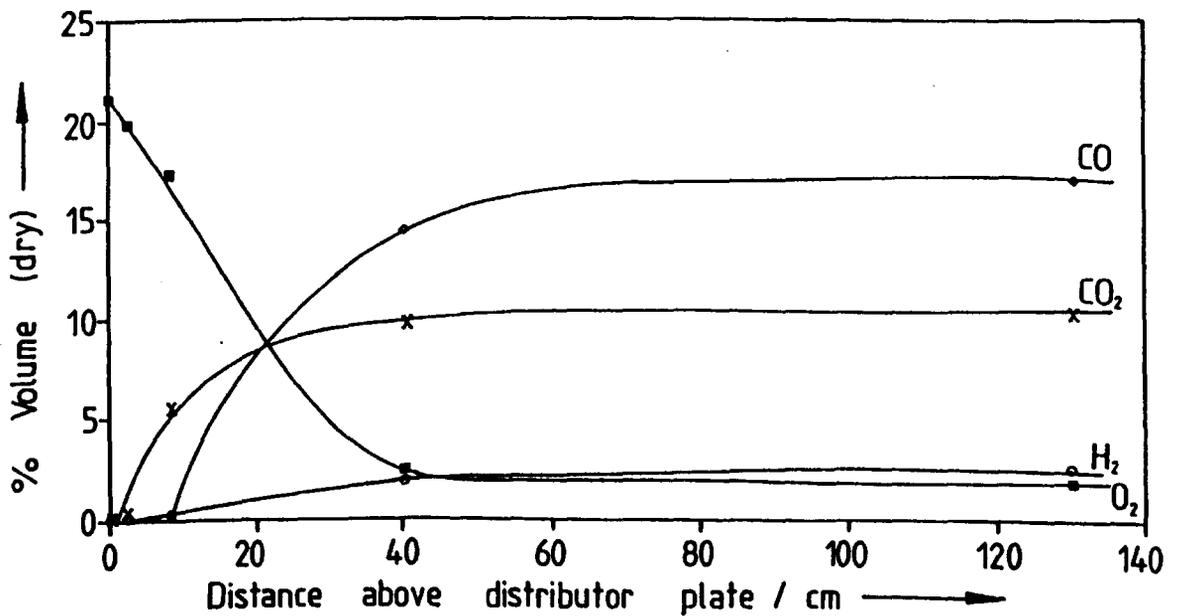
When the fuel particles are introduced to the top
of the freeboard, they fall against the flow of the hot
product gases. During this time devolatilisation occurs
with subsequent pyrolysis of the tars and hydrocarbons.
The air at the distributor plate forms bubbles which rise
through the bed constantly being swept out into the dense
phase. The oxygen is therefore consumed in the combustion
of the char as it passes through the bed. Figures 3.7 to
3.13 show clearly that the bulk of the oxygen (> 50%), in
the air, is consumed in the lower portion of the bed
leaving the upper portion of the reactor rig relatively
free of oxygen. It is therefore in this upper region of
the reactor rig where the initial devolatilisation occurs.
The combustible volatiles are permitted to escape into the
upward flow of the product gas. Any combustible volatiles
retained by the fuel char particles will then be completely
combusted in the lower oxygen rich region of the bed due to
the induced recirculation of the solids, or they are
carried out of the bed in the product stream as elutriated
solids.

3.4.3.1 Effect Of Fuel Feedrate On Bed Profiles

The effect of fuel feedrate on concentration
profiles of the gaseous species along the length of the
reactor rig can be better understood if we consider the
oxygen consumption, carbon-dioxide and carbon-monoxide

production at various distances above the distributor
plate.

The basic assumptions used to evaluate the
percentages of these three gases are that at the
distributor plate there is 100% oxygen available, while at
the exit there is 0% available. The production of
carbon-dioxide and carbon-monoxide is assumed to be 0% at
the distributor plate and 100% at the exit. All the other
percentages are then evaluated correspondingly as shown in
Table 3.2.

Firstly considering the availability of oxygen
along the length of the reactor rig with respect to fuel
feedrate, as shown in Figure 3.16. At the distributor
plate there is 100% of the incoming oxygen available for
consumption. While at 2cm above the distributor 40-60% of
the oxygen has been consumed, leaving between 60-40%
available for consumption. The greater consumption of
oxygen is associated with lower fuel feedrates. Conversely
a greater percentage of the oxygen is available at 2cm
above the distributor with increasing fuel feedrates. At
8cm above the distributor a similar trend is observed.
While at 40cm the available oxygen is constant for various
fuel feedrates, at about 4%. From these observations the
main conclusion which can be made is that the majority of
the oxygen is consumed in the lower and middle sections of

the bed (ie. 0-2cm and 2-8cm above the distributor plate).
Also a greater oxygen consumption is obtained with lower
fuel feedrates.

FIGURE 3·16

% Available oxygen at various distances above the distributor with respect to fuel feedrate (bed depth = 10 cm).
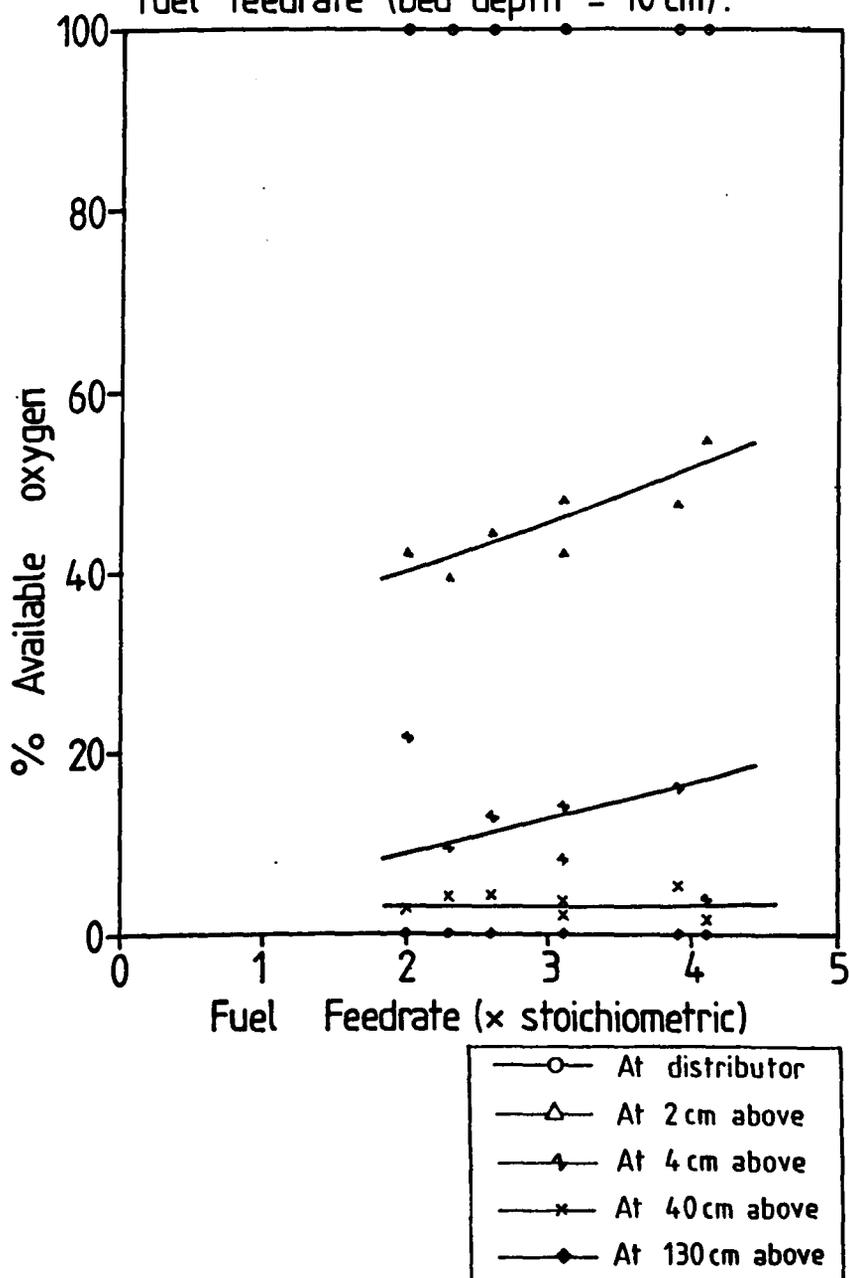
TABLE 3·2

Gas concentrations At Various Distances Above The Distributor Plate

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FUEL FEEDRATE (× S)	% GAS CONCENTRATION AT VARIOUS DISTANCES ABOVE THE DISTRIBUTOR PLATE																			
	0 cm				2 cm				8 cm				40 cm				130 cm			
	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2
2·0	100·0	0·0	0·0	0·0	37·7	6·6	8·7	0·42	20·8	21·0	50·7	0·23	6·2	65·8	85·5	0·74	0·0	100·0	100·0	0·64
3·1	100·0	0·0	0·0	0·0	37·8	0·0	11·3	0·0	7·7	11·9	52·5	0·18	3·8	44·0	71·8	0·50	0·0	100·0	100·0	0·81
3·1	100·0	0·0	0·0	0·0	43·5	3·0	6·5	0·36	14·3	34·6	58·0	0·47	1·3	58·6	74·0	0·62	0·0	100·0	100·0	0·79
4·1	100·0	0·0	0·0	0·0	51·0	0·0	1·9	0·0	2·7	29·9	76·4	0·35	1·3	79·2	90·7	0·78	0·0	100·0	100·0	0·89
2·3	100·0	0·0	0·0	0·0	41·9	2·1	5·4	0·25	10·5	70·8	79·7	0·58	4·0	83·3	86·5	0·62	0·0	100·0	100·0	0·65
3·9	100·0	0·0	0·0	0·0	46·0	0·0	0·0	0·0	19·0	64·5	60·5	1·03	6·6	81·8	79·8	0·99	0·0	100·0	100·0	0·96
2·6	100·0	0·0	0·0	0·0	44·4	2·4	6·5	0·33	12·6	18·3	70·6	0·23	3·7	61·0	83·7	0·65	0·0	100·0	100·0	0·89

To explain what is occuring in the lower and middle section of the bed we need to consider how this oxygen is being consumed or rather what products are being formed. Since the main products of oxygen are carbon-monoxide and carbon-dioxide, we need to quantify these products with respect to the lower and middle section of the bed.

Considering the lower section of the bed (0-2 cm above the distributor plate). The production of both carbon-monoxide and carbon-dioxide decreases with increasing fuel feedrate, as shown in Figures 3.17 and 3.18. This explains the decrease in oxygen consumption with increasing fuel feedrates. From Figure 3.19 the CO/CO_2 ratio decreases with increasing feedrate, therefore the production of carbon-dioxide exceeds that of carbon-monxide. The reason for this is that since there is a greater percentage of oxygen available, the production of carbon-dioxide is favoured. This reaction also produces heat.

In the middle of the bed (2-8 cm above the distributor plate), the production of carbon-monoxide and carbon-dioxide, both increase with increasing fuel feedrate, as shown in Figures 3.17 and 3.18. Infact between 20-35% of the overall carbon-monoxide and between 50-70% of the overall carbon-dioxide has been produced in this section. This confirms the availability curves of

oxygen, Figure 3.16, which shows that most of the oxygen, between 80-90%, has been consumed within the lower and middle sections of the bed. The CO/CO₂ ratio starts to increase in the middle section with increasing fuel feedrates. Hence the production of the carbon-monoxide starts to increase with respect to carbon-dioxide. This can be explained by the fact that less oxygen is available for consumption, thus promoting the production of carbon-monoxide.

FIGURE 3·17

% Carbon-monoxide produced at various distances above the distributor with respect to fuel feedrate (bed depth = 10 cm).

—o—	At distributor
—△—	At 2 cm above
—⊬—	At 4 cm above
—x—	At 40 cm above
—◆—	At 130 cm above
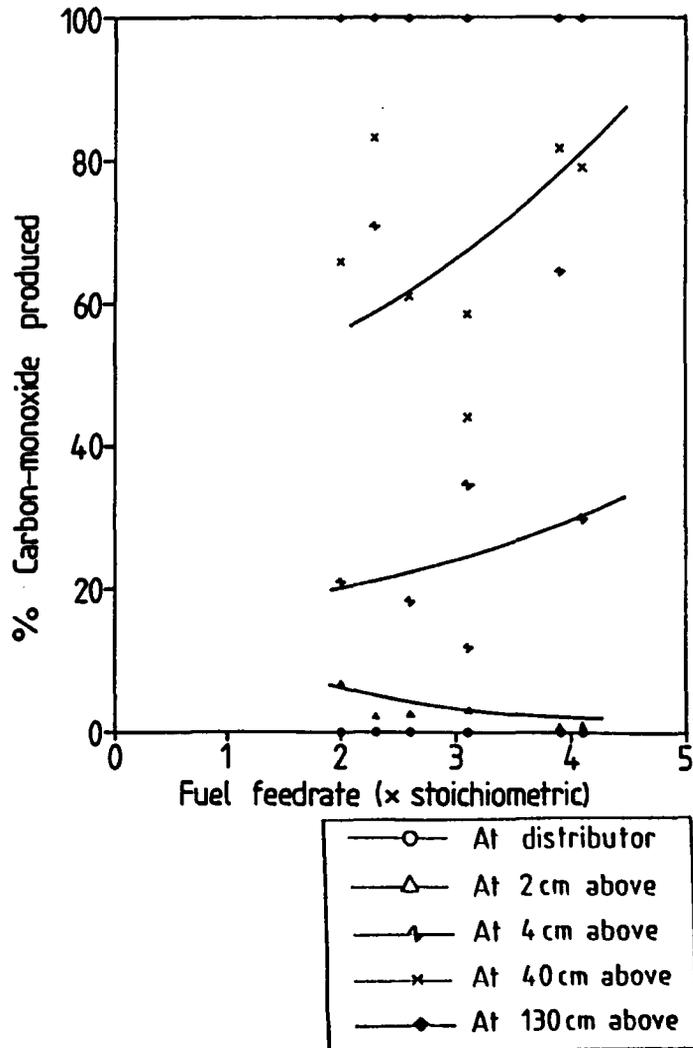
FIGURE 3·18

% Carbon-dioxide produced at various distances above the distributor with respect to fuel feedrate (bed depth = 10 cm)

Legend:
- ─○─ At distributor
- ─△─ At 2 cm above
- ─◣─ At 4 cm above
- ─×─ At 40 cm above
- ─◆─ At 130 cm above
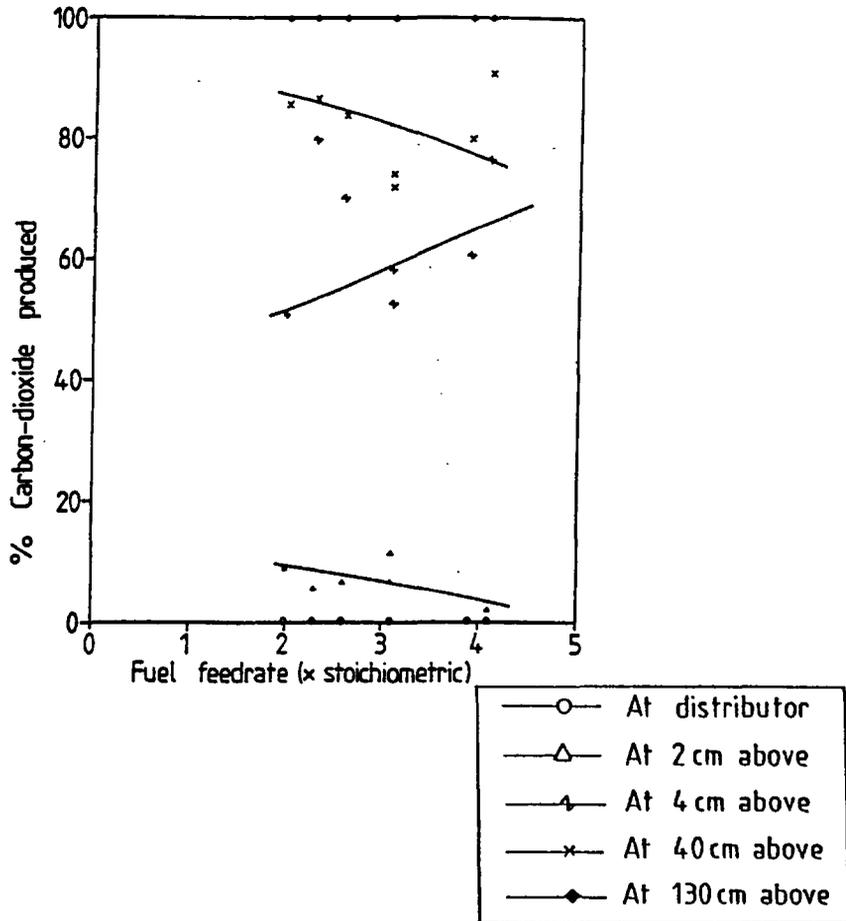
FIGURE 3·19

CO/CO_2 ratio at various distances above the distributor with respect to fuel feedrate (bed depth = 10 cm).
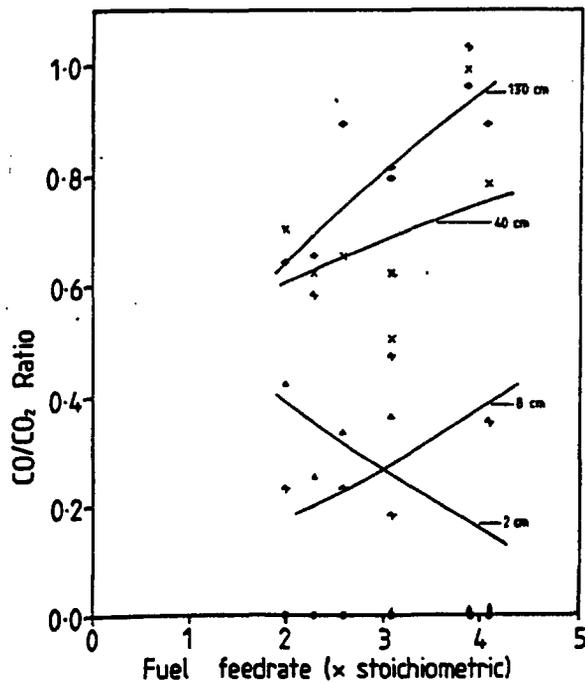
In the freeboard region (40cm above the
distributor plate), production of carbon-dioxide starts to
decline in favour of carbon-monoxide, with increasing fuel
feedrate. This is confirmed by the increasing of the
CO/CO_2 ratio with increasing fuel feedrate.

3.4.3.2 Effect Of Bed Depth On Bed Profile

The effect of bed depth on concentration profiles
of the gaseous species along the length of the reactor rig
was only attempted, from the limited results obtained,
shown in Table 3.3. The results cannot be easily
explained since there is no obvious trend. The only
observation which follows any trend is that for 40cm above
the distributor plate. Where the available oxygen
increases with bed depth. This is further confirmed by the
decline in the CO/CO_2 ratio.

There are no significant conclusions from these
results, perhaps considering further bed depths might have
provided some trends.

Since the woodchips are being fedd by dropping
through the freboard region, bed depth might not play a
significant role in the determination of the gasification
reactions occuring within the reactor rig. Nevertheless
the quality of mixing which defers with bed depth might be
an important factor, in that it will obviously affect the
availability of oxygen along the length of the reactor rig.

Which in turn will affect the production of carbon-monoxide and carbon-dioxide. Since limited results were obtained regarding the effect of bed depth no quantative comparisons were possible.

TABLE 3·3

Gas Concentrations At Various Distances Above The Distributor Plate With Respect To Bed Depth

| BED HEIGHT (cm) | % GAS CONCENTRATION AT VARIOUS DISTANCES ABOVE THE DISTRIBUTOR PLATE | | | | | | | | | | | | | | | | | | | |
| | 0 cm | | | | 2 cm | | | | 8 cm | | | | 40 cm | | | | 130 cm | | | |
	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2	O_2	CO	CO_2	CO/CO_2
5	100·0	0·0	0·0	0·0	85·1	0·0	0·2	0·0	72·6	0·0	54·2	0·0	2·1	84·7	96·3	1·48	0·0	100·0	100·0	1·68
10	100·0	0·0	0·0	0·0	41·9	2·1	5·4	0·25	10·5	70·8	79·7	0·58	4·0	83·3	86·5	0·62	0·0	100·0	100·0	0·65
15	100·0	0·0	0·0	0·0	83·1	0·0	4·5	0·0	80·0	0·8	8·0	0·02	30·5	28·2	62·9	0·10	0·0	100·0	100·0	0·22

3.4.4 The Effect Of Varying Process Conditions On Product Gas Quality And Yield

The gasification of solid fuel in a medium of air produces a large volume of gas (per KG of fuel) than in a medium of oxygen or steam. This is due to the presence of large volume of nitrogen from the air which acts as a dilutent to any combustible prduct gas.

In general gasification with air gives a gas of CV less than 6 MJ/Nm^3 due to the high nitrogen content. If a gas of greater heating value is required then oxygen/steam or hydrogen media must be employed.

The experimental results derived to investigate the effect of varying process conditions on product gas quality and yield are presented in Appendix 2, Tables 1 to 15 (a-e). These results represent all the data collected to study the following :

      (i)   the effect of fuel feedrate

      (ii) the effect of temperature

      (iii) the effect of bed depth

3.4.4.1 Effect Of Fuel Feedrate

The effect of fuel feedrate was investigated for various temperature ranges. The rate of introduction of woodchips into the reactor was varied over a range of fuel factors (multiples of the stoichiometric fuel feedrate, S) from 1.9 to 4.3. The results obtained from these experiments are presented in Figures 3.20 to 3.25 (a-c). The notation for fuel feedrate is given by 'S' (or fuel factor).

The yield (Nm^3/Kg) and output (Nm^3/min) of the major gaseous components are shown in Figures 3.20 to 3.25 (a-b, respectively). It is immediately obvious that the total specific yield of product gas decreases with fuel feedrate, while the output increases. The output of the product gas increases due to the larger amounts of pyrolysis products entering the gas stream (ie. CO, H_2, CH_4, C_nH_m, CO_2, etc.). While the yield of the product gas per Kg of fuel decreases due to the amount of air, or in particular N_2, which decreases per Kg of fuel as the fuel feedrate is increased.

The yield and output of H_2, CH_4 and C_nH_m tends to be fairly constant or show a gentle increase as the fuel feedrate is increased, depending upon bed depth and temperature range. The accumulation of these gaseous components contributes less. than 10% of the total gaseous output or yield. CO, CO_2 and N_2 are the major gaseous

gaseous components. The trends in the yield and output of CO and CO_2 are not obviously clear nor can they be easily explained. If we consider Figures 3.20 to 3.22 (a-b), which are results obtained at a constant bed depth of 10 CM, but for various temperature ranges. It is clear from the curves of the output of CO, which increase with increasing fuel feedrate. While the output of CO_2 decreases slightly. This can be explained simply, due to the decrease in the availability of O_2 for combustion at increasing fuel feedrates, favouring the production of CO. Also the greater increase in CO production compared to the slight decrease in CO_2 production, is explained by the fact that the available O_2 produces 2 molecules of CO compared to 1 molecule of CO_2, assuming carbon is available. Considering the yield of CO and CO_2, both decrease with increasing fuel feedrate at lower temperature ranges. At higher temperature ranges (1021-1080 K), the yield of CO increases slightly, while the decrease of CO_2 is relatively greater with increasing fuel feedrate. This can be explained again by the lower availability of oxygen at higher fuel feedrates favouring the production of CO. The effect of bed depth will be explained in another section.

The quality of the fuel gas improved markedly with an increase in fuel feedrate as shown in Figures 3.20 to 3.25 (c). The higher calorific values at higher fuel feedrates are mainly due to the the larger volumes of

decomposition products (H_2, CO, CO_2, CH_4, $C_n H_m$) entering the product gas per unit volume of air input. The result is a marginally greater volume output of product gas which is richer in combustible components than at lower fuel feedrates. Fuel gases with calorific values in excess of 4.5 MJ/Nm^3 (4.2 MJ/Nm^3 net) have been obtained consistently, at higher temperature ranges and fuel feedrates, which represents a useful alternative to a coal derived low calorific producer gas.

Although the results show a useful increase in the gas quality with increase in fuel feedrates, the calorific output from the gasifier, as defined in terms of MJ out per Kg of woodchips fed in, shows a decrease with increasing fuel feedrates, Figure 3.26. The drop in efficiency at higher fuel feedrates is due to the 'diminishing returns' effect, whereby an increase in the fuel feedrate in not matched by a corresponding rise in the calorific value of the fuel gas produced. In other words the calorific value of the gas, in MJ/Nm^3 tends to approach an upper limit at higher fuel feedrates which results as drop in efficiency.

FIGURE 3·20 (a)

Bed depth = 10 cm, Temperature = 941 – 990 K
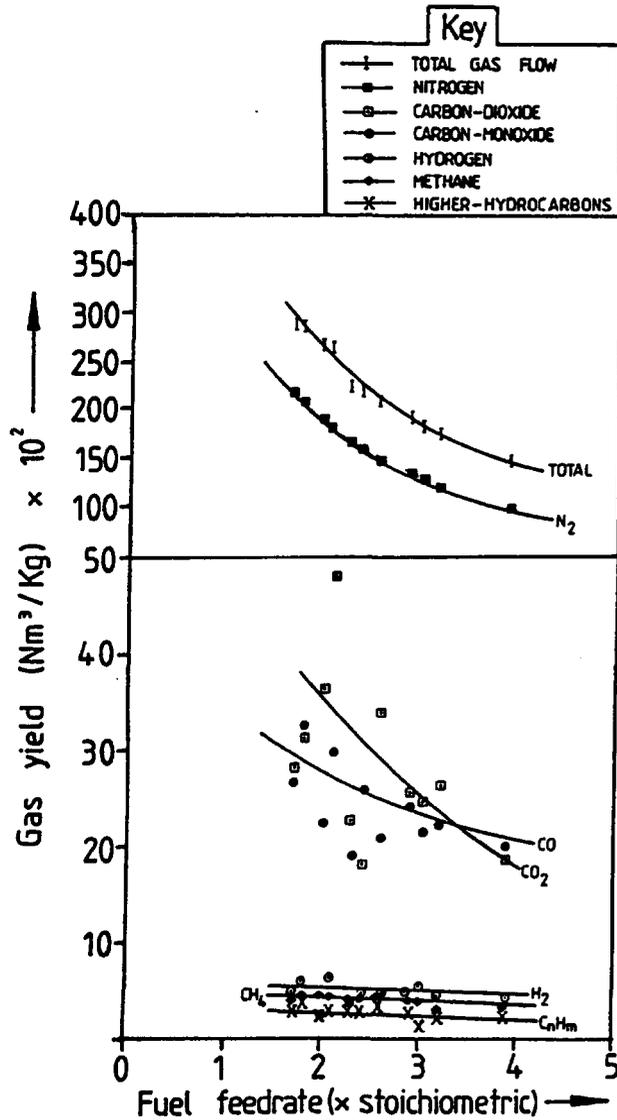
FIGURE 3·20 (b)

Bed depth = 10 cm, Temperature =941–990 K
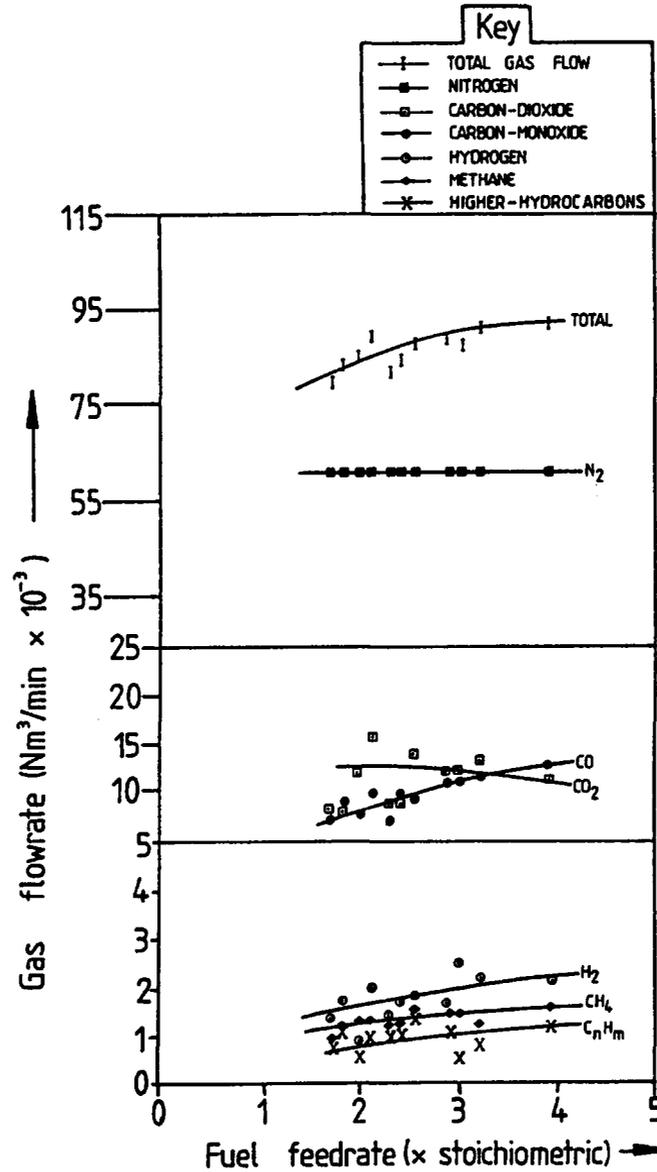
FIGURE 3·20 (c)

Bed depth = 10 cm
Temperature range = 941-990 K

Key

	TOTAL GAS FLOW
	NITROGEN
	CARBON-DIOXIDE
	CARBON-MONOXIDE
	HYDROGEN
	METHANE
	HIGHER-HYDROCARBONS

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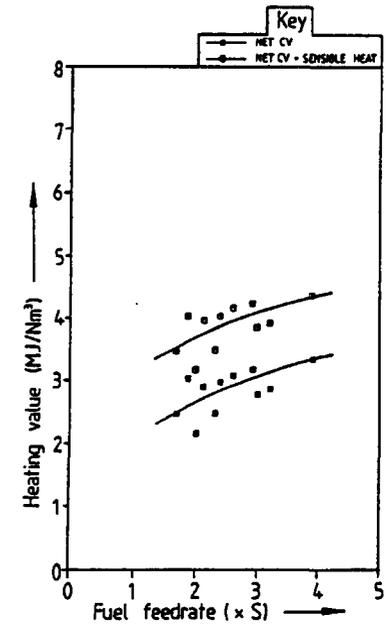
FIGURE 3·21 (a)

Bed depth = 10 cm, Temperature = 991 – 1020 K
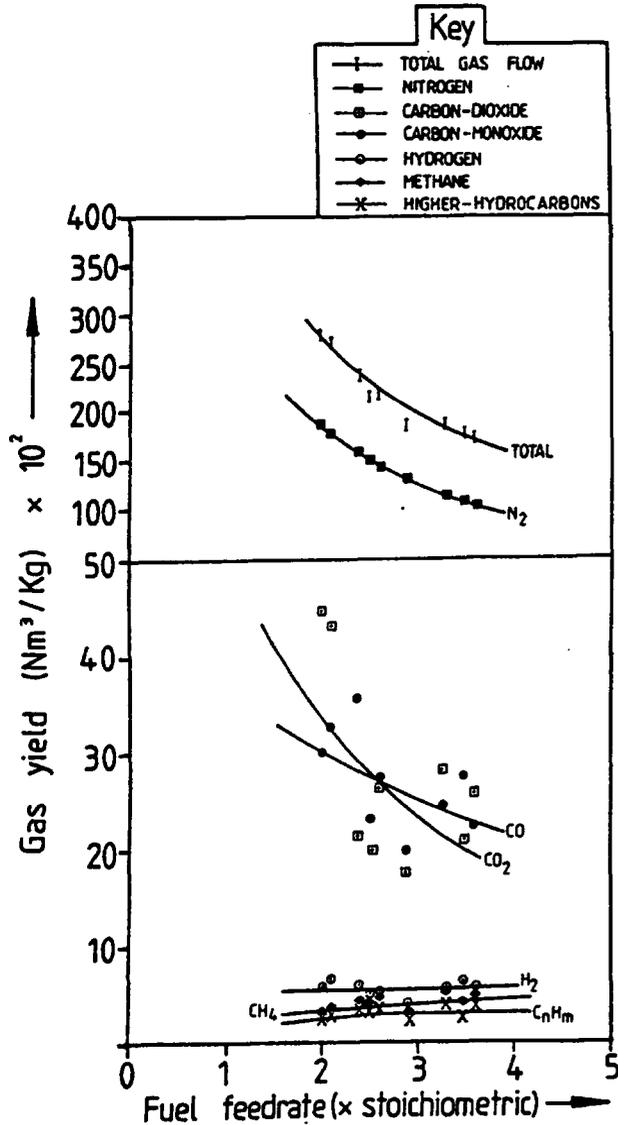
FIGURE 3·21 (b)

Bed depth = 10 cm, Temperature = 991 – 1020 K
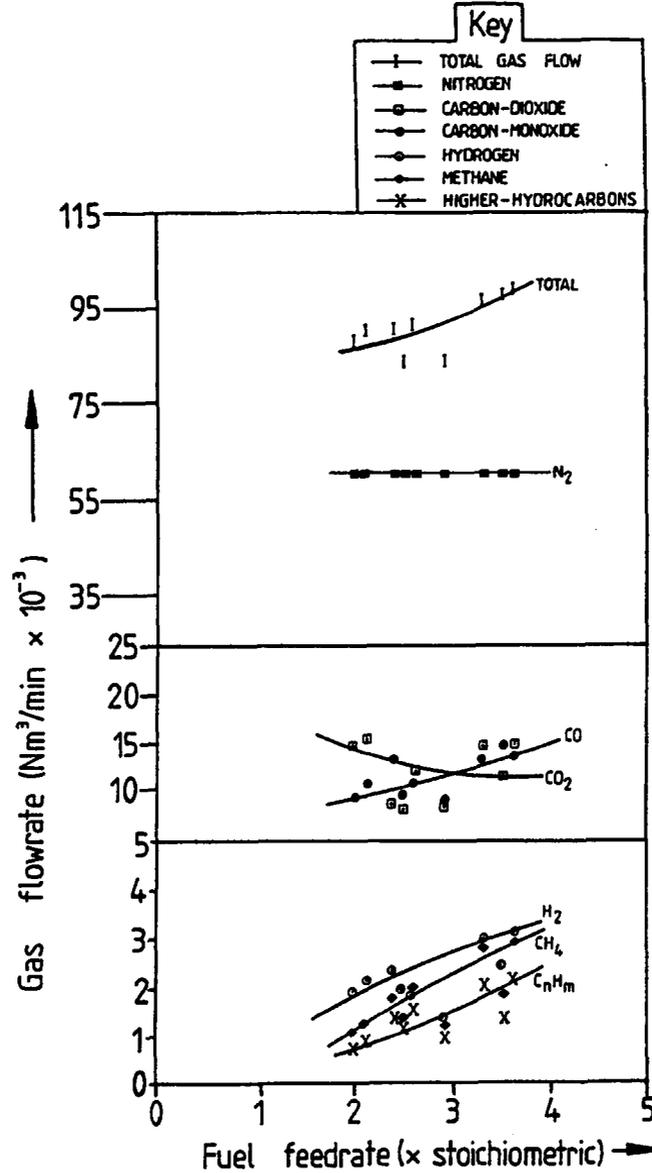
FIGURE 3·21 (c)

Bed depth = 10 cm

Temperature range = 991 – 1020 K
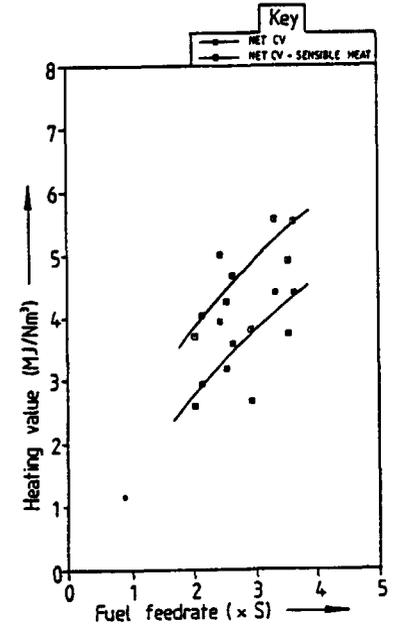
FIGURE 3·22 (a)

Bed depth = 10 cm, Temperature =1021–1080 K
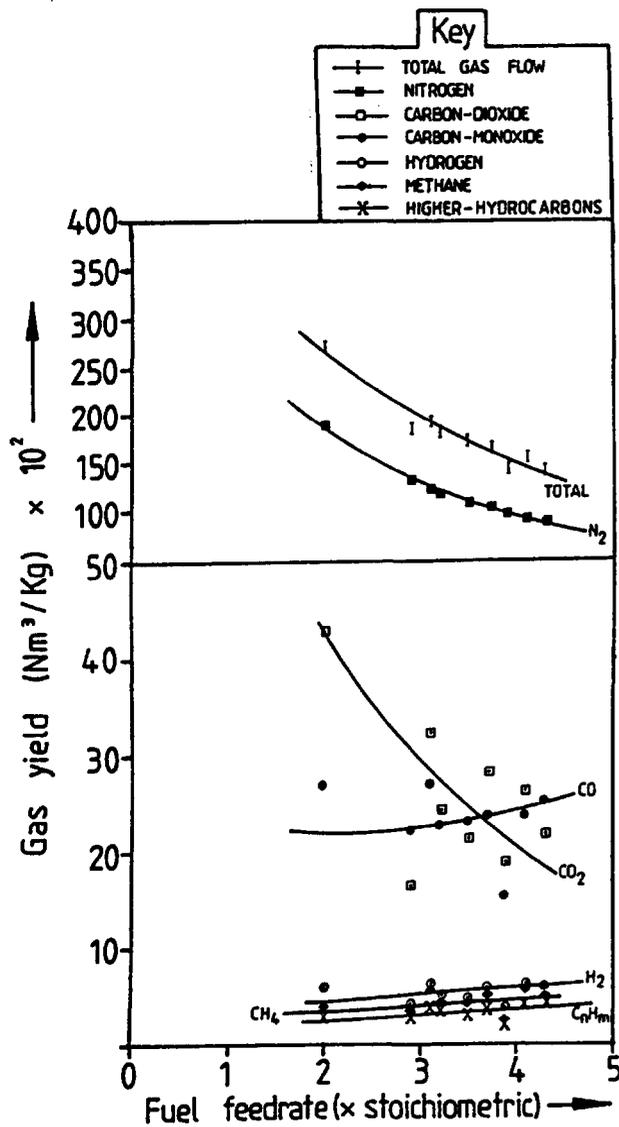
FIGURE 3·22 (b)

Bed depth = 10 cm, Temperature =1021–1080 K
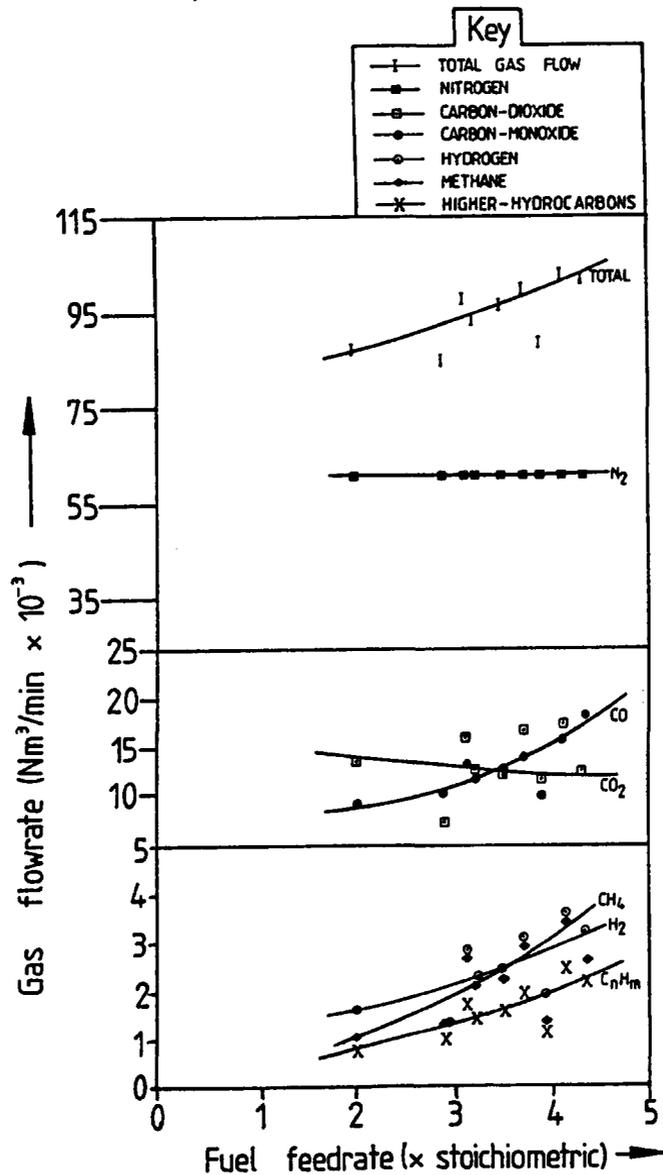
FIGURE 3·22 (c)

Bed depth = 10 cm
Temperature range =1021–1080 K
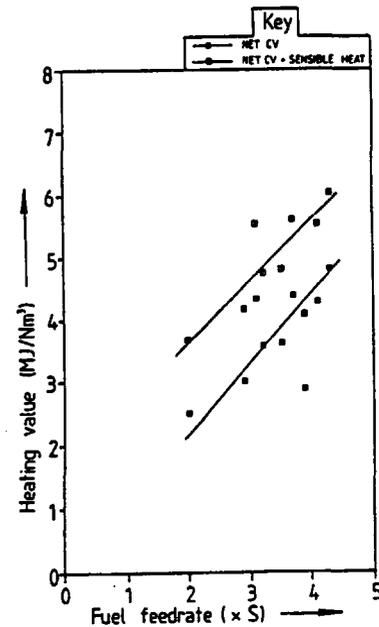
FIGURE 3·23 (a)

Bed depth = 5 cm, Temperature = 900-1060 K
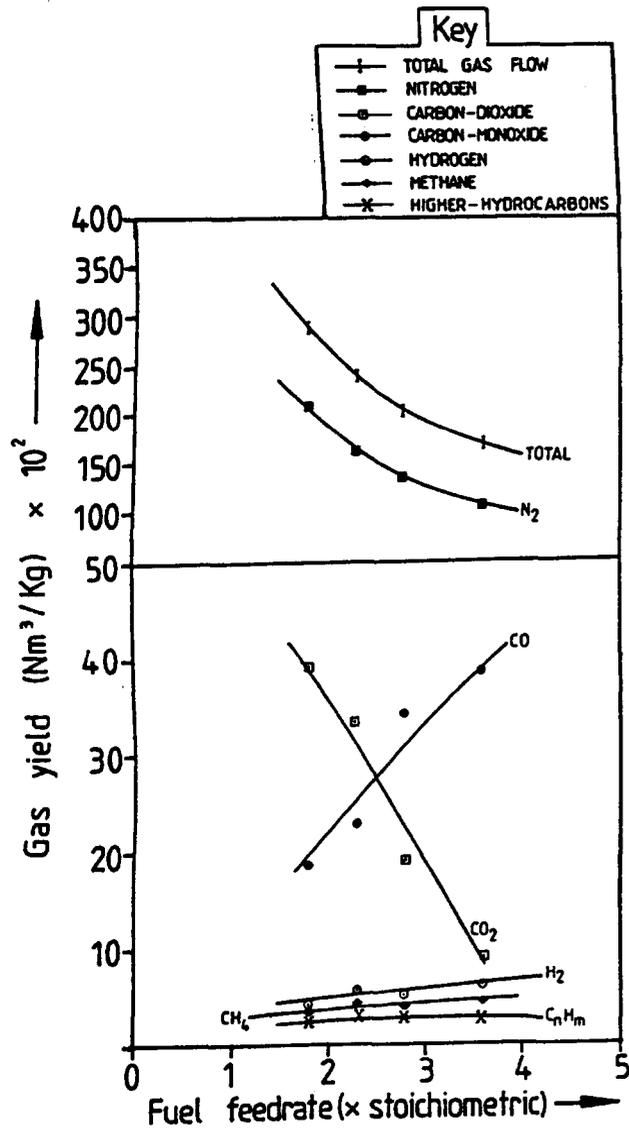
FIGURE 3·23 (b)

Bed depth = 5 cm, Temperature = 900-1060 K
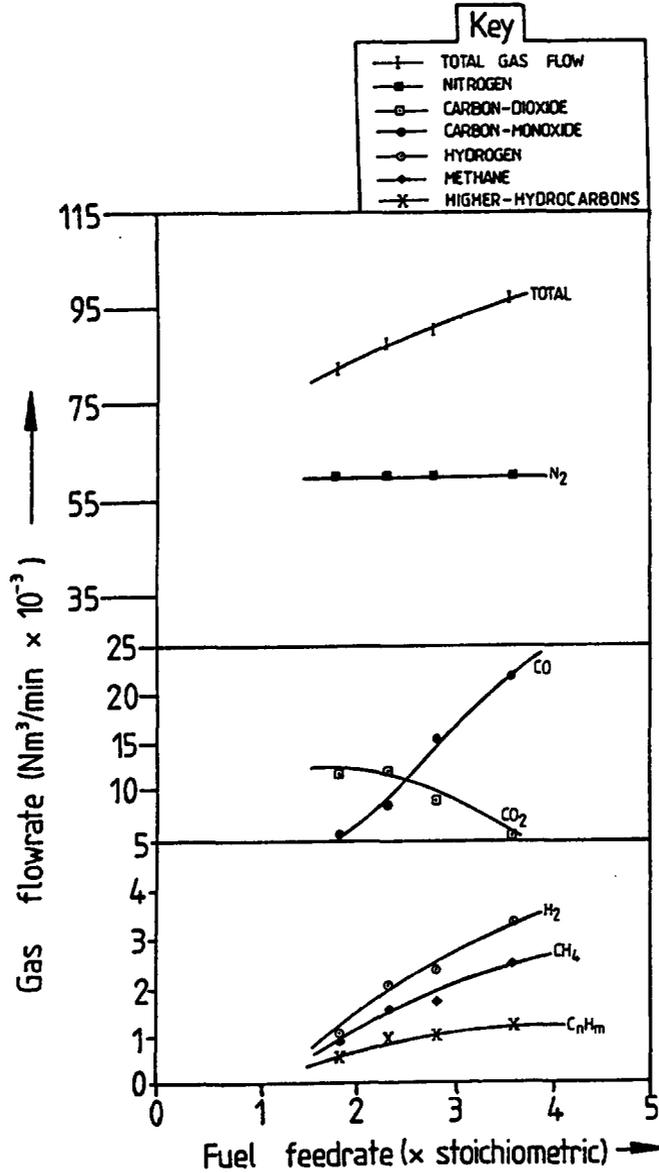
FIGURE 3·23 (c)

Bed depth = 5 cm
Temperature range = 900-1060 K

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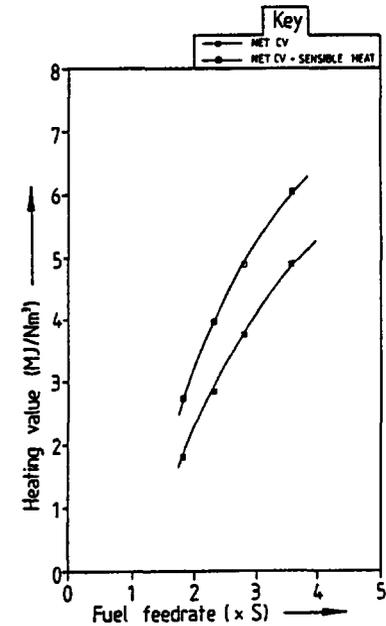
FIGURE 3·24 (a)
Bed depth = 20cm, Temperature = 801-900 K
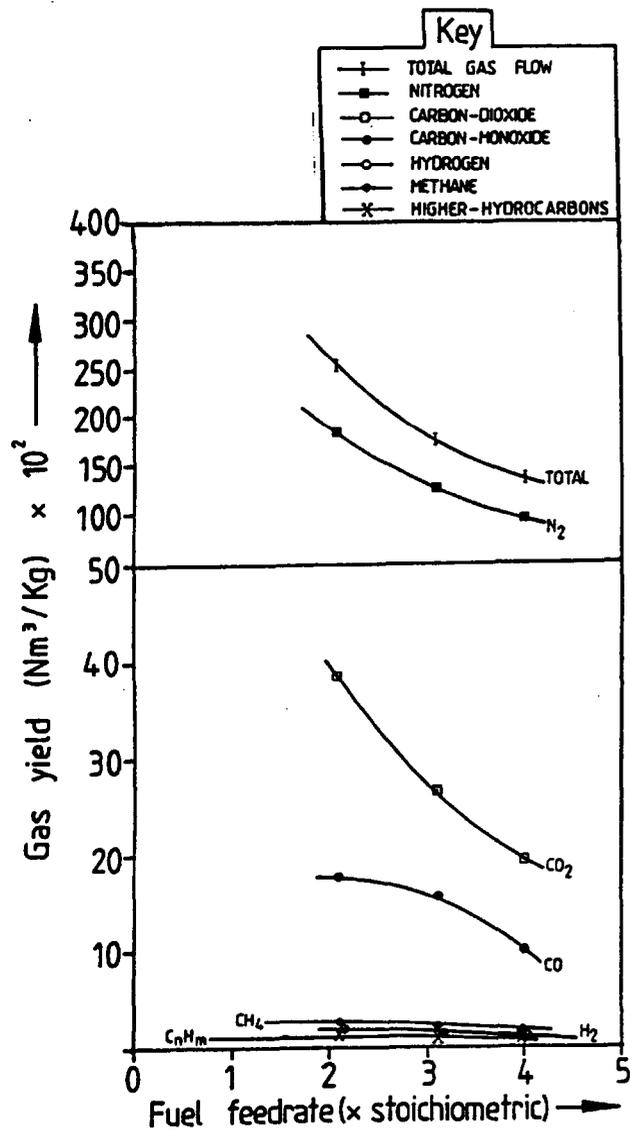
FIGURE 3·24 (b)
Bed depth = 20 cm, Temperature =801-900 K
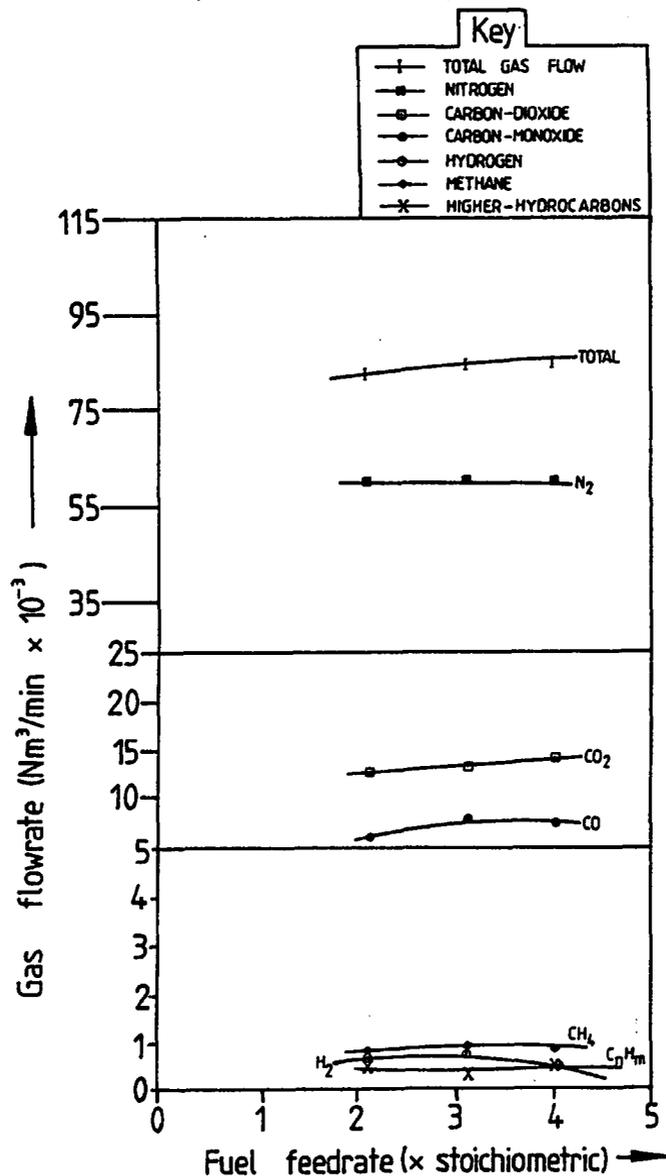
FIGURE 3·24 (c)
Bed depth = 20 cm
Temperature range =801-900 K
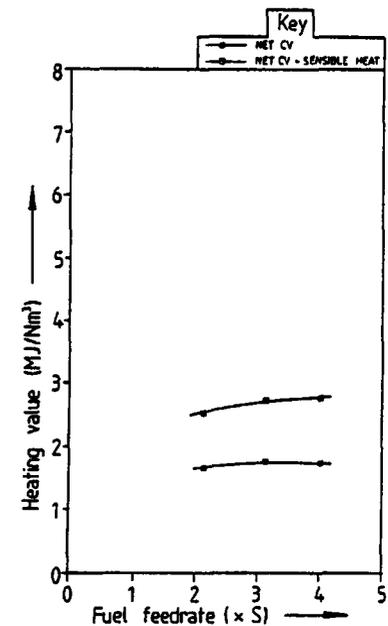
FIGURE 3·25 (a)

ed depth = 15 cm, Temperature = 951 – 1050 K
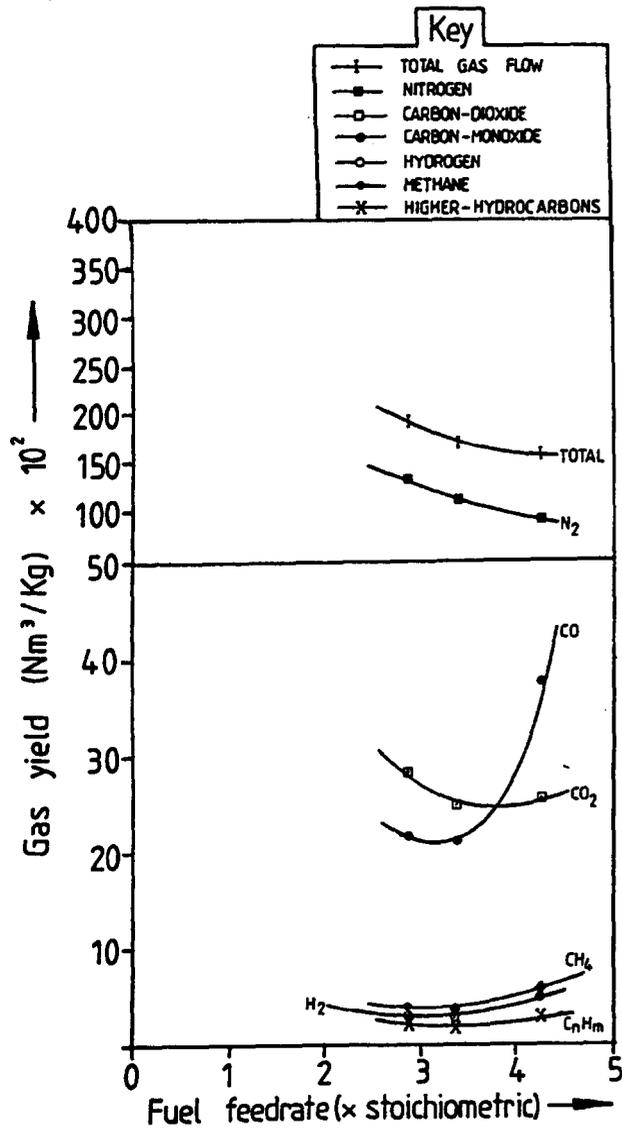
FIGURE 3·25 (b)

Bed depth = 15 cm, Temperature =951-1050 K
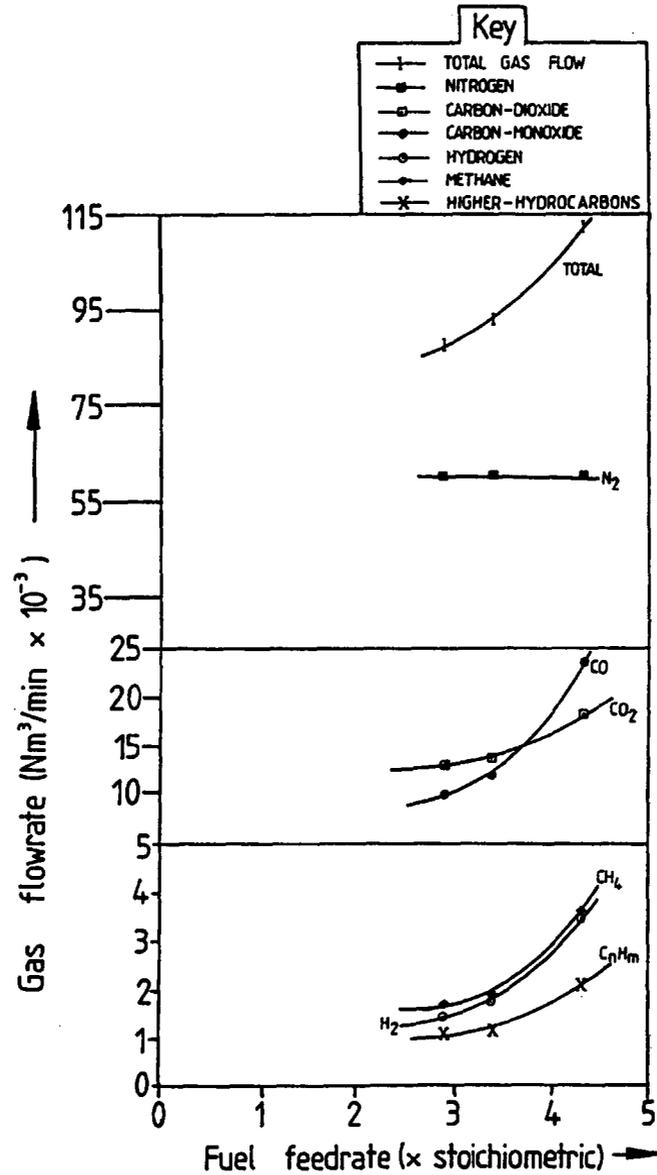
FIGURE 3·25 (c)

Bed depth = 15 cm

Temperature range =951-1050 K
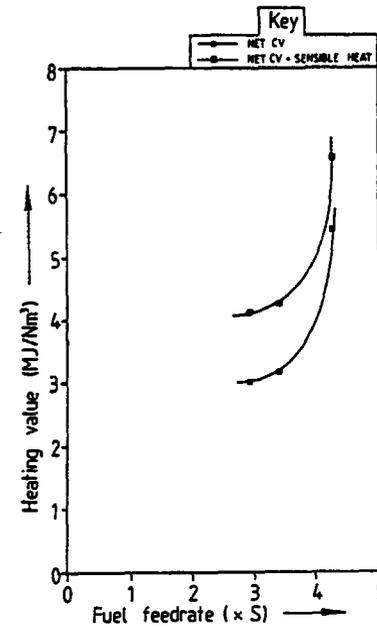
FIGURE 3·26

Gasification Efficiency (%)
Calorific Value (MJ/Kg Fuel) Gross
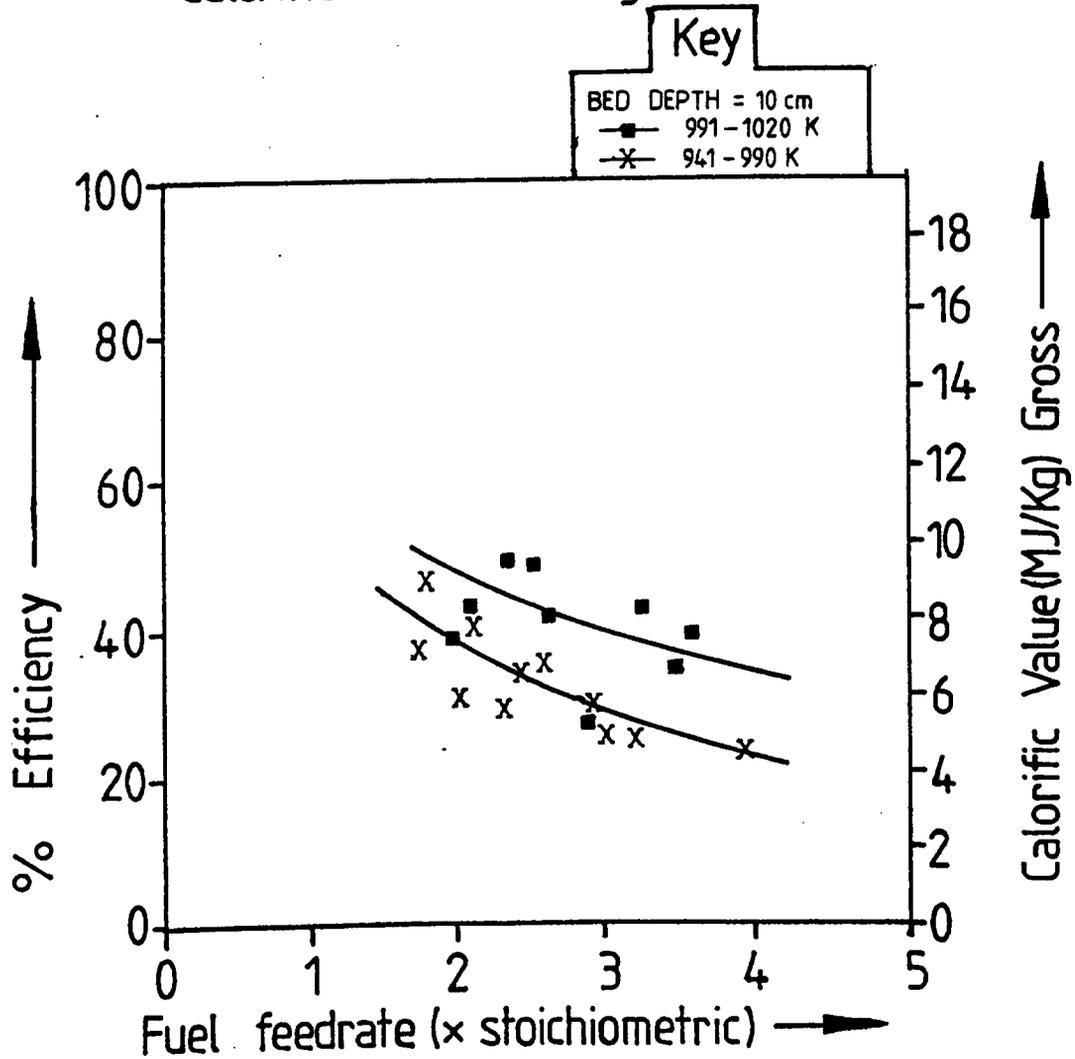
3.4.4.2 Effect Of Temperature

The relative output of gas is extremely dependent on the cracking temperature employed in the reactor. Higher cracking temperatures (> 900 K) tend to give higher yields of gas and less tar, whereas the opposite is true for lower temperatures [VAN DEN AARSEN et al., 1982 and MORI, 1979]. A fluidised bed of hot sand particles, therefore, has the useful characteristic of rapid heat transfer to the fresh fuel particle.

Figures 3.27 and 3.28 (a-b) show how temperature influences the gas output (Nm^3/min) and calorific value (MJ/Nm^3), (a) and (b) respectively.

It is noticable that the total gas output shows an increase with increasing temperature. Since the output of the major constituent of the gas, N_2, remains constant for particular fuel feedrates. Then the increase in the total gas output must be attributed to the increase in the output of the other gaseous components.

The general trend is that there is an higher output of the combustible components (H_2, CO, CH_4 and C_nH_m), but a levelling off in the CO_2 output. This is probably due to the effects of the reduction of CO_2 by carbon to form CO, which is favoured at higher temperatures as indicated by the following equation :

$$C \ + \ CO_2 \ \text{-------} \text{>} \ 2CO \qquad\qquad [3.8]$$

FIGURE 3·27 (a)

Bed depth = 10 cm, Fuel feedrate (×S) = 2·3–2·6

Key

- TOTAL GAS FLOW
- NITROGEN
- CARBON-DIOXIDE
- CARBON-MONOXIDE
- HYDROGEN
- METHANE
- HIGHER-HYDROCARBONS
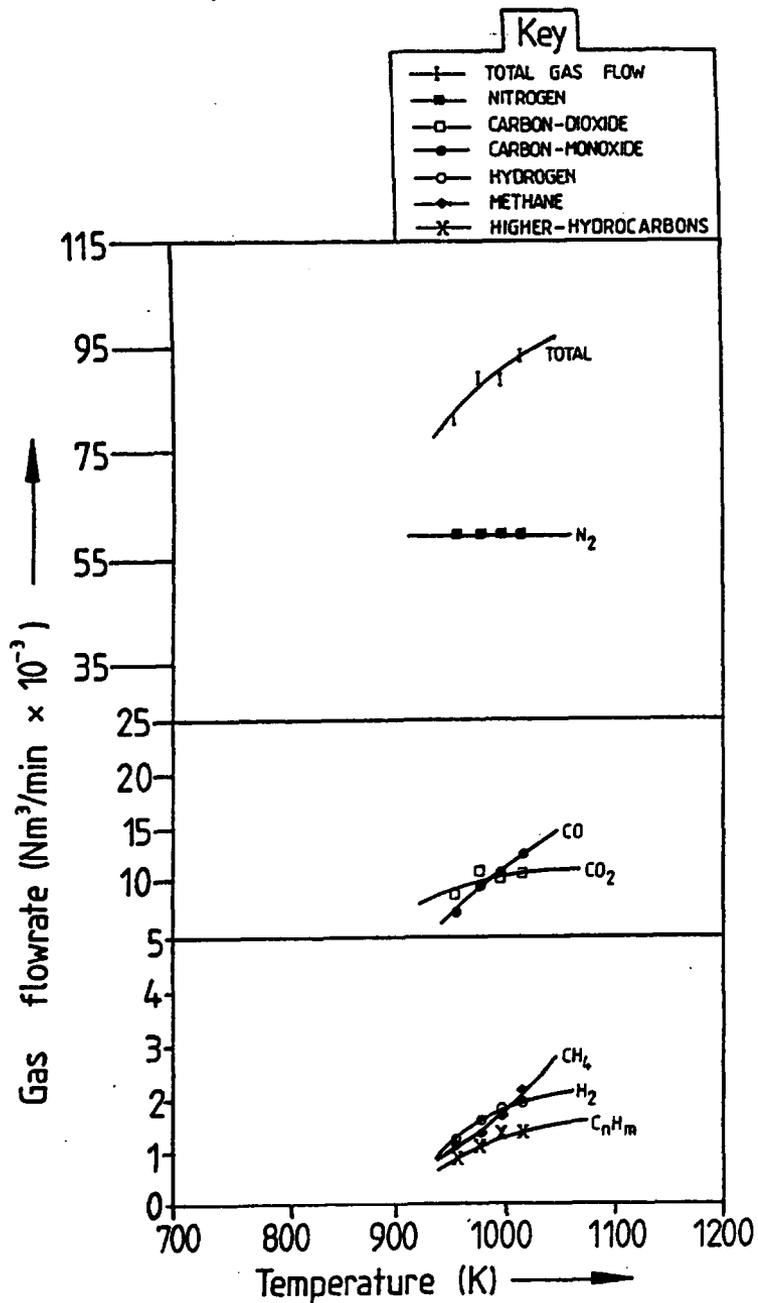
FIGURE 3·27 (b)

Bed depth = 10 cm

Fuel feedrate (×S) = 2·3 – 2·6

Key

- NET CV
- NET CV + SENSIBLE HEAT
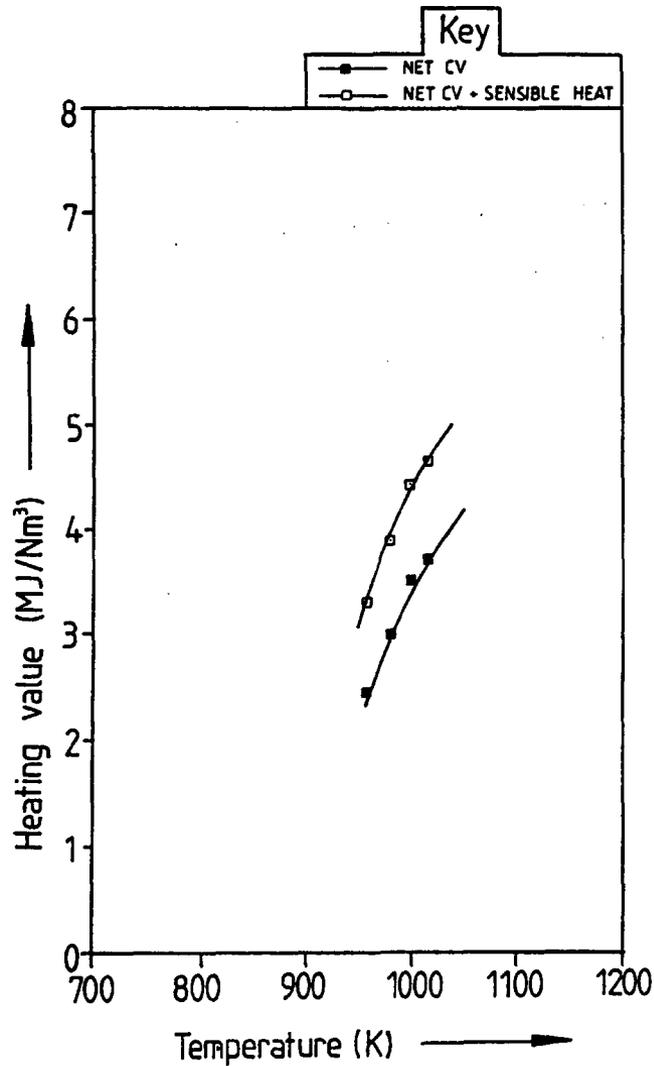
FIGURE 3·28 (a)

Bed depth = 10 cm, Fuel feedrate (× S) = 3·0 - 3·3

Key
TOTAL GAS FLOW
NITROGEN
CARBON-DIOXIDE
CARBON-MONOXIDE
HYDROGEN
METHANE
HIGHER-HYDROCARBONS
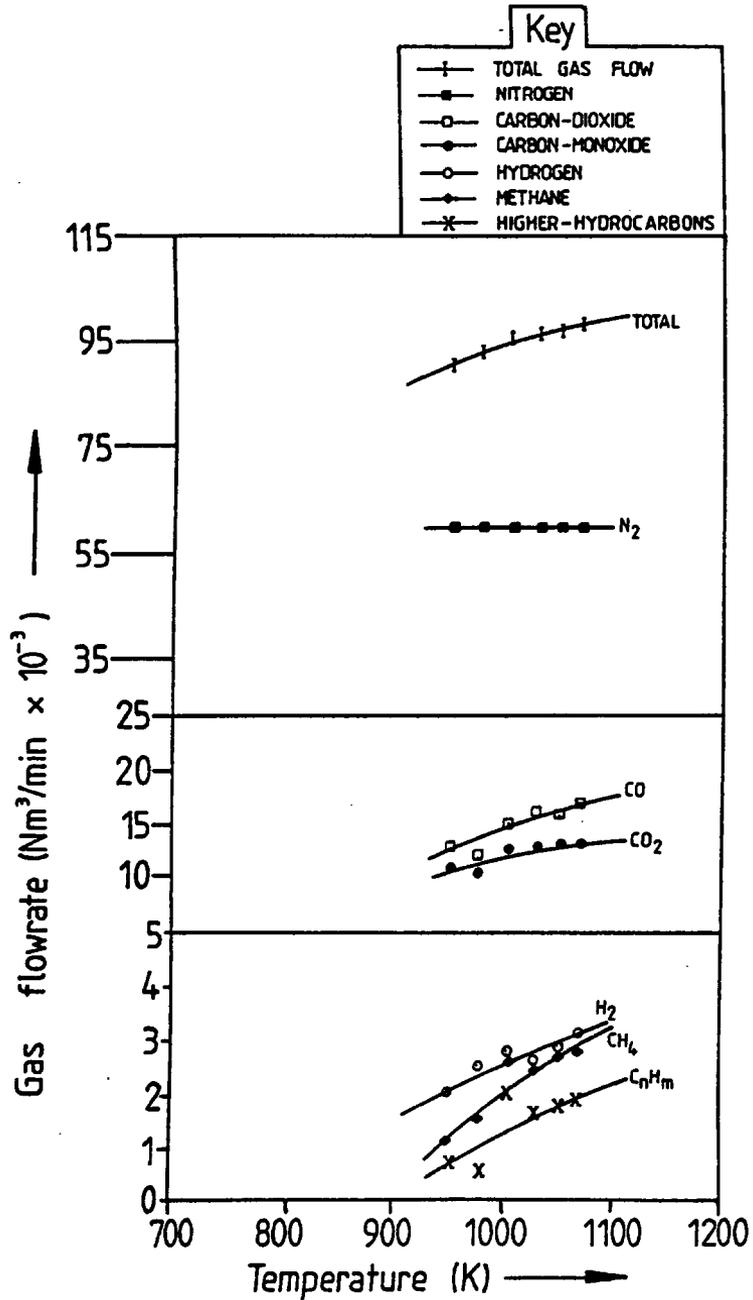
FIGURE 3·28 (b)

Bed depth = 10 cm

Fuel feedrate (× S) = 3·0 − 3·3

Key
NET CV
NET CV + SENSIBLE HEAT
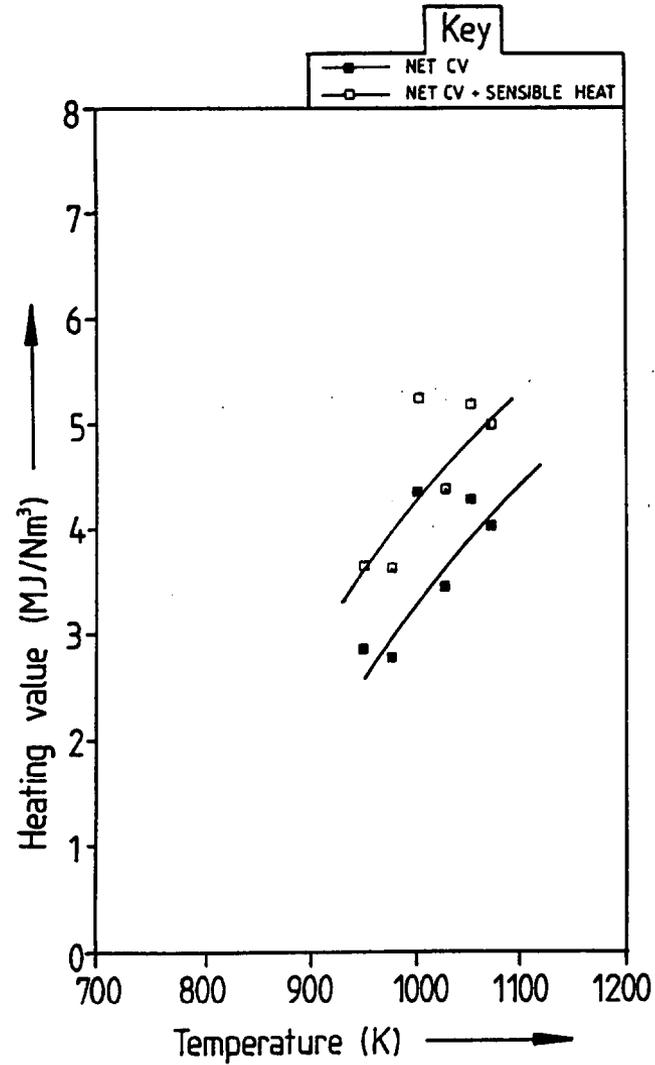
The increase in H_2 output can be explained by the more complete pyrolysis of the tar and in addition by the char-steam reaction at higher temperatures :

$$C \ + \ H_2O \ \text{--------} \rightarrow \ CO \ + \ H_2 \qquad [3.9]$$

This reaction also produces CO. The increase in CH_4 and other higher hydrocarbons can be attributed to the more complete pyrolysis of tar.

By performing mass balance calculations from fuel ultimate analyses and gas analyses the efficiency of conversion of elements (H, C, O) from solid to gas can be calculated. Figure 3.29 (a and b) gives all the efficiencies, including thermal and gasification, with respect to temperature. This shows a general increase in coversion efficiencies with increasing temperatures.

The increase in the gasification efficiency with increase in temperature is due to the higher proportion of combustibles in the fuel gas from increased yields of pyrolysis and gasification products. Figures 3.27 and 3.28 (b) show how the calorific value of the product improves with increased temperatures. The Figures also shows the improvement to the calorific values if the sensible heat were to be taken into consideration.

FIGURE 3·29 (a)

Bed depth = 10 cm, Fuel feedrate (×S)=2·3-2·6

Key
—|— % CARBON EFFICIENCY
—□— % OXYGEN EFFICIENCY
—○— % HYDROGEN EFFICIENCY
—●— % THERMAL EFFICIENCY
—X— % GASIFICATION EFFICIENCY
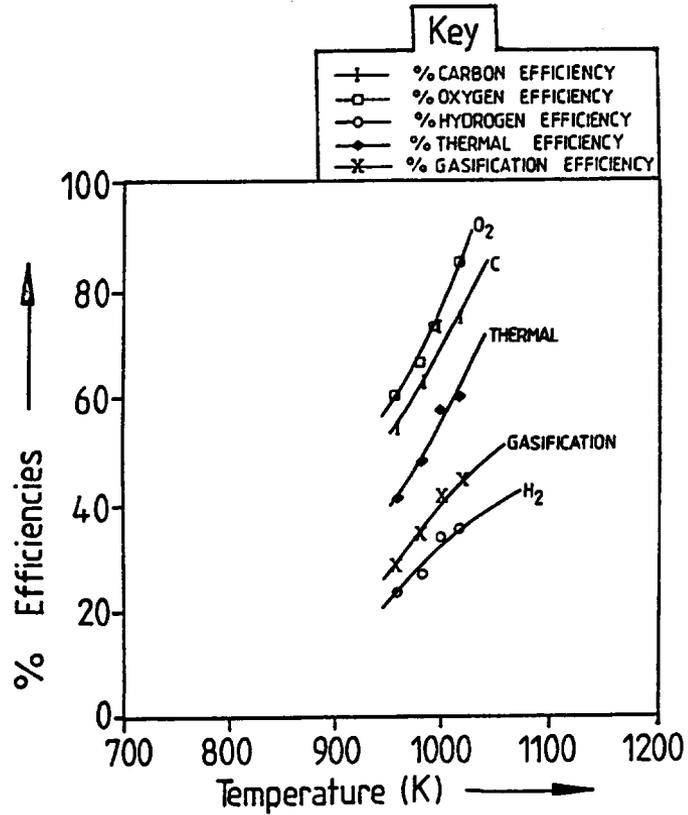
FIGURE 3·29 (b)

Bed depth = 10 cm, Fuel feedrate (×S)=3·0-3·3

Key
—|— % CARBON EFFICIENCY
—□— % OXYGEN EFFICIENCY
—○— % HYDROGEN EFFICIENCY
—●— % THERMAL EFFICIENCY
—X— % GASIFICATION EFFICIENCY
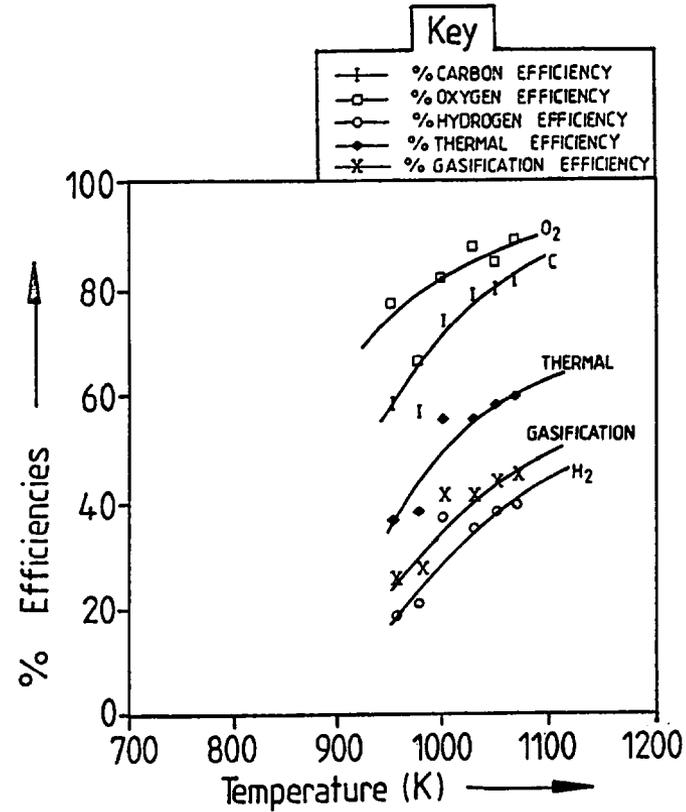
3.4.4.3 The Effect Of Bed Depth

The study of woodchip gasification and pyrolysis presented in this chapter included the effect of varying static inert bed depth along with the other main variables of bed temperature and fuel feedrate. Static bed depths of 5, 10, 15 and 20 cm were used and the results are presented in Figure 3.30 (a-d).

The general trend of the results showed that at deeper bed depths better quality fuel gases were produced. This is explained by the fact that there is a separation of the initial rapid pyrolysis of fuel particles occuring at the top of the bed from the combustion occuring in the lower parts of the bed. The combustible gases released during pyrolysis are therefore less likely to be burned in a deep bed than a shallow bed. The violent nature of solids recirculation in a bubbling fluidised bed brings fuel particles into the combustion region before complete release of volatiles has occured which inevitably means that CO, H_2, hydrocarbon gases and vapours will br consumed but to a lesser extent in the deeper beds.

The factors which affect the amount of volatile combustion which takes place are :

(1) Rate of pyrolysis and volatile release which is dependent on reactor temperature.

(2) Rate of solid recirculation within the fluidised

bed which is dependent on fluidising velocity

(3)   Rate at which solid fuel particles are
      transferred from the bed surface to the
      combustion zone in the lower sections of the bed
      which is dependent on bed depth.

FIGURE 3·30 (a)

Fuel feedrate (×S) = 1·8 − 2·4

Temperature   range = 870 − 940 K
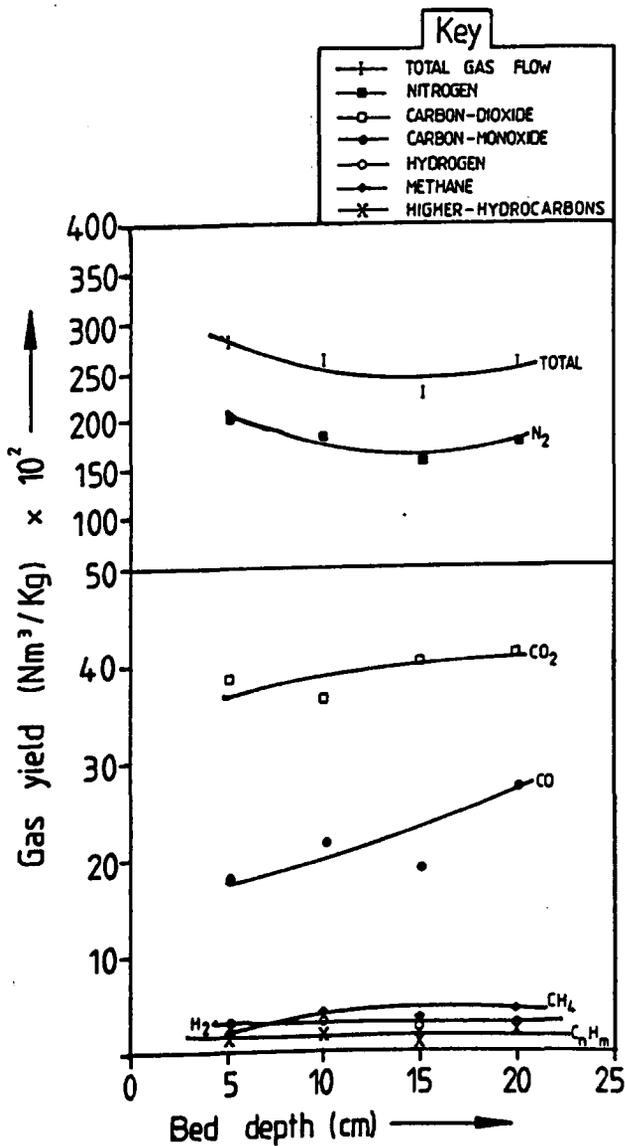
FIGURE 3·30 (b)

Fuel feedrate (×S) = 1·8 − 2·4

Temperature   range = 870 − 940 K
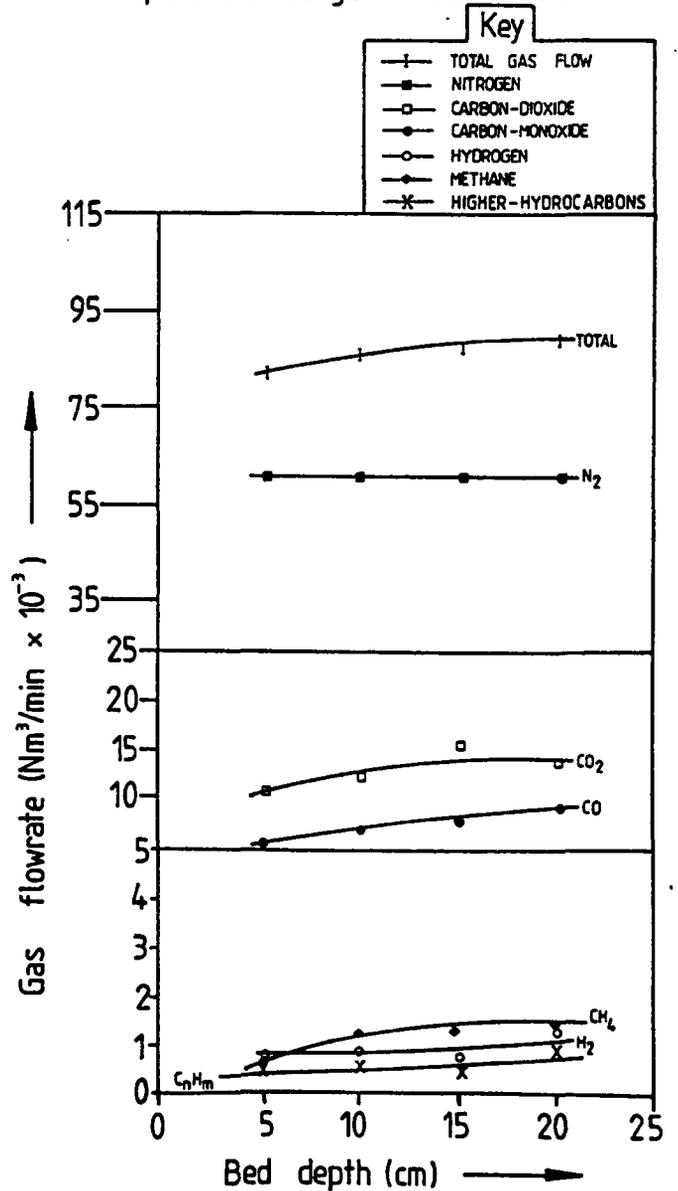
FIGURE 3·30 (c)

Fuel feedrate (×S) = 1·8 − 2·4

Temperature range = 870 − 940 K

Key
NET CV
NET CV + SENSIBLE HEAT

Heating value (MJ/Nm³)

Bed depth (cm)
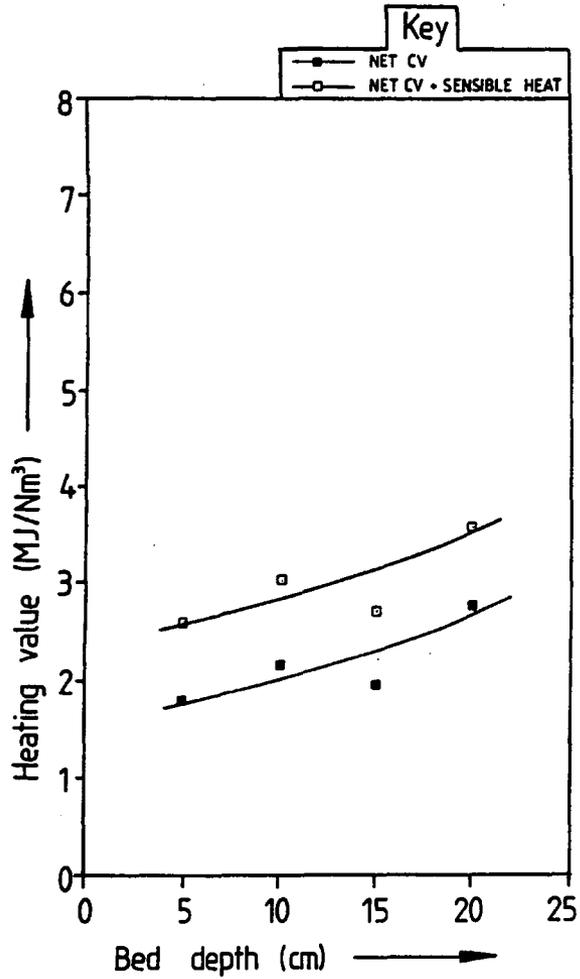
FIGURE 3·30 (d)

Fuel feedrate (×S) = 1·8 − 2·4

Temperature range = 870 − 940 K

Key
% CARBON EFFICIENCY
% OXYGEN EFFICIENCY
% HYDROGEN EFFICIENCY
% THERMAL EFFICIENCY
% GASIFICATION EFFICIENCY

% Efficiencies

Bed depth (cm)
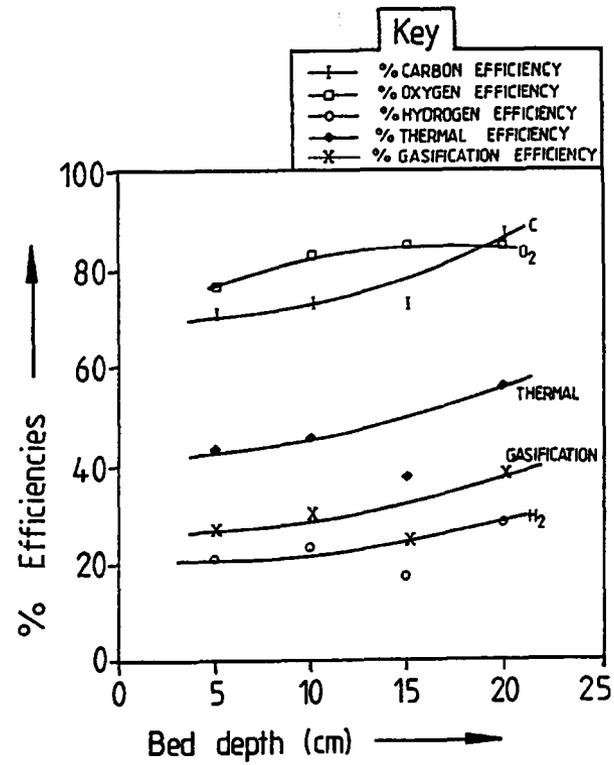
3.4.5 Losses Associated With The Partial

Combustion/Pyrolysis Of Woodchips

Losses to the woodchip-fuel gas coversion process can be accounted in thermal units and are reflected in the efficiency rating of the process. The sources of losses which will be discussed in this section are the following :

(1) Elutriation of carbonaceous material (including sensible heat).

(2) Sensible heat in product gas flow.

(3) Condensible combustibles, ie. oils and tars, in the product gas flow.

If the fuel gas can be consumed close to the point of production, then losses (2) and (3) can be turned to advantage by burning the hot gas laden with tar vapours. This can effectively increase the calorific value of the gas and will produce a more radiant flame. On the other hand, if the gas was consumed at a location remote from the producer, then it would have to be cooled and cleaned before transportation through pipework to prevent blockage which incurs losses to the efficiency of the system.

3.4.5.1 Solids Elutriation

Solid fuel particles, when exposed to conditions of combustion at high temperatures in a fluidised bed, tend

to reduce in size. The burning fuel particles shrink due to consumption of carbon and also undergo processes of attrition and breakage. If we consider a fresh fuel particle entering the hot fluidised bed then rapid volatile release will leave a fragile porous char particle prone to abrasion from other bed particles. The rapid rise in temperature may also cause the particle to break up as it enters the bed. The on going process of particle size reduction in the bed will cause a distribution of particle sizes to occur with the lower limit being controlled by fines having terminal velocities less than the velocity of the upward flowing air, and combustion/pyrolysis products will be carried out of the reactor [WALL, 1974]. If the fines contain carbon and/or volatile combustibles then, if the particles are not returned to the bed or burned elsewhere, this represents a loss to the process. An assessment of the losses have been presented in Tables 1 to 15 (d), Appendix 2, for all the experiments. Figure 3.31 shows how increasing the fuel feedrate increases the rate of solids carry-over with effectively a linear dependency at the operating conditions stated. Over the range of feedrates studied, the amount of solid carried over, as a percentage of fuel input, remained constant as shown in Figure 3.31.

The results indicate that elutriation of fines comprises about 0.6% of the fuel input to the reactor and

FIGURE 3·31

Bed depth = 10 cm
Temperature range = 1021−1040 K
Stoichiometric fuel feedrate = 16·0 g/min
Moisture = 11·0%
U_0 = 0·25 m/s (1000 K)
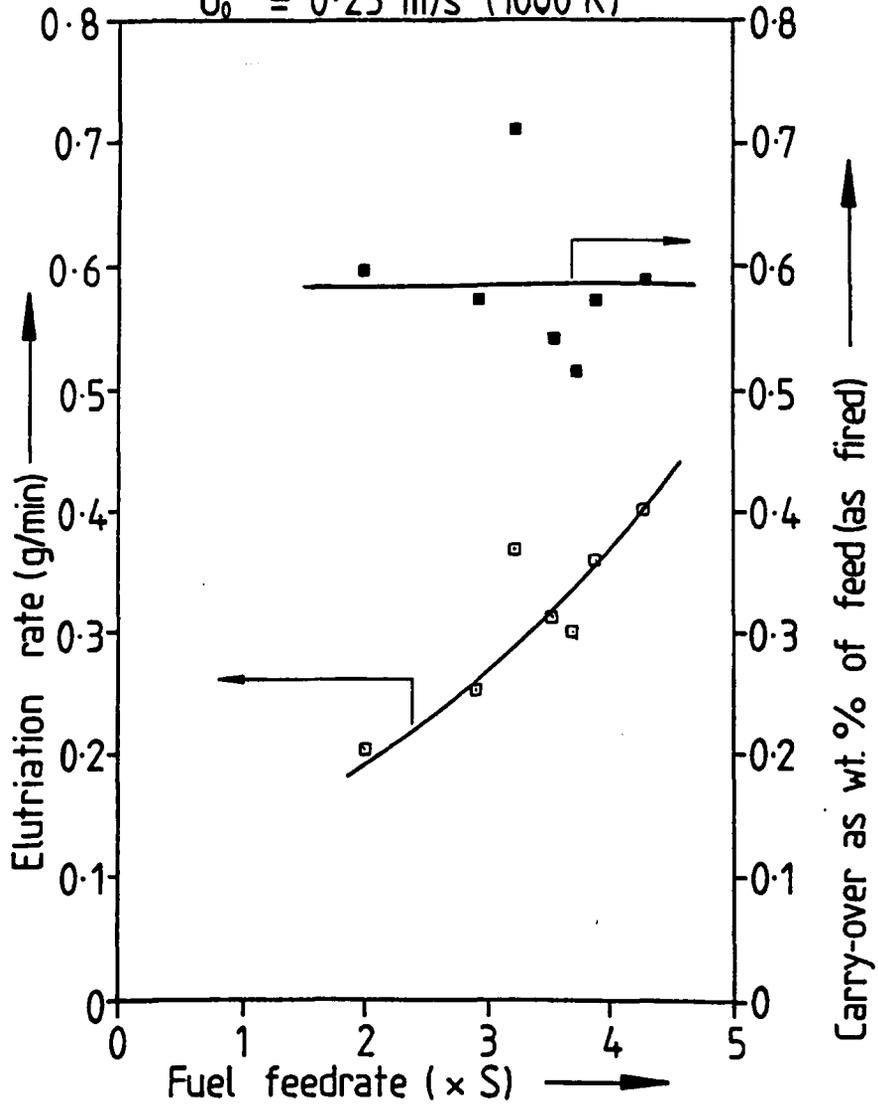
is directly proportional to the rate of fuel feeding. An
average proximate analysis of the elutriated material is
given in Table 3.4, where it is clear that the elutriated
solids are only slightly lower in carbon and hydrogen
content compared to the original fuel. Only the ash
content is greatly increased in comparison.

The potential heat in the carry-over amounts to
0.09 MJ/Kg of woodchips (as fired). In other words, 0.5%
of the potential heat or CV of the woodchips is not
utilised in the reactor.

A major factor in determining the rate of
elutriation is the rate at which the fluidising air passes
through the reactor. It is clear that at higher upward gas
velocities more particles of larger diameters will be
elutriated giving a greater overall elutriation rate with a
wider size distribution. At lower fluidising velocities
only the finest particles are elutriated, thereby reducing
carry-over losses.

TABLE 3·4

Average Proximate Analysis

(% by weight)

	Carry-over	Fuel (daf)
Carbon	43·6 %	53·8 %
Hydrogen	3·1 %	6·7 %
Oxygen	19·6 %	37·1 %
Nitrogen	0·5 %	1·9 %
Ash	33·2 %	0·5 %
Heating Value	15·68 MJ/Kg	19·11 MJ/Kg

3.4.5.2 Sensible Heat

Sensible heat has been evaluated for all the experimental runs and presented in all the calorific value (net) figures to compare the additional potential. This potential can only be utilised if the product gas is being used close to the point of production, or the product gases are being cooled via a heat exchanger.

3.4.5.3 Condensible Combustibles

The condensible combustibles are tars and oils produced during the pyrolysis stage. Previous studies [VAN DEN AARSEN et al., 1982, SAKODA et al., 1981] have shown that greater than 90% of the condensible combustibles are broken down at temperatures exceeding 950K. The oil and tar products can therefore still be regarded as a significant product of the process. It could also be a useful component of the fuel gas if it is led directly to a combustion chamber while still hot where the organic olis and tars are burned together with combustible gas components. The combustion of the oil and tar vapours tend to produce a more luminous flame compared to a pure carbon-monoxide and hydrogen flame, giving enhanced radiative heat transfer to refractories or water tubes.

3.5 SUMMARY AND CONCLUSIONS

The work presented in this chapter into the possibilities of low grade fuel gas production from woodchips has shown that a useful lean gas in high yield can be obtained using air. The most effective conditions for the production of the best quality fuel gas would be the following guidelines :

(1)   The maximum reactor temperature possible within limits of rate of heat transfer and construction materials but without air-preheat (which would incur an efficiency penalty) possibly up to 1200 K.

(2)   A rate of solid recirculation which for a given bed depth provides maximum heat transfer from the combustion zone at the bed bottom, to the pyrolysis zone at the bed top. A bed operating in a vigorously bubbling mode at between 3 to 6 times the minimum fluidising velocity would be satisfactory.

Woodchips and other forms of wood wastes have the potential, therefore, to be an important industrial fuel in countries where there is an excessive natural resource or where presently it is being consumed inefficiently.

Other studies which might play a part in any
further work would be :

(1) The yield and characterisation of the oils and
tars formed during pyrolysis/gasification with a
view to assessing by-product recovery.

(2) The processing of woodchips with steam, steam/air
or steam/oxygen mixtures for the production of a
synthesis gas, rich in carbon-monoxide and
hydrogen.

(3) Employing greater bed depths and feeding fuel
within the bed.

CHAPTER FOUR

COLD MODELLING

4 COLD MODELLING

4.1 INTRODUCTION

A large number of fluidised bed applications require the bulk transfer of fluidised solids. The fluidised bed offers a number of advantages over most methods of contacting, such as high rates of heat and mass transfer and solids mobility. A recirculating fluid bed (incorporating a centralised jet and draft tube or two jets) has additional advantages by controlling the recirculation of fluidised solids and improving the solids handling. In these systems, given process conditions, it is important to be able to predict the recirculation rates of the bed material to ensure circulation is obtained throughout the bed. Tests carried out in cold 2-D fluidised beds have been found useful in identifying the critical design parameters. The merits of cold 2-D beds will be discussed later.

Gas-solid fluidisation was first employed as an industrial technique (coal gasification) in the Winkler Process in the early 1930's. The chemical reactions involved in the gasification processes are all heterogeneous gas-solid reactions accompanied by large

energy changes.   Heat and mass transfer considerations are an essential criterion in the design of such processes. Solid fuel gasification processes are classified on the basis of the technique by which heat transfered.   Within each group will then contain processes which are similar as regards their energy utilisation, their range of application and the composition of the gas produced.   As a rough criterion we can distinguish between allothermic and autothermic processes.

Fluidised bed gasification processes are applicable to both the above applications.   The adaptability of fluid beds to varying loads without serious ill-effects on the thermal efficiency of the process, due to the special flow properties of the turbulent layer, coupled with the very homogeneous temperature distribution, is the reason why the fluid bed technique is employed in most gasification processes.

Further improvements to the basic fluidised bed technique have been made in recent years by combining two fluid beds.   In these systems combustion and gasification processes are isolated, thus yielding a higher quality gas ie. not diluting the fuel gas with combustion products. However, there is a slight problem in this technique concerned with the transference of heat from the exothermic combustion chamber to the endothermic gasification chamber. This is presently being investigated by many researchers.

The capital cost of twin-fluidised bed systems would also be higher. A compromise between the quality of the fuel gas and the capital costs involved can bring into contention a single fluidised bed with internal recirculation, known as a recirculating fluidised bed.

The concept of recirculating fluid bed with a draft tube is by no means new. Probably the first application was described by TASKAEV and KOZINA (1956). They utilised this system for low temperature carbonisation of coals in a 15cm diameter column with a 2.5cm diameter draft tube. Later a 'seeded coal process' was developed by CURRAN et al. (1973) employing the same concept to smear the 'liquid' raw coal undergoing the plastic transition onto the seed char, and the recirculating char during low temperature pyrolysis. Further studies performed by WATKINSON et al. (1983) on spouted beds with jets, have reported two major advantages of this system over conventional fluid bed systems, when using high caking coals. There is no grid upon which agglomerates can build up and since all the gas enters through a single orifice the resultant high velocity jet at the apex of the cone would tend to break up any agglomerates that might form within the bed. The draft tube with jets or two jet system, although not having a single inlet orifice, has a similar flow pattern to the spouted bed. It would therefore follow that the same advantages would be

applicable to these systems.

However, one of the largest concerns when using reciriculating fluidised beds, to commercialise many chemical processes, is scale-up. This is believed to be due to the absence of an experimentally verified hydrodynamic theory. The outcome of the interest shown in recent years in understanding these systems has led to several hydrodynamic studies being undertaken in this field. The present study is therefore intended to supplement other work in this field, but in particular to present a viable proposition for its use in biomass gasification processes with respect to the findings from woodchip gasification. There has also been work done in this present study using a novel two jet system instead of draft tubes and comparison of induced circulation rates between the systems is made.

4.2 HYDRODYNAMIC STUDIES

Hydrodynamic studies of recirculating fluidised beds were initiated by YANG and KEAIRNS (1974,1978). They identified the design parameters on the basis of the data collected in a semi-circular plexiglass column 6.1m in

height and 0.3m in diameter, using two different
distributor plates (flat and conical) and two sizes of
draft tubes. Solid circulation rate in their studies was
obtained from the particle velocity measurements at the
downcomer side, either by following visually the tracer
particles at the wall with a stop watch or by using a
'radio pill'. A 'radio pill' is a radio transmitter
enclosed in a spherical or cylindrical plastic shell
between 1.3cm and 2.5cm in diameter, so that the density of
the radio pill approaches that of the bulk solids in the
downcomer side. The bed sand size used in their work was
of an average diameter of 0.75mm. From their studies they
could only obtain bulk flow measurements giving flow
patterns, since the radio pill employed was between 17 to
33 times larger than their bed particles. It would
therefore not be expected to flow individual particle
movements. The insertion of ariels in the bed to enable
tracking of the radio pill could also cause distortions in
the bulk particle flow patterns, but this might not be
visually observable. However, since the bed dimensions are
so large any small distortion may not be significant. The
main conclusions from their work was the prediction of the
bulk flow patterns, as shown in Figure 4.1.

Later work done by ISHIDA et al. (1975), observed
solids circulation in a fluid bed incorporating a draft
tube. They used a semi-cylindrical tube of 27.9mm internal

diameter and 2.8mm thickness as an inner column made of transparent PVC plate, which was attached by a perpendicular glass plate to an outer column of 280mm i/d. Porous alumina catalyst beads of 40 to 60 mesh (U_{mf} of 3.6cm/s) were used as solid bed particles and air was used as the fluidising gas. Their work only dealt with the concept of equilibrium bed heights and observation of solid circulation rates. No quantative measurements were made in their work regarding the solid circulation rates or particle velocities. They reached a similar conclusion on flow patterns as previous work.

FIGURE 4·1

Recirculation Bed Concept

Gas out

Solids out

Draft tube

Downcomer

Solid flow
Gas flow

Alternative solids feed

Gas distributor plate

Gas and solids feed
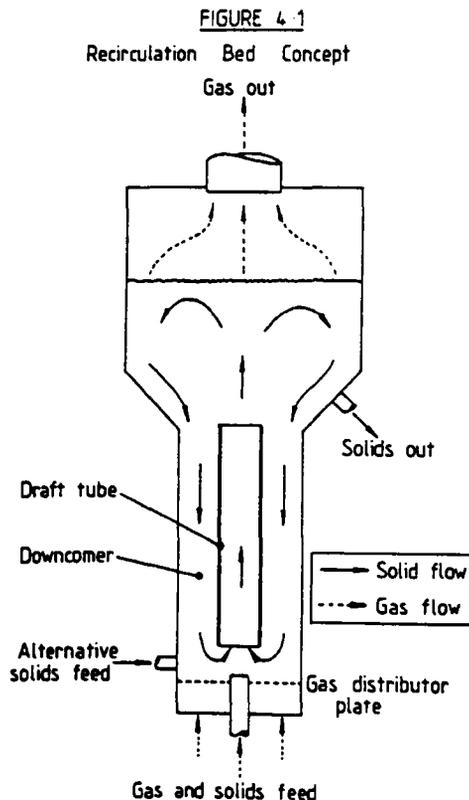
Further work was done by LANAUZE and DAVIDSON (1975), and LANAUZE (1976), who studied induced particle circulation rates in a 0.3m diameter air fluidised bed of sand with centralised draft tubes. They employed the same experimental technique as used by Yang and Keairns, ie. the radio pill technique. From their results they concluded that circulation rates were affected by the gap height between the distributor and the draft tube. They also stated that the circulation rates were not affected by the draft tube length or the height of the bed above the draft tubes. A model was developed to predict the circulation rates, assuming that the driving force for circulation was the density difference between the draft tube and the annulus, and the energy dissipated by particle shear at the walls.

Recent work by GIDASPOW et al. (1983) measured the time-averaged porosity profiles in 2-D recirculating bed, with a circular jet and a rectangular jet, by the use of a gamma-ray densitometer. They claimed to have discovered from their studies that a region of maximum porosity to exist at a finite length above the jet inlet in the case of the circular jet. The maximum length moved up the bed with increasing jet velocity. It was also shown that jet penetration heights for both the jets were close to each other. Latter part of their work was concerned with the development of an experimentally verified hydrodynamic

model to predict solid circulation rates around a jet in a fluidised bed gasifier.

In summarising, the work done on recirculating fluidised bed hydrodynamics has greatly increased recently but as yet there has not been a satisfactory model to predict the circulation rates applicable to the gasification of biomass. The purpose of this study has been to extend the understanding of the recirculating fluidised bed and present a reasonable practical approach to recirculating biomass gasifiers. Unfortunately this study's scope did not enable any experimental work to be undertaken using the recirculating fluidised bed.

4.3 TECHNIQUES FOR PARTICLE TRACKING

Cold 2-D beds are normally adopted for studying hydrodynamics of fluidisation, but the tracking technique is the most important aspect in relation to the reproducibility of the results obtained. In the past it is this that has led to limitations in many studies and has yielded conclusions which might not be totally accurate. It is therefore necessary to investigate the advantages and disadvantages of particular techniques with respect to the

obtainable results. In this section, preceding the experimental section, it is therefore necessary to give an indication of the requirments of a tracking technique and the limitations of the present experimental technique employed in this study.

The essence of an effective technique is that the information should be reproducable with sufficient ease and clarity related to the bed behaviour. All techniques are limited by the imformation produced with respect to particular aspects of fluidising behaviour. That is, they have strong points in one area only. This is almost inevitable when dealing with complex fluid dynamics existing in a fluidised bed. Ideally it would be preferable to simultaneously track particle and bubble velocities locally, determine bubble (or in more turbulent regimes; the void) sizes in a randomly changing, fluctuating system. Unfortunately this is not possible. It is necessary to develop techniques that permit a reasonable view of the events occuring by looking at either extremely specific (local) areas or rather generally at the bed. When considering specific points within a bed it is important to obtain sufficient data throughout the bed to produce statistically meaningful results. In an attempt to obtain data in quantity, to provide results of sufficient accuracy, some very sophisticated equipment has been developed with a price tag that fits accordingly.

In a general sense it would be useful to know the
qualities that are desirable in a technique investigating
fluidisation. This would provide a basis for comparing the
merits of different techniques. The following criterion
might be adopted (either as desirable or otherwise, as the
case may be) :

(i)   The 'make up' of the bed with respect to bed
particles. Are the particles realistic if used in an
industrial application ? (or could realistic particles be
used). Particle qualities would depend on : material type,
size/range, shape and density. Furthermore in the context
of the bed, voidage and density can be considered.

(ii) Bed dimensions . 2-D bed is limited as an
approximation to a 3-D bed, since boundary layer effects
are likely to be significant. Dimensions approaching
meters are commonly employed, especially in the case of 2-D
beds. Bed geometry does appear to play a significant role.

(iii) The type of fluid , either liquid or gas. Could
either be used? Usually techniques involving gas
fluidisation can be applied to liquid fluidisation but not
vica-versa. A realistic simulation requires gas
fluidisation. For a gas; bubble formation and movement,
tracking of bubbles, bubble size and shape, would be
considered. In addition it is important to consider these
parameters with regard to ease of distinction.

(iv) Bed internals . Does the technique require only a

free-bed or does it require other additional bed internals.
It is obviously important, economically and for comparison,
that the basic free-bed is able to handle numerous bed
internals for example draft tubes or jets.

(v)   Tracer particles . A small number of particles
are 'labelled' for this purpose, enabling a distinction to
be made form the bulk of the bed particles. However it is
obviously important to maintain the same particle
properties, or at least relatively close. In certain cases
because the tracked particle(s) are limited by the type of
material, size or shape required, it becomes necessary to
alter the bulk of the bed particles accordingly.

(vi) Non hazardous . In the case of radioactive and
X-ray techniques this could be a problem. To a far lesser
extent UV requires elementary precautions, that is a pair
of goggles. In one particular tracking technique (the
L.D.V. technique) the fluid required, as the tracking
medium, is an extremely volatile liquid, consequently a
severe fire hazard and a powerful solvent (a mixture of
benzyl and ethyl alcohol).

(vii) The equipment required to monitor the fluidised
bed or the labelled particles therein. For light sensing
techniques it is useful to make visual odservations
initially. In this case it is relatively easy to check
correct fluidisation is occuring and that the required data

is obtained. Considerations should be made for relative ease of handling and calibrations required (problems with dead time for X-ray and radioactive techniques are significant and can be a limitation). Data must be sufficiently accurate to be meaningful.

(viii) The speed with which data can be treated to provide results. This is generally much less of a problem with the advent of computers, if signals representing the data can be digitised, a computer program can be written.

(ix) Versatility of the technique for various applications. This might include diverse areas such as ; mixing and separation, heat transfer studies, recirculation studies etc.

(x) Inexpensive (generally dependent on simplicity). 2-D beds have a distinct advantage in this respect. Readily lending themselves to techniques involving light and hence camera or cine photography, even simple visual observation. Video equipment is also increasingly available (now sufficiently developed to provide high resolution pictures). This avoids developing film negatives and readily facilitates virtually automatic digitisation. Such equipment is more expensive but is not implicit in technique. Still costing less compared to X-ray equipment or detectors (eg. scintillators or GM tubes) designed to pick up the movements of specially prepared radioactive 'particles'.

The previous points might be regarded as desirable aspects of a technique for investigating fluidisation. Tables 4.1 and 4.2 have been drawn up by TA'EED et al. (1984) with respect to the these mentioned criteria, for 14 research beds including the research bed for this present study.

It is useful to note the following :

For techniques involving radioactivity. Monitoring of bubbles is not possible. Only a single radioactive particle is used for tracking. The resolution of detector arrays is insufficient for simultaneous tracking of particles. Particle size can be large, also artificial and therefore spherical in shape. This is restricting for realistic applications. Since attempts are made to determine particle movement throughout the bed the assimilation of sufficient data is lengthy, being in the order of hours.

X-ray techniques tend to be more flexible. For particle tracking it is important to use bed material of differing X-ray opacity, leading to thoroughly untypical bed (since the path length of X-ray to the tracked particle diameter is important). This is illustrated in the tables. Individual particles cannot be followed because it is impracticable. Monitoring bubbles is relatively straight forward permitting any bed shape. However overlapping

TABLE 4·1

Research	METHOD OF TRACKING	REALISTIC SIMULATION OF OBJECTIVE	SPECIFIC ADVANTAGES	LIMITATIONS
A Masson, H. et al.	Tracks isolated spheres (5-20mm ∅)(contain radioactive particles) by Electronic Transition Identificator. Radioactive.	Needs further modelling	Non-invasive	Radioactive hazard. Material unrealistic. Spheres too large. Tracks one sphere only.
B Fitzgerald, T. et al.	Ferrite Tracer introduced and monitored by arrays of induction probes. Magnetic field.	Aims to look at solid speeds for large bed simulation, therefore bed realistic	Large bed (2000 pounds) Non-invasive	Drop in tracer which can form clumps. Drop cuts in probe signal. Tubes must be used.
C Loew, O. et al.	Stereo-Photogra-Metric method. Measures displacement fields + cine filming use dyed particles. (6% of bed) Stereo Photogrametric		Produces optical 3-D image. Detailed qualitative information. Non-invasive	Powerful computer 1 particle tracked. Radioactive hazard.
D Chen, M.M. et al.	Radioactive tracer technique with 1 particle for 5 hrs. using 16 scintillators. Radioactivity.	Comparing particle circulation with and without internals.	Virtually automated from 'setting up' to results. Non-invasive	Powerful computer 1 particle tracked. Radioactive hazard.
E Merry, J.M. et al.	Outer circulation photographed, (+ dye); Inner movement tracked with radio pill, aerial in bed. Photography Radio Signal. Two beds, gas & liquid fluidised	Not aiming to produce even fluidisation, since simulating 'Gulf stream'.		Basically 2-D in both cases. Radio pills motion not very representative of bed material.
F Hager, W.R. et al.	Tinted beads arranged above a tube surrounded by white beads, using cine & photography upon fluidising. Photography.	Looking at flow regimes around cylindrical objects especially defluidisation around cap region. Bed material not realistic.	Simple	2-D specification for 'cap' observation with tubes. Initial flow only.
G Peeler, J.K. et al.	Certain special tubes have glass windows 'fitted flush inside. Light via fibre optics & mirrors allows inside bed viewing between tubes. Photography	Looking at movement of dense phase around tubes to help determine heat trans. Bed type well considered. Good simulation.	Large bed Gives view inside bed for region around tubes	Must use tubes, unclear how dense phase is measured. Cannot track particles.
H Cranfield, R.R.	Impregnated particles provide magnetic field response. Bubbling measured inductance probe + phtotography. Magnetic	Induced bubbling to determine flow and size, or free bubbling not traced. Not v. good for indicating movement.	Non-invasive	Tracer movement is crude Does not correlate 2-D with 3-D bed.
I Bernard, J.M et al.	Uses(L)fluidised system, with matched Refractive index. Signal processing & data reduction. Laser interference + video for cf. L.D.V.	Modelling for turbullent reacting flow. Very unrealistic material; size & type.	Measures particles in 3-D. Very specific local values. Non-invasive	Measuring section 4x10cm∅ only (of bed). Accuracy by complex stats. correlation Lq needs same n. Lq a fire hazard & powerful solvent.
J Nienow, A.W. et al.	Aluminium particles & sheets used to produce tracing material for X-ray cine. X-Rays	Attempting to verify mixing & segregation mechanism for larger varying shaped particles; Flotsam & Jetsam. Fairly realistic.	Bed sectioning not required. Can use small numbers of large particles. Non-invasive	Particles of interest much larger (<1.5cm ∅ & 2.4x2.4x0.08 card). Time averaged. X-rays need screening.
K Kondulov, N.B. et al	Radioactive isotope for tagging. Tracking with scintillation counters. Detectors cover 3-D axis + filming. Radioactive	Monodispersed particle bed. 1 tracer particle is realistic CF to rest of bed.	Non-invasive Fairly close tracking of sphere.	Only tracks one sphere Interpretation by oscillogram comparison tricky.
L Canada, G.S et al.	Flooding the two beds for filming and observation. Record ΔP, and void fraction with probes simultaneously. Cine	Trying to isolate effect of particle size & flow behaviour. Uses probes and visual observations. The void age is then related to regimes. Not looking closely for bubbles.	Large flow regimes covered.	Filming, thus 2-D Cannot detect particles
M Yung, J. et al.	Fluorescent particles when bombarded with (N_2) laser pulses. Fluorescence is recorded from observation. Laser & Fluorescence	Showed some mathematical models to accommodate 2-D effects and find functional relationship P_w (wake fraction). Realistic in this.	Found particle tracks from probes bombarded. Fairly distinct.	2-D. Does not actually track bubble. Purely on chance if one appears when laser pulses. Tracking particles do not last. Small area of bed examined.
N Ta'eed, O. et al.	Dyed particles fluorescent under U.V. light are tracked with and without internals. A rotating chopper allows a number of exposures on one still. Ultra-violet Photography.	Provides realistic conditions for fluidised bed analysis while using sufficiently discrete sand size.	Speed of Data Simplicity; can use wide range of bed material. Flexible (bubble and particle Vs)	2-D

TABLE 4·2

bed mat.	ø/mm	ρb g/cm	Umf cms	ū sec/cm²	H/cm	Dist	Tube (s)	Detection Particles	Bubbles	ACCURACY	COST	OPERATOR SKILLS. ESTABLISHED OR NOT	INTERPRETATION OF DATA	SPEEDS OF RESULTS.
A	glass & Tr.	0.3 uniform	1.09-1.29	with air	18 x 13 medium	1.9 : low	0.04; fine	No	One sphere continuously	X	Response time 0.1 ms	Computer	Yes	Complex	Slow
B	course sand & Tr.	1.04		59 (air)	100 x 100	76.2	5.08 cm (x 64 tubes)		Detects ferrite material within 15.24cm of probe	X	Response time <10ms Results not v. reproducible	Video & data acquisition equip. z80 micro.	No	V. Tortuous	Rather slow. (loading bed etc.)
C		2 - 2.78; large	0.85	79.4(air)	29.2x2.7	57	0.165mm	cylinder, sq. ovals	Particle simultaneously	Bubble shape	Maximum Error, Vp = ±2.4 mm/s	Computer £40,000+	9-12 mons training / First time used.	Results easy	Long time to process results.
D	glass beads & Tr.	0.42-0.6		38.8 & 56.9 air	19.05 ø	19.05	Fine grain	Various rod bundle sizes		X	Vp 3% error Spatial resolution 5mm	Computer tracking facility. Tr-metallic scandium resin coated	Yes	Graphic patterns readily interpreted	± 5hrs + computing time
E	Gf sand,coal ash,Sialm W; Pb shot	0.13;0.3; 0.05 / 0.06;0.2	2.5; 5; 0.4 / 2.6; 9.3		15xwidth / 0.9xwid.	25 / 12.5	0.012mm brass prs / Not even		Front & central Vp	At front of bed			Yes		
F	glass beads	0.56			30.5 x 1.9	61	Porous Al. plate	1 of 2 cylinders abv.	Track particle circulation above 'cap' area.	X	Fairly rough (turnover time)	Low	Not very difficult. / As colour tracer;yes	Straight forward	Fast
G	silica sand	0.297-1.19	1.61	30 (air)	122 x 122	122	3.8cmø (x 92)		Dense phase	Observes movement	Not very accurate. 3-D difficult to calibrate.		Average / Not established		
H	alkalised alumina	1.52-1.76		47 (air)	two beds 610/610x20	15 (at Umf)	Not even	No	X	Velocity	Fair		Not much. / Fairly well established	Vague	Fair amount of time
I	soft glass (some dyed)	0.6	2.7	? (benzyl & ethyl alc)	7.6 ø	5.7	Al. 'honey-combs'	No	At one particular position in bed measures	X	Special resolution 0.5mm Error V=±3%	Very Expensive	First time used.	Data needs computer processing	Fair amount of time
J	Alumina	0.265	1.45	2.73(air)	0.14 x depth	10	Plastic porous	No	Large particle	X			Not much	Observational Straight forward.	
K	alumino-silicate	0.265-0.29		84 [air]	17.2	50		No	Single particle movement over 10-20mins	X	Enlarge oscillograms for acc. Perturbation of 3-2mm error.		Fair amount to decifer oscillograms. / 1st time		Very long time consuming (without computer) (10-20/run mm)
L	soda lime glass, Ba, Si	0.0595-.07 / 0.233-.233			i) 30.5 sq. ii) 61.0 sq.	i) 25 ii) 30	Bubble cap type	No	X	Bubble regimes	Fairly		Fair amt. since gt. deal of obv V.well est.	Straight forward since qualitative	Very long time consuming
M	40% ZnS, 60% R...	0.174	0.540		40 x 1.5	224	Not even distribution	No	Seen while it fluoresces	Observe			Reasonably skilled to observe. / 1st time.	None given	Time consuming (observing)
N	sand	0.1 - 1.3	2.0	10-170	13 x 30	5 - 8	high density polythene	Yes	Velocity vectors for both		Good	Cheap (under £1000)	Very Little / Yes	Computer processing gives v. clear results	Quick - 6 hrs turnover

bubbles cannot be distinguished and the bed thickness is limited (to less than 40cm).

Magnetically sensitive probes are an indirect technique for monitoring bubble formation in conjunction with bed internals, such as tubes, etc. The data produced is a fairly crude guide to the fluctuating quantity of gas around the bed internals, no information can be obtained on bubble size. Particle mixing rates can be found if particles of different magentic permeability are used (ie. metal to provide sufficient sensitivity).

Photographic techniques are generally used in conjunction with 2-D beds, since light cannot penetrate an opaque bed. This is the most established technique and has largest number of variations. Both particles and bubble tracking is possible, but as yet not simultaneously.

Hopefully the tables should provide a clear picture of the potential for the technique used in this study, albeit with the limitations of any 2-D bed. This is not severely restricting as can be testified by comparison with limitations of other techniques. Many research beds are designed to provide relative parameters, that is, velocities are relative with respect to all parts of the bed. Arguably the most important aspect of the technique is the implicit simplicity of a 2-D bed. This provides the the technique with the versatility and scope for studying many different parameters associated with fluidisation.

The results obtained by the technique are complete in detail both quantitatively and qualitatively for particles and bubbles, rapidly as a bonus (especially if adapted for use with video equipment). The capital and operational costs are also significantly lower than other methods.

4.4 EXPERIMENTAL TECHNIQUE

The experimental method is uniquely simple to set up, operate and to evaluate results. A 2-D glass bed of size 0.5m x 1.0m x 0.008m (6mm thickness glass) is set onto a perspex plenum chamber and sealed with common domestic silicon sealant. A fan provides fluidising air (up to 3m/s) through both sides of the plenum chamber and through a high density polythene distributor plate. There is, in the case of draft tubes, a separate fan which supplies air to the centralised jet, 0.11m above the distributor plate. The draft tubes are placed on either side of the jet at equal distances of 0.04m, as shown in Figure 4.2. The draft tubes were attached to the bed with double sided sticky tape. In the case of the jets, since the bed anatomy is axially symmetrical, only two jets were employed to simulate a novel recirculation technique. The jet

FIGURE 4·2

Draft Tube System Dimensions
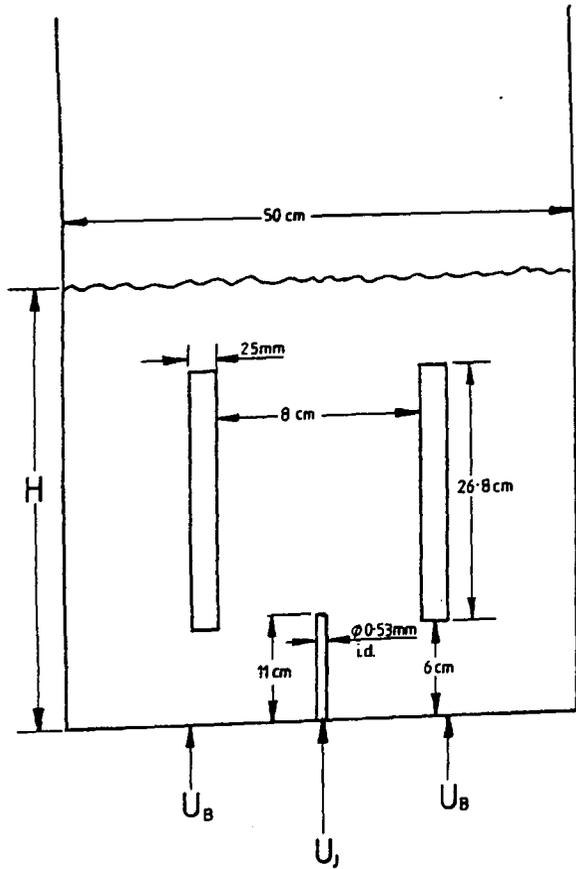
FIGURE 4·3

Jet system dimensions
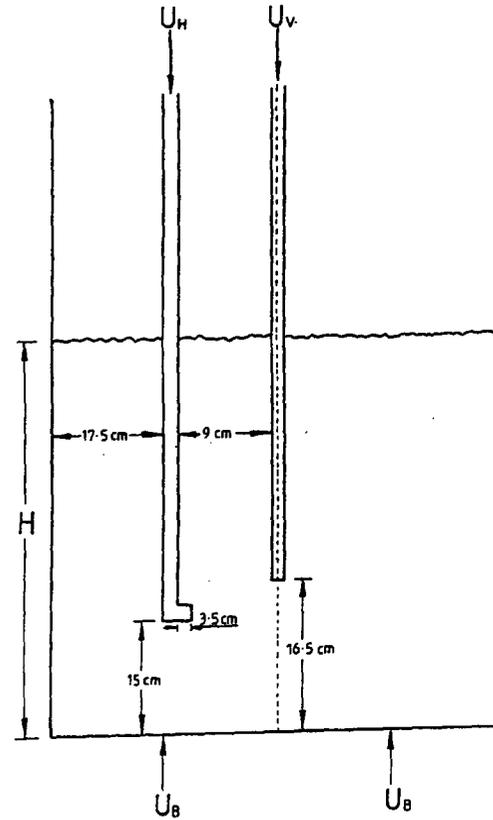
dimensions are given in Figure 4.3. The dimensions of the draft tubes and jets were designed to allow for circulation across the tubes where the particle motion was induced whilst supporting the stationery parts of the bed. This enabled the optimum position and size of the tube to be evaluated with respect to bed velocity. Bed material, may be of any type or size, in this case sand (average diameter 0.414mm, sphericity 0.68, density 2.452 Kg/m^3, minimum fluidising velocity 25cm/s), is poured into the bed to the required bed height. The bed is surrounded by four 60cm fluorescent strip lighting fittings with reflectors, using 20W 365.5nm blacklight blue tubes, as shown in Figures 4.4 (a) and (b). A small proportion of the sand is dyed with a fluorescent sulphur based dye, which adheres to the sand well without altering the sand properties significantly. Under the UV light, the particles provide a sharp contrast between the 'black' sand and the fluorescing particles. A camera with a 1sec. f5.6 shutter and UV filter was placed behind a rotating 20cm 16slit chopper, with speed of 15 slits/second for sand at bed velocity of 1m/s. the chopper speed and camera shutter speed are varied for different conditions. The chopper speed was measured accurately with a photo-electric diode connected to a timer, this enabled the velocity of the particles or bubbles to be evaluated.

FIGURE 4·4

(a) Schematic diagram – Jets
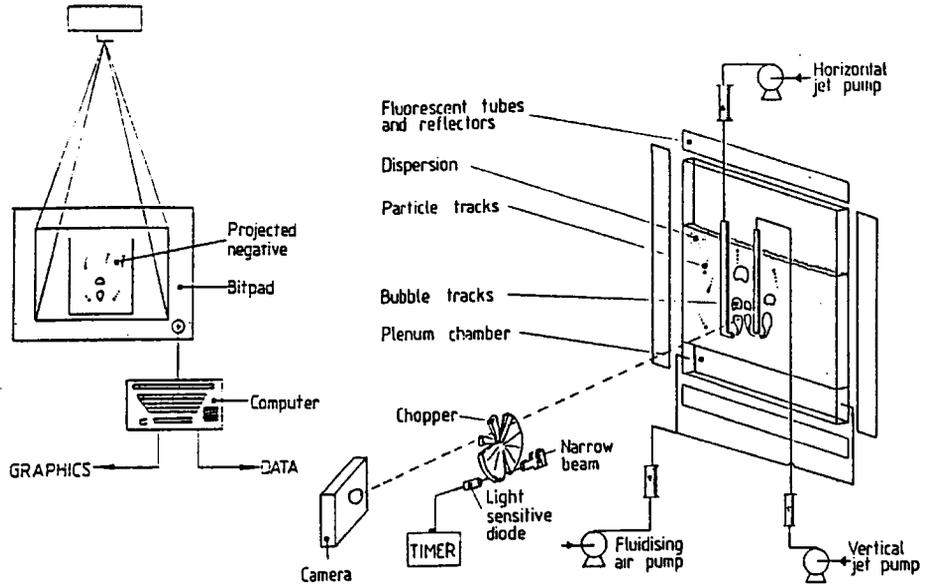
(b) Schematic diagram – Draft tube
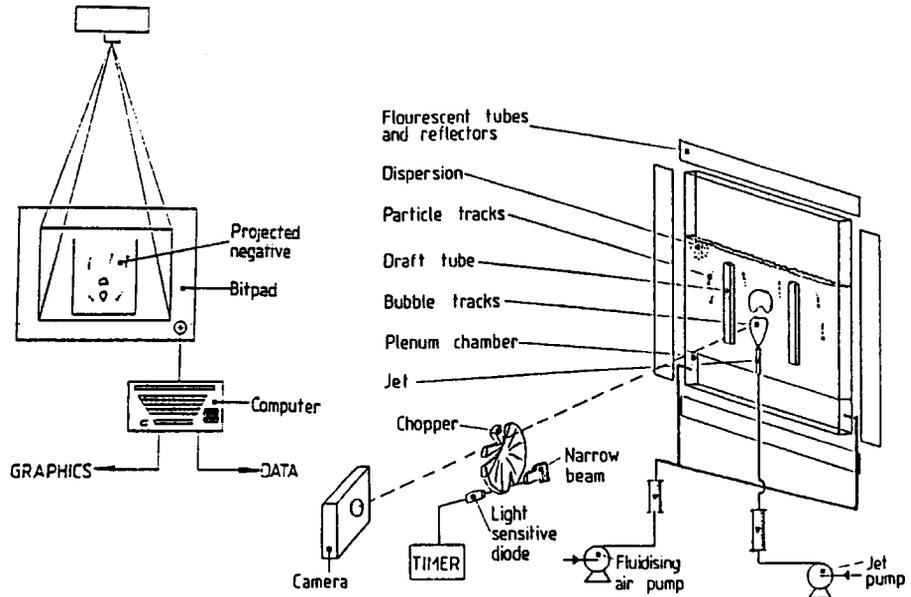
When a picture was taken the dyed particles left
an image of tracks on the B/W 400 ASA negative which was
then developed to high contrast with microphen developer.
The negatives were then projected down from an enlarger
onto a 50cm bit pad which was attached to a microcomputor.
By digitising the control points, the tracks were recorded
and processed both numerically and in graphical 'picture'
format. The bubbles were similarly photographed and
tracked by placing a white fluorescent card behind the bed,
covering 15 of the slits on the chopper, hence having only
one open slit, and running at typical chooper speeds of 2.7
slits/second. This enabled the bubbles to be 'frozen' as
well as providing adequate bubble displacement. Exact
bubble shapes were digitised by removing the chopper and
taking fast shutter speed (1/500 second) shots under UV
light or white light if required.

This technique eliminates the errors encountered
using large tracer particles, as in the case of the 'radio
pills'. It also elinimates the errors arising from any
intrusive obstacles which might impede the flow, such as
aerials which would need to be inserted in the bed if the
radio pill technique was adopted.

In the present study three sets of data have been
collected regarding the tracking of particles (and bubbles)
to determine circulation rates within the fluidised bed.
The first set of data concerns the determination of the

circulation rates and flow patterns in an empty bed (ie. no obstacles). The second set concerns the circulation rates and flow patterns using draft tubes and a centralised jet, as shown in Figure 4.5. The final set is again concerned with the determination of circulation rates and flow patterns but in this case employing a novel jet circulation system, using horizontal and vertical jets, as shown in Figure 4.6. Measurements were taken to explain the effect of the following parameters on the hydrodynamics of the recirculating fluidised bed :

(A) Empty Bed

    (i) Bed velocity (50, 75, 100 and 125 cm/s).

    (ii) Effect of tracer size (0.414 and 1.008 mm).

    (iii)Bed height constant at 30cm.

(B) Draft Tube System

    (i) Bed height (30, 35 and 40 cm).

    (ii) Fluidising bed velocity (16.70, 20.83 and 27.5 cm/s).

    (iii)Centralised jet volumetric flowrate, was constant at 80lit/min (8.33×10^{-4} m^3/s).

(C) Jet System

    (i) Twin jet volumetric flowrate of 66, 80 and 97 lit/min (11, 13.33 and 16.17×10^{-4} m^3/s)

    (ii) Fluidising bed velocity (16.70, 20.83 and 27.50 cm/s).

(iii) Bed height was constant at 30cm.

In all the tests the following measurements were taken :

(1)   The bubble stills enabling the size and shape of the bubble to be evaluated,

(2)   the bubble tracks enabling the determination of the bubble rise velocities,

(3)   the particle tracks enabling the flow patterns and the individual particle velocities (hence circulation rates) to be evaluated, and

(4)   dispersion measurements were also taken, by dropping particles on top of the bed, enabling dispersion rates to be evaluated.

Since the 2-D bed determines relative velocities (not absolute) in the bed, the results are comparative.

FIGURE 4·5

Recirculation patterns with draft tube system
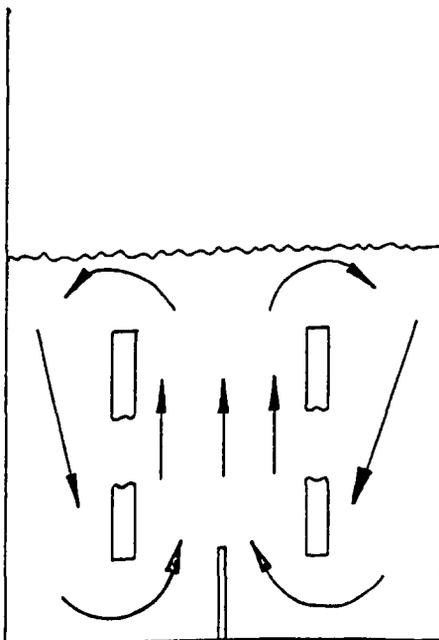
FIGURE 4·6

Recirculation patterns with jet system
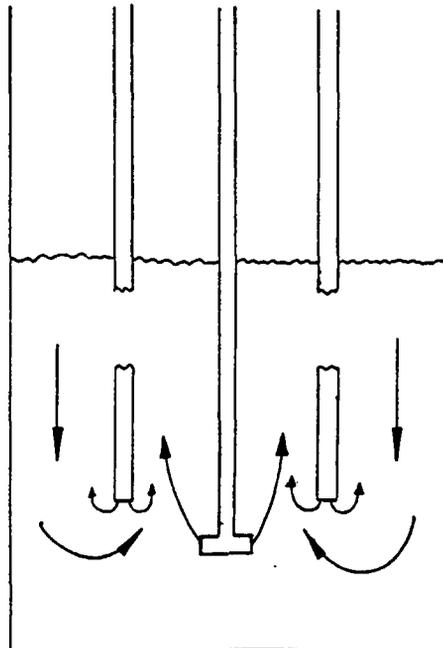
4.5 RESULTS

Initially the justification of using tracer particles the same size as the particles under investigation will be discussed using the dispersion data obtained for a free-bed (ie. no obstacles), employing two different sized tracer particles. The results obtained for all the three sets of data will then be discussed individually. Later each techniques' merits will be compared, with predictions and possible utilisation of their respective systems.

4.5.1 Effect Of Tracer Size

Tracking of particles is a very complex technique and usually relies heavily on the tracers' ability to follow the particles under investigation as realistically as possible. This has led to various different and elaborate techniques employing different types of tracers, as discussed previuosly in this chapter. The most common assumption made fron previous studies (in particular the radio pill techniques) has been that the tracers would follow the particles under investigation as long as the density of the tracers approached that of the particles being tracked. The influence of the size of the tracer on

its tracking ability has not been fully realised. Infact
some investigators in this field have assumed that the size
has no great effect as long as the densities are similar.
This present set of results considers the effect of two
different sized tracer particles on the dispersion rates in
a cold 2-D free bed. Static bed height of 30cm, fluidising
bed velocity of 50cm/s and bed particle (sand) size of
0.414mm is used in both cases, with the tracer particles of
the following specifications :

      (a)   Same sized tracers

            (average diameter = 0.414 mm)

      (b)   Larger sized tracers

            (average diameter = 1.008 mm)

     The numerical data obtained from the experiments
are given in Tables 4.3 and 4.4. The tables show the
displacement rates and the bandwidth in the X and Y
directions with respect to time. Displacement rates are
values obtained taking the mean modulus of the movement of
the tracer particles in the specified directions.
Bandwidth is obtained by taking the mean modulus values of
the standard deviation in the specified directions and
gives us the width of the displacement rates.

     Figure 7 (a) and (b) shows the graphical output of
the dispersion patterns obtained from the processing of the
particle tracks from the negatives using two different
sized tracer particles.

TABLE 4·3

<u>DISPERSION DATA</u>– FREE BED (H = 30 cm , U_B = 50 cm/s)

Mean modulus values	Time/seconds									
	0·000	1·014	2·028	3·042	4·056	5·070	6·084	7·098	8·112	9·126
Displacement in X-direction, cm	0·00	0·32	1·56	1·37	1·48	0·94	1·43	0·29	1·34	0·61
Displacement in Y-direction, cm	0·00	2·93	3·21	2·95	5·06	5·83	8·64	11·57	12·85	14·90
Bandwidth in X-direction, cm	0·00	0·12	1·56	0·97	1·28	1·08	1·81	0·47	3·11	0·87
Bandwidth in Y-direction, cm	0·00	0·06	0·43	1·32	0·54	2·03	3·38	2·93	1·71	5·18

TABLE 4·4

<u>DISPERSION DATA</u>– FREE BED USING LARGER TRACERS
(H = 30 cm , U_B = 50 cm/s)

Mean modulus values	Time/seconds						
	0·000	1·544	3·088	4·632	6·176	7·720	9·264
Displacement in X-direction, cm	0·00	2·26	2·39	3·31	2·19	0·39	0·79
Displacement in Y-direction, cm	0·00	16·38	20·23	23·82	25·30	26·03	28·90
Bandwidth in X-direction, cm	0·00	0·60	0·18	0·71	0·22	0·44	0·32
Bandwidth in Y-direction, cm	0·00	0·61	0·51	1·28	1·24	2·78	1·40

FIGURE 4·7 (a)

Graphical output from negatives of free bed, same sized tracers

VERTICAL HEIGHT / cms ⟶

PLENUM CHAMBER ⟶

FIGURE 4·7 (a) cont'd

VERTICAL HEIGHT / cms

PLENUM CHAMBER ⟶
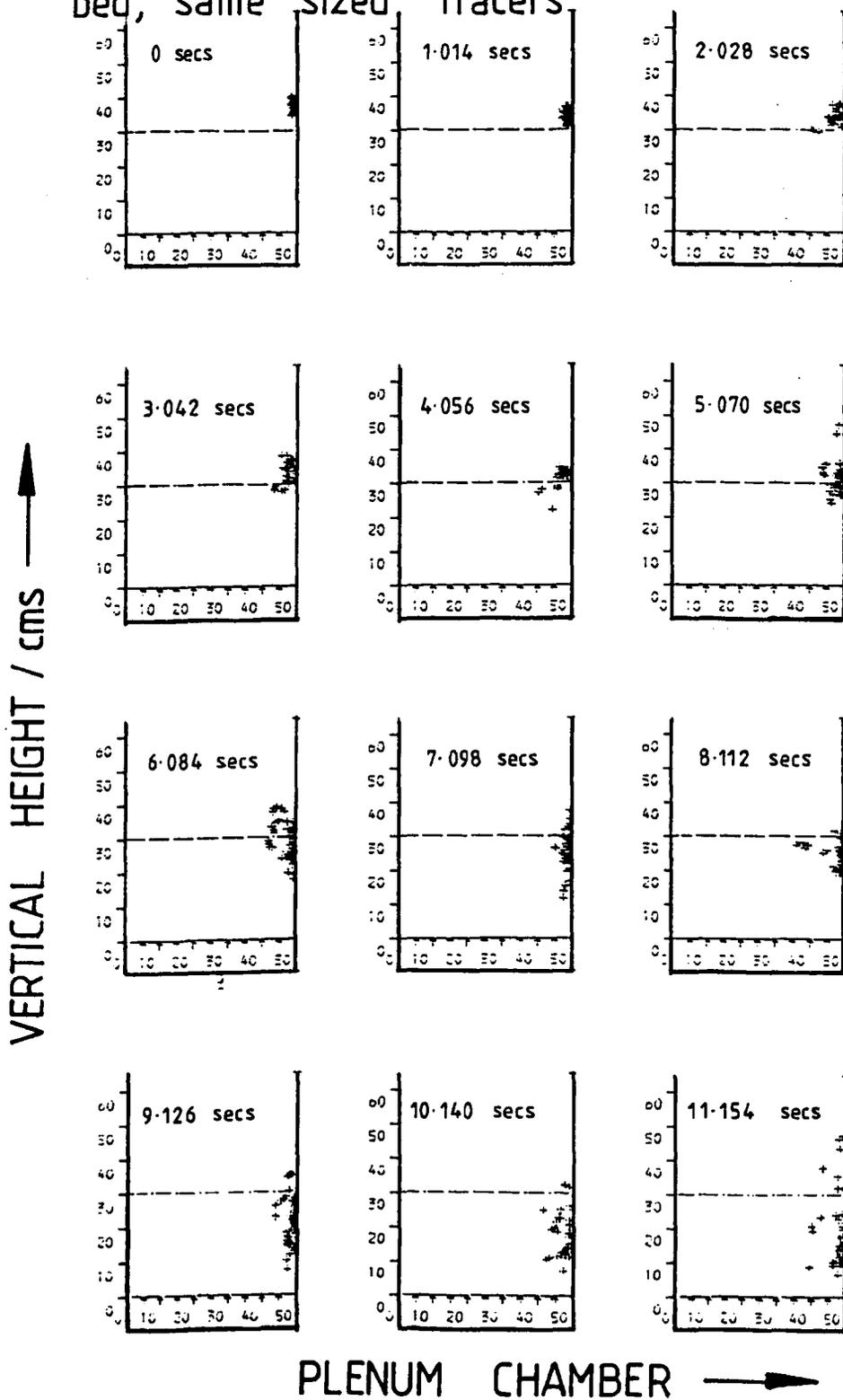
FIGURE 4·7 (b)

Graphical output from negatives of free bed, large sized tracers

VERTICAL HEIGHT / cm

PLENUM CHAMBER ⟶
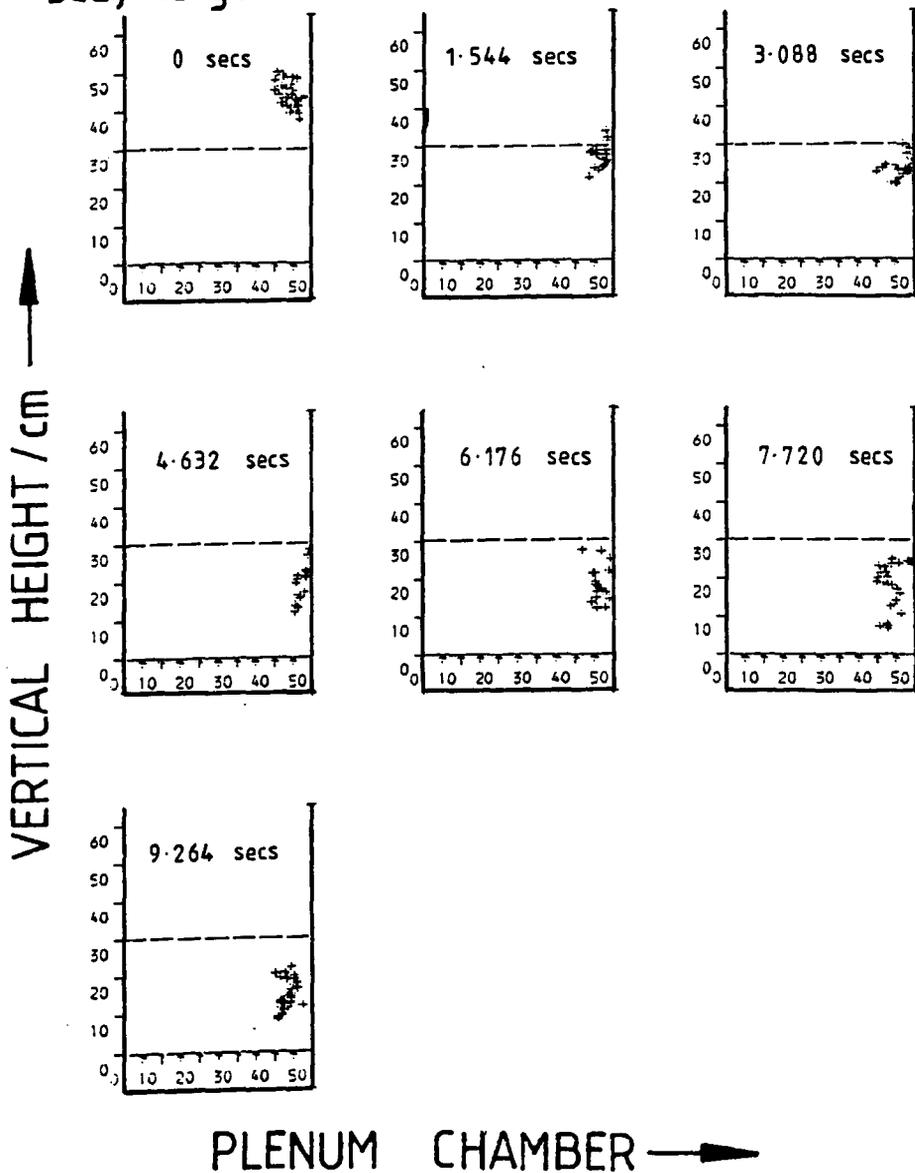
Figures 4.8, 4.9 and 4.10, (a) and (b), show the graphical output based on the numerical data from the tables. Even at a glance at Figures 4.8 (a) and (b), dispersion rates for different sized tracers, it is obvious that the large sized tracer does not follow in the same directional displacement nor magnitude as the same sized tracer, even though they might have similar trends in certain directions. Discussions on the indivdual displacements and bandwidths for the X and Y directions which follow will make it quite clear that the size of the tracer particles is an important parameter regarding the accuracy of the tracking technique.

Figure 4.9(a) shows the mean modulus displacement in the X-direction. From the graph it can be seen that horizontal movement fluctuates with time for both the tracers, with a larger displacement being obtained fro the larger sized tracer.

Figure 9(b) shows the mean modulus bandwidth in the X-direction. Bandwidth is basically the horizontal width of the particle movement. This has a similar trend as the displacement curves, but a larger bandwidth being obtained in the case of the same sized tracers.

Figure 10(a) shows the mean modulus displacement in the Y-direction (downwards). This shows significant difference between the two tracers, large tracer falling at a markedly faster rate.

FIGURE 4·8   DISPERSION DATA

(a) Free-bed , H = 30cm, U_B = 50 cm/s
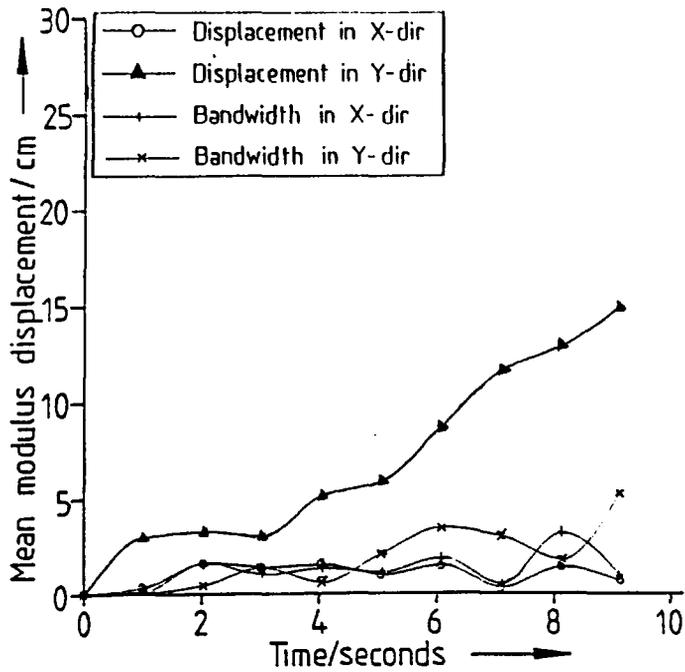
(b) Free-bed / large tracers
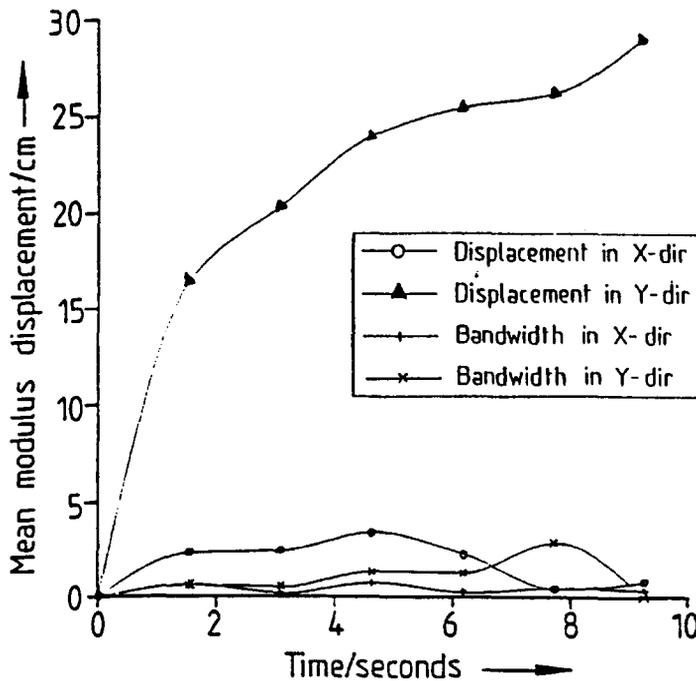
(a) Dispersion Data - Free Bed (H = 30cm, U_B = 50cm/s)
    Effect of tracer size on X-dir. displacement.
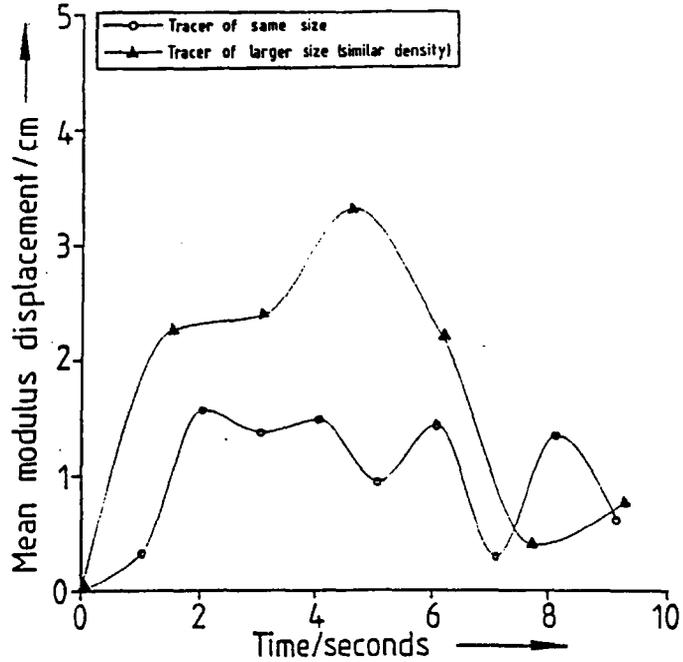
(b) Dispersion Data - Free Bed (H = 30cm, U_B = 50cm/s)
    Effect of tracer size on X-dir. bandwidth.
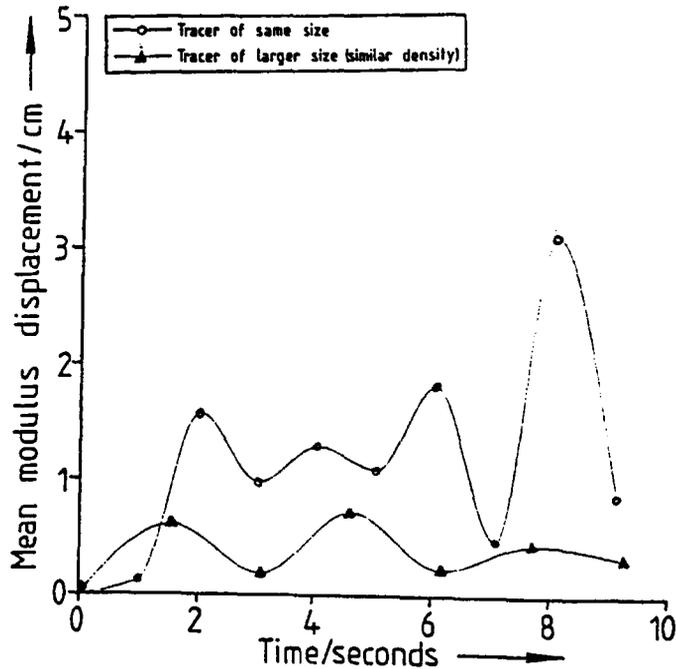
FIGURE 4·9

Effect of tracer in X-direction

(a) Dispersion Data – Free Bed (H = 30cm, U$_B$ = 50cm/s)
Effect of Tracer size on Y-dir. displacement.
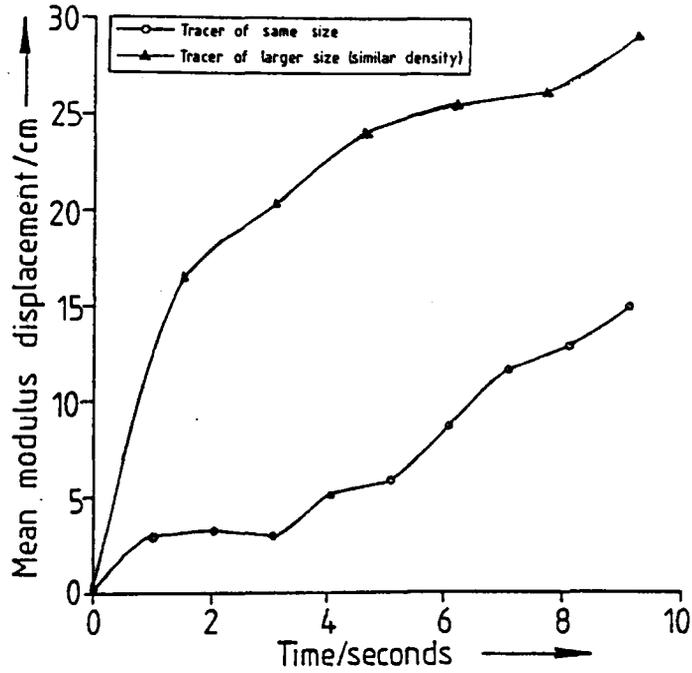
(b) Dispersion Data – Free Bed (H = 30cm, U$_B$ = 50cm/s)
Effect of tracer size on Y-dir. bandwidth.
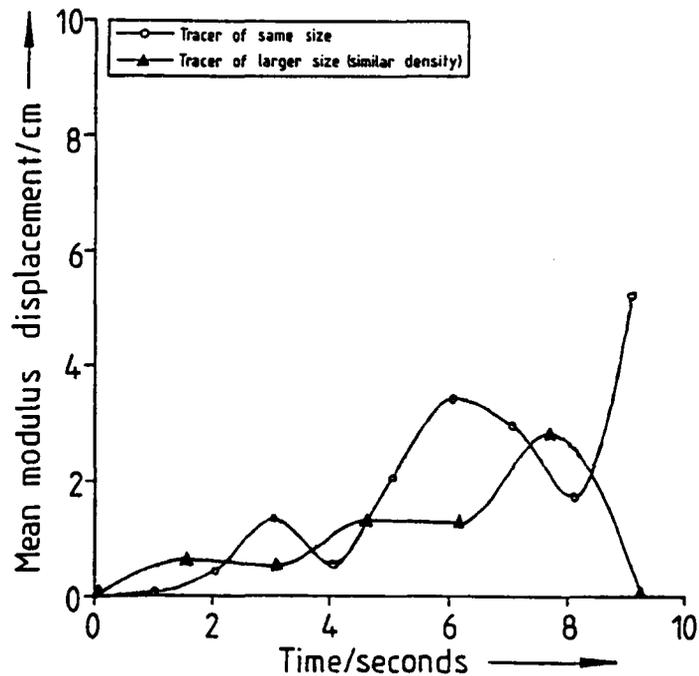
FIGURE 4·10

Effect of tracer size in Y-direction

Figure 10(b) shows the mean modulus bandwidth in the Y-direction. These curves show a similar trend and magnitude of fluctuations for both the tracers.

As a conclusion from the figures it can be said that the large sized tracers are a poor indication in the tracking of smaller particles, in particular in the Y-directional displacement, even though their densities might be similar. It would therefore be inaccurate to establish any hydrodynamic behaviour regarding bed particle movement using larger tracers. In this study, same sized tracers as the particles under investigation have been employed to obtain a more accurate answer to the hydrodynamic problems encountered.

4 .5.2 Dispersion Data

Dispersion rates have rarely been investigated in the past. It is important to be able to predict the dispersion rates since this would give an indication of the rate at which any material, fed from the top, would disperse within the bed. Previous researchers have neglected the possibility of feeding material on the top of the bed and have thus ignored the dispersion rates. In this study dispersion rates have been investigated, for both the draft tube system and the jet system, with regards

to the displacement and bandwidth in the vertical (X) and horizontal (Y) directions with respect to time. The effects of bed depth, bed velocity and jet velocity have been investigated.

4 .5.2.1 Draft tube system

Figures 4.11, 4.12, 4.13, 4.14 and 4.15 show the representation of the digitised negatives in graphical form. Tables 4.5, 4.6, 4.7, 4.8 and 4.9 are corresponding numerical data obtained from the digitised negatives. Figures 4.11,4.12 and 4.13 show the increase in fluidising bed velocity (16.70, 20.83 and 27.50 cm/s respectively) while the bed depth remains constant at 30cm. It can be seen that the displacement of the dispersing particles in the Y-direction (downwards) increases with increasing bed velocity. The trends in the bandwidth of dispersion in the X and Y-directions and the displacement in the X-direction, are not so clear. Comparing Figures 4.12, 4.14 and 4.15, which represent the increase of bed height (30, 35 and 40 cm respectively) for constant fluidising bed velocity of 20.83 cm/s. It is not very obvious what the displacement abd bandwidth trends are, or if there are any trends. The only clear observation which can be made from the figures is that at a bed height of 40cm the dispersion rates greatly reduced. For a clearer picture graphical output were processed to yield numerical values, Tables 5 to 9.
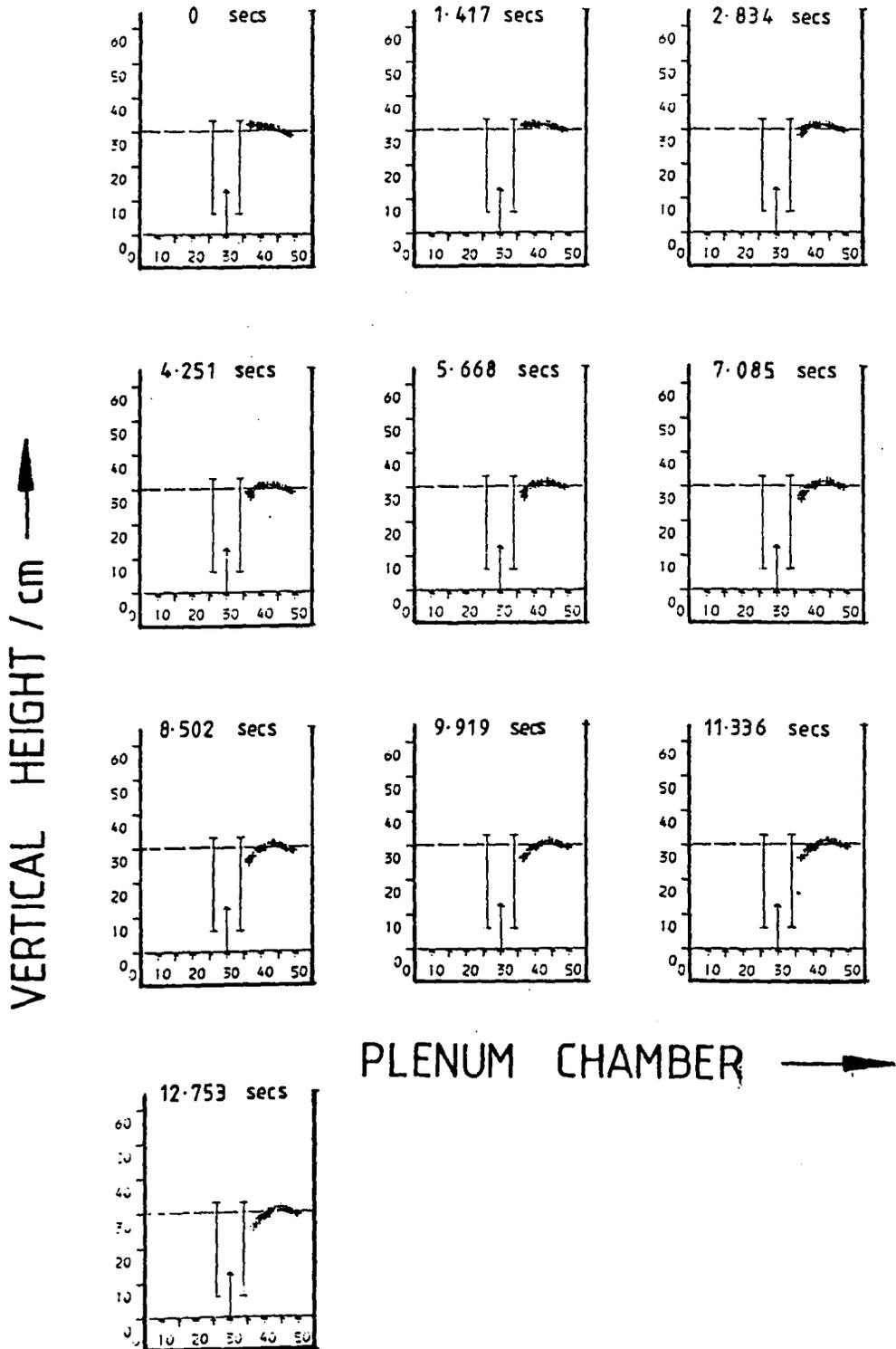
FIGURE 4·11

Draft tubes (dispersion)

$U_B = 16·70$ cm/s, $H = 30$ cm

PLENUM CHAMBER ⟶
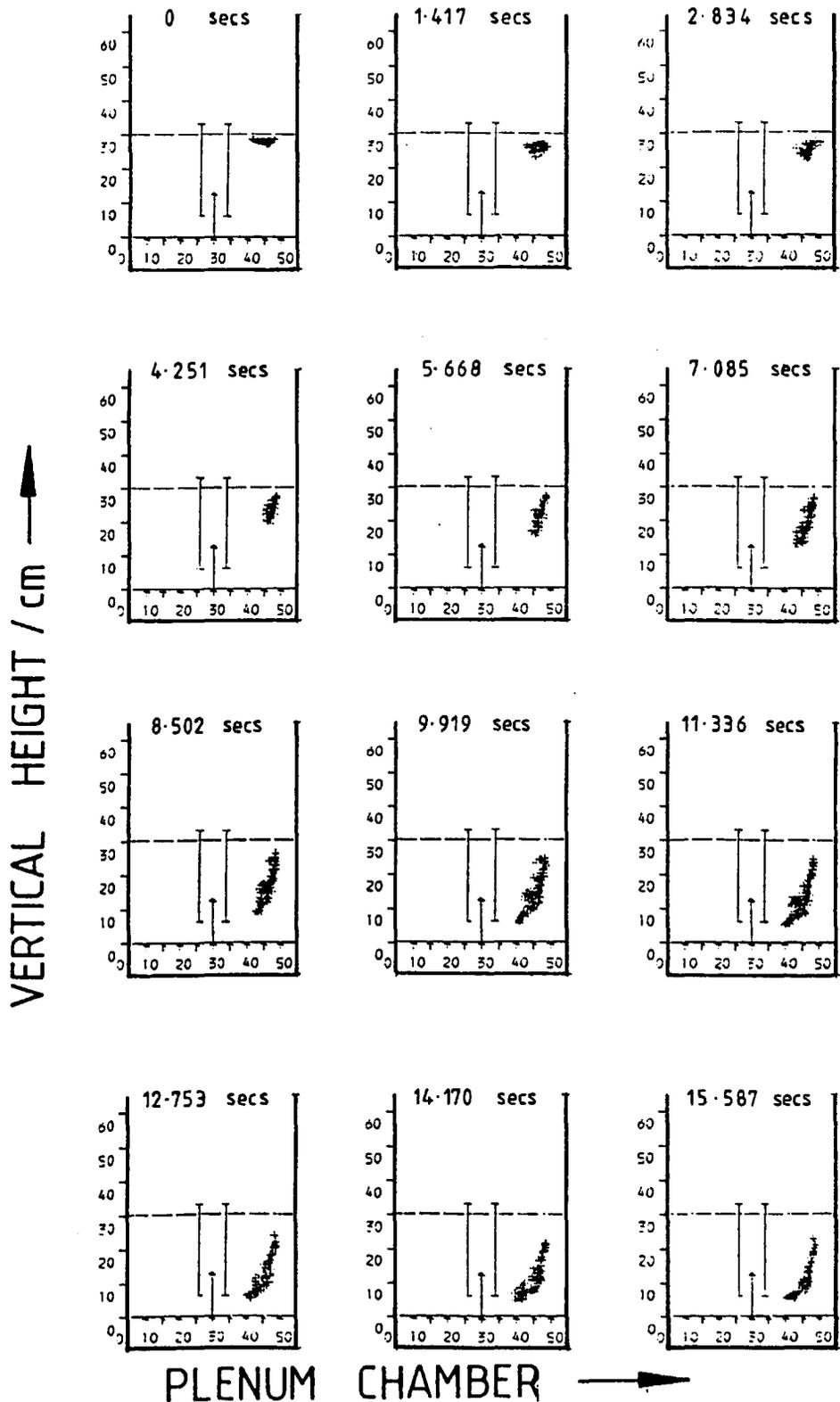
FIGURE 4·12

Draft tubes (dispersion)
$U_B = 20·83$ cm/s, $H = 30$ cm
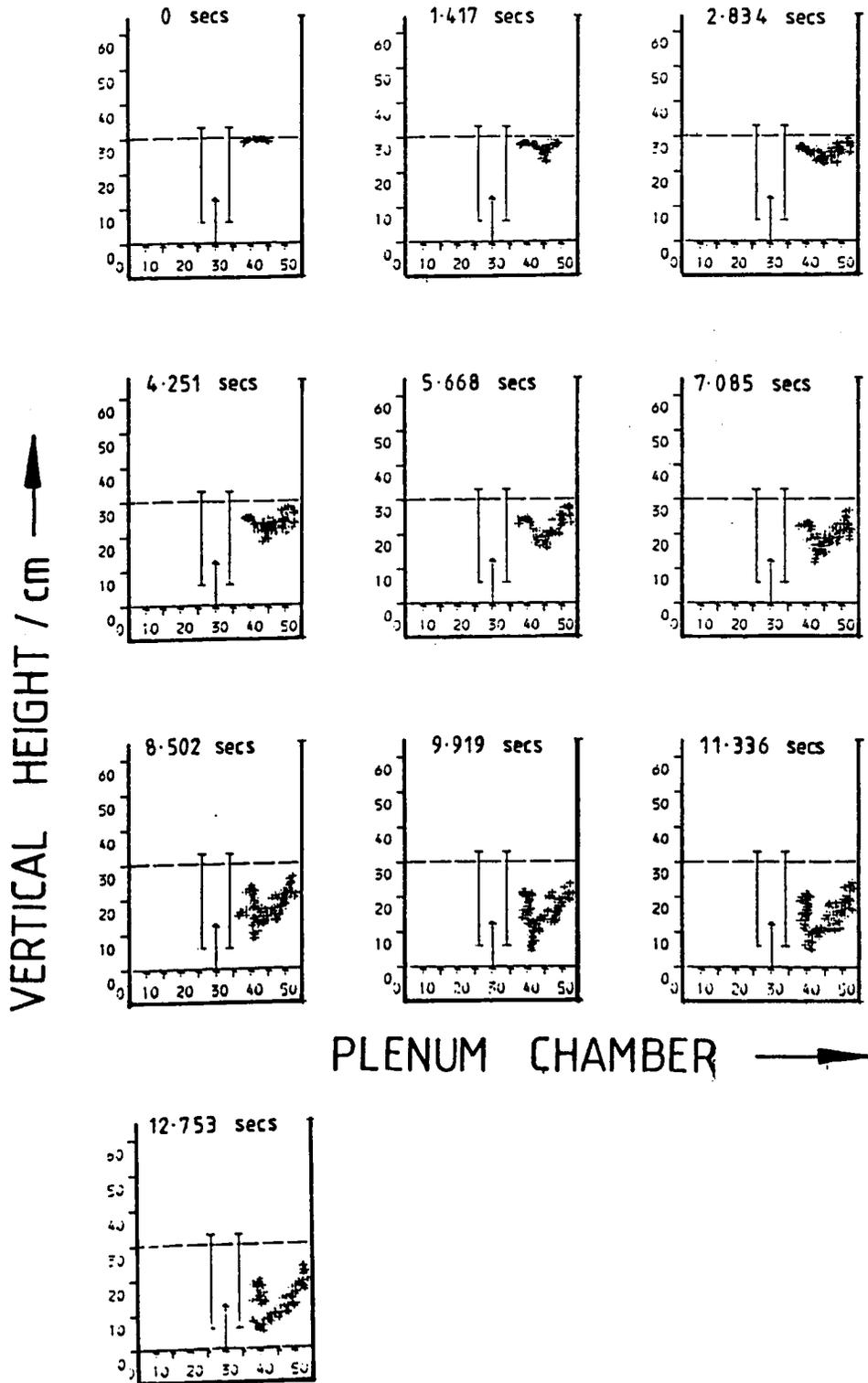
FIGURE 4·13

Draft tubes (dispersion)
$U_B = 27·50$ cm/s, $H = 30$ cm
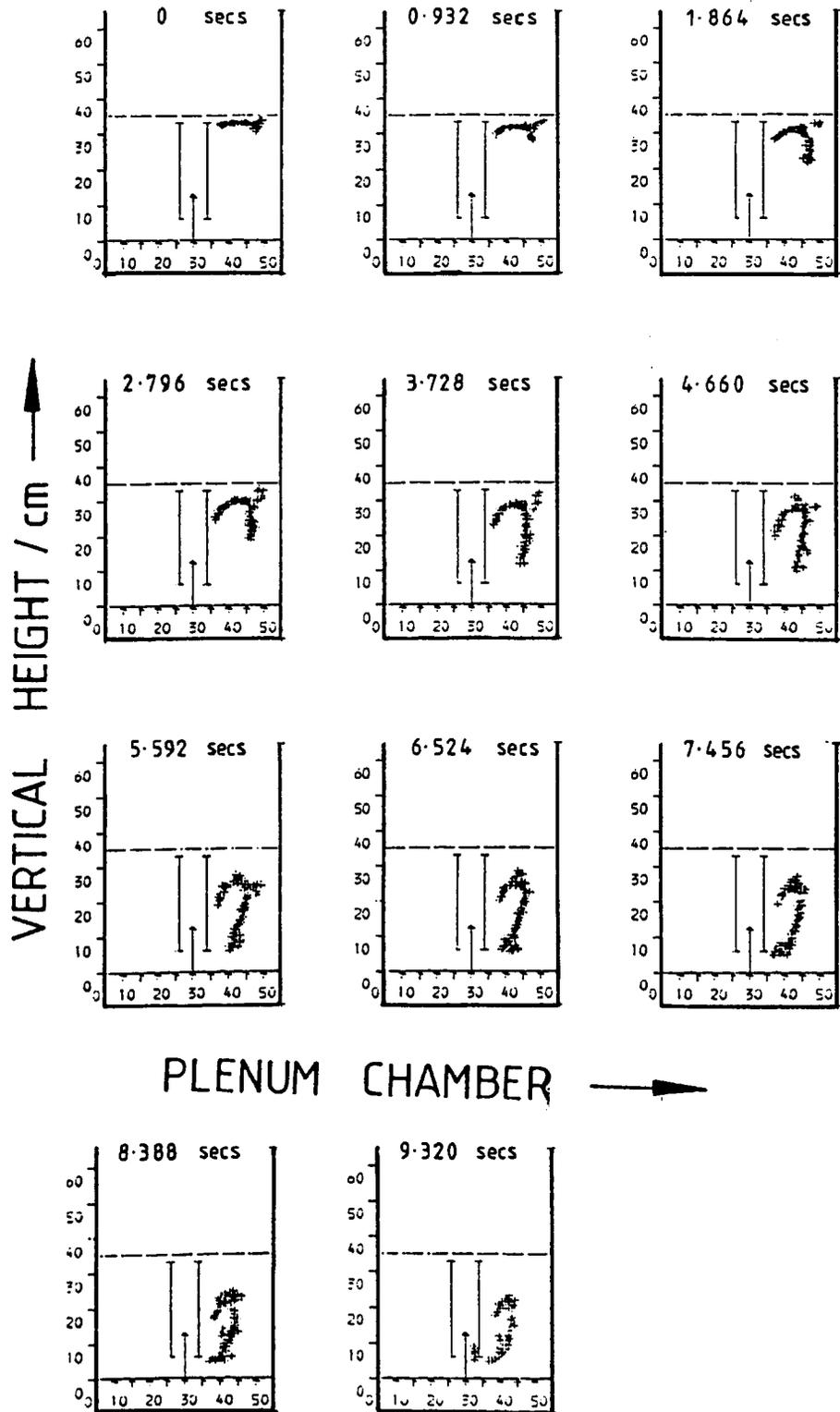
FIGURE 4·14

Draft tubes (dispersion)
$U_B = 20·83$ cm/s, $H = 35$ cm
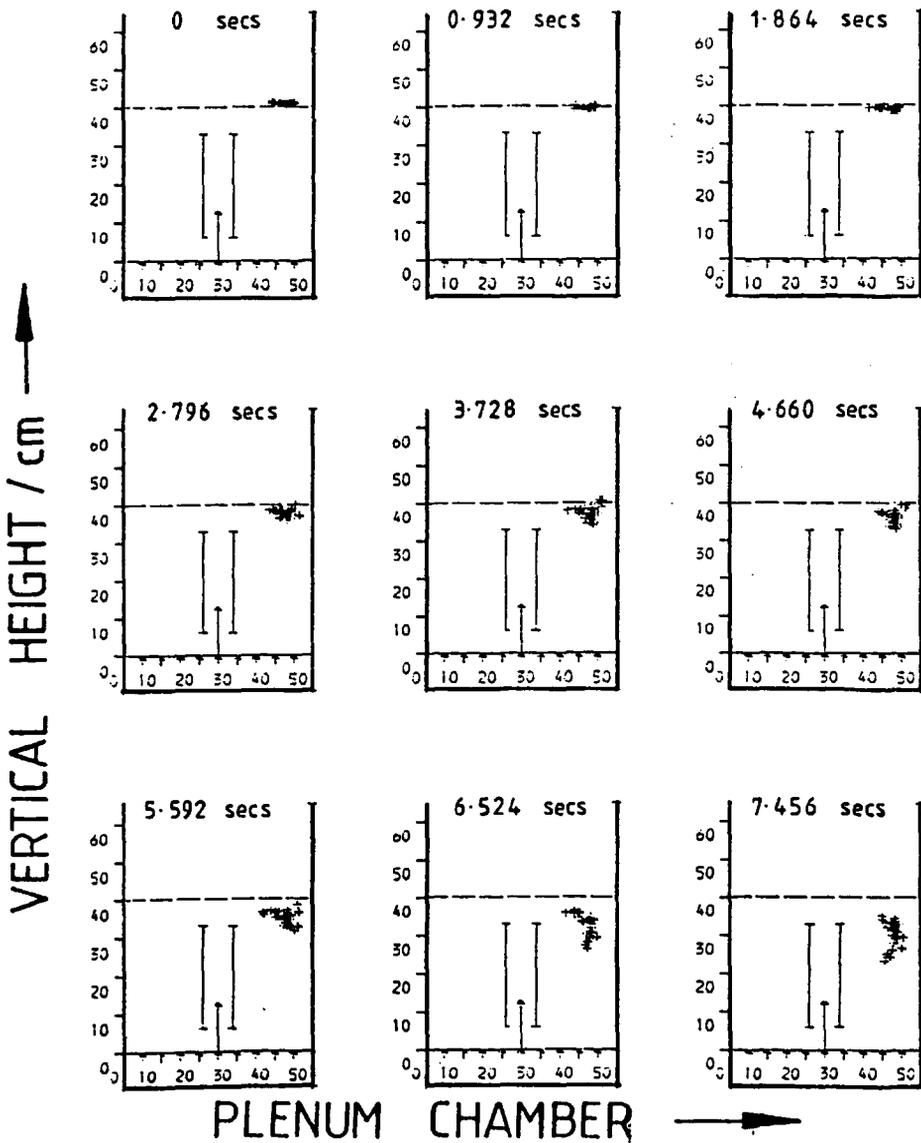
FIGURE 4·15

Draft tubes (dispersion)
$U_B = 20.83$ cm/s, $H = 40$ cm

TABLE 4·5

DISPERSION DATA – DRAFT TUBE (H = 30 cm, U_B = 16·70 cm/s)

Mean modulus values	Time/seconds							
	0·000	1·417	2·834	4·251	5·668	7·085	8·502	9·919
Displacement in X-direction, cm	0·00	0·77	0·98	0·31	0·33	0·69	0·62	1·07
Displacement in Y-direction, cm	0·00	0·04	0·58	1·27	1·34	1·50	2·09	1·88
Bandwidth in X-direction, cm	0·00	0·23	0·54	0·68	0·41	0·37	0·58	0·46
Bandwidth in Y-direction, cm	0·00	0·16	0·03	0·07	0·39	0·49	0·63	0·52

TABLE 4·6

DISPERSION DATA – DRAFT TUBE (H = 30 cm, U_B = 20·83 cm/s)

Mean modulus values	Time/seconds							
	0·000	1·417	2·834	4·251	5·668	7·085	8·502	9·919
Displacement in X-direction, cm	0·00	1·60	1·96	2·37	2·00	1·69	1·27	0·58
Displacement in Y-direction, cm	0·00	2·04	2·97	3·94	6·20	8·29	11·18	13·42
Bandwidth in X-direction, cm	0·00	0·00	0·23	0·49	0·31	0·03	0·17	0·54
Bandwidth in Y-direction, cm	0·00	0·55	0·87	1·68	2·58	3·33	3·99	4·24

TABLE 4·7

DISPERSION DATA – DRAFT TUBE (H = 30 cm, U_B = 27·50 cm/s)

Mean modulus values	Time/seconds							
	0·000	1·417	2·834	4·251	5·668	7·085	8·502	9·919
Displacement in X-direction, cm	0·00	2·36	3·71	4·38	4·48	4·25	3·13	2·47
Displacement in Y-direction, cm	0·00	2·61	3·82	5·81	7·19	9·64	11·34	13·89
Bandwidth in X-direction, cm	0·00	0·70	2·07	1·59	2·09	1·69	2·06	1·99
Bandwidth in Y-direction, cm	0·00	1·16	1·32	1·76	2·70	2·87	3·42	4·00

TABLE 4·8

DISPERSION DATA – DRAFT TUBE (H = 35cm, U_B = 20·83cm/s)

Mean modulus values	Time/seconds										
	0·000	0·932	1·864	2·796	3·728	4·660	5·592	6·524	7·456	8·388	9·320
Displacement in X-direction, cm	0·00	0·59	0·90	0·47	0·21	0·33	0·81	1·04	1·86	1·89	3·45
Displacement in Y-direction, cm	0·00	1·54	3·93	5·46	8·66	10·57	13·84	15·37	16·89	17·59	18·98
Bandwidth in X-direction, cm	0·00	0·26	0·62	0·20	0·51	1·27	1·19	1·68	1·53	1·88	0·31
Bandwidth in Y-direction, cm	0·00	0·51	2·27	2·79	4·56	5·13	5·85	6·48	6·74	5·66	5·42

TABLE 4·9

DISPERSION DATA – DRAFT TUBE (H = 40cm, U_B = 20·83 cm/s)

Mean modulus values	Time/seconds								
	0·000	0·932	1·864	2·796	3·728	4·660	5·592	6·524	7·456
Displacement in X-direction, cm	0·00	0·30	0·41	0·11	0·20	0·28	0·01	0·61	0·35
Displacement in Y-direction, cm	0·00	1·08	1·89	3·27	4·00	4·73	6·07	8·22	10·23
Bandwidth in X-direction, cm	0·00	0·16	0·09	0·17	0·03	0·32	0·21	0·19	0·64
Bandwidth in Y-direction, cm	0·00	0·20	0·24	0·64	1·32	1·30	1·28	2·35	2·60

By plotting graphs for each individual condition
(each Table), with respect to the displacement and
bandwidth in the X and Y directions against time, we obtain
Figures 4.16 to 4.20. Further splitting the curves for
displacement in each direction and plotting them against
bed height and bed velocity, we get Figures 4.21 to 4.24
(a) and (b).

The Y-directional displacement can be clearly
seen, as shown in Figure 4.21 (a) and (b), to be affected
by bed height and bed velocity. From Figure 4.21 (a) it
can be observed that the Y-directional displacement is at a
maximum when the bed height is at 35cm. This incidently is
the same as the draft tube height. Figure 4.21 (b) shows
the effect of bed velocity on the Y-directional
displacement. It can be seen that an optimum is reached
when the bed velocity approaches the minimum fluidising
velocity (U_{mf} = 25 cm/s). This can be explained by the
fact that as bed velocity is increased, so does the space
between particles, this enables the movement of the
particles to be increased. Thus increasing the dispersion
rates. When the bed velocity is increased above the
minimum fluidising velocity, the fluidising medium now
leaves the bed in the form of bubbles and that the space
between particles is not now increased significantly.
Therefore, as a consequence the dispersion rates are not
increased significantly.

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FIGURE 4·16

Dispersion Data - Draft Tube
H = 30 cm, U_B = 20.83 cm/s
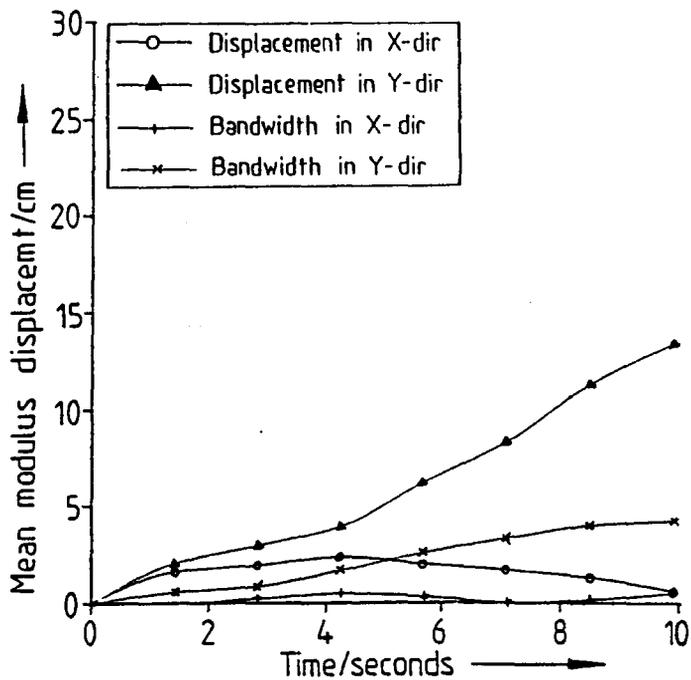
FIGURE 4·17

Dispersion Data - Draft Tube
H = 35 cm, U_B = 20.83 cm/s
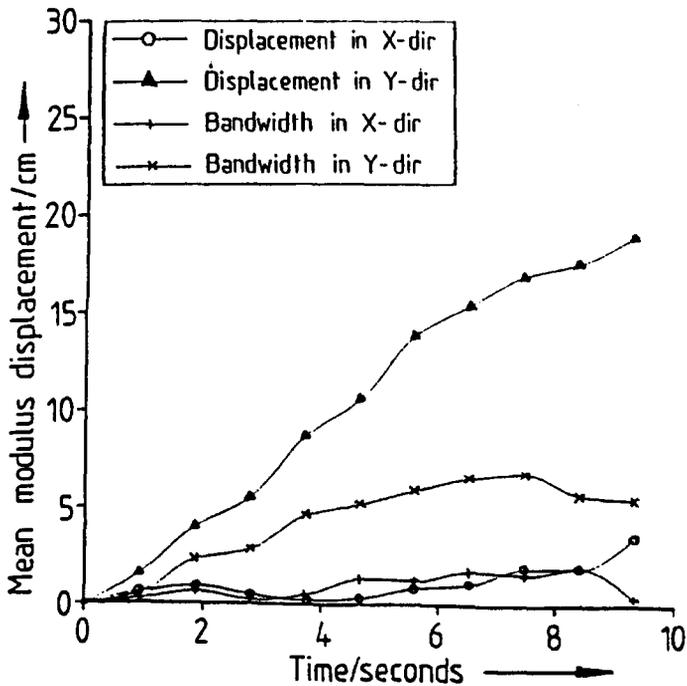
FIGURE 4·18

Dispersion Data – Draft Tube
H = 40 cm, U$_B$ = 20.83 cm/s
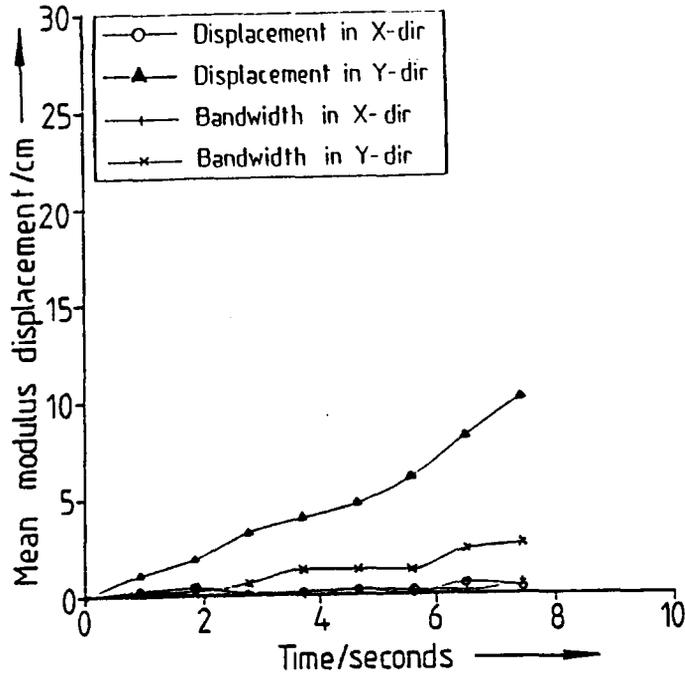
FIGURE 4·19

Dispersion Data – Draft Tube
H = 30cm, U$_B$ = 16.70 cm/s
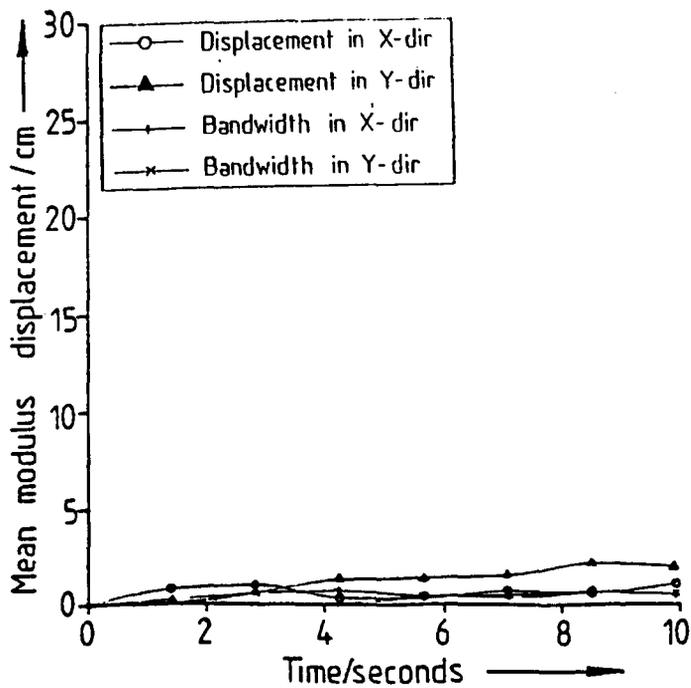
FIGURE 4·20

Dispersion Data - Draft Tube
H = 30 cm, U_B = 27.50 cm/s
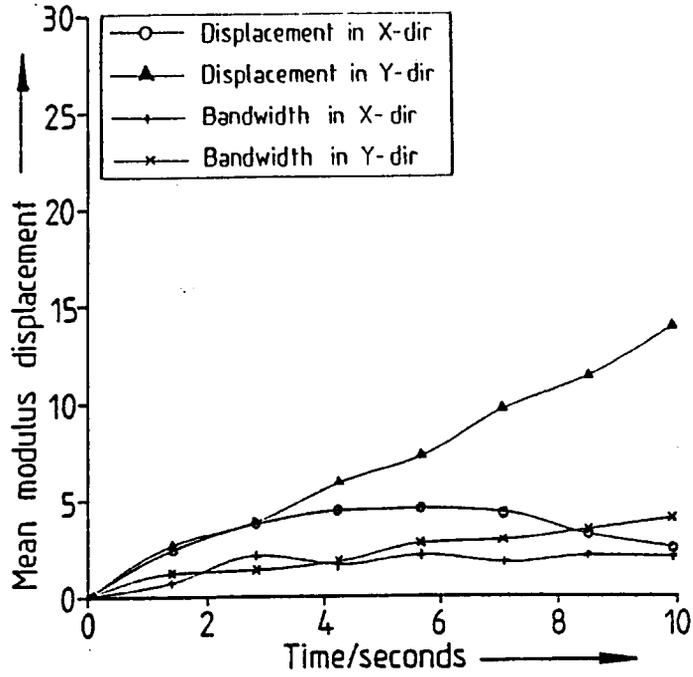
(a) Dispersion Data – Draft Tube
Effect of bed height on Y-dir. displacement.
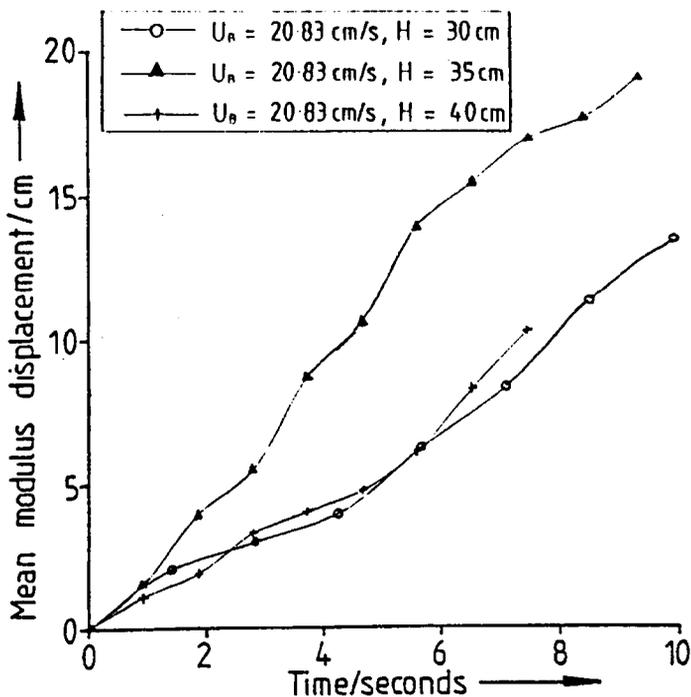
(b) Dispersion Data – Draft Tube
Effect of bed velocity on Y-dir. displacement.
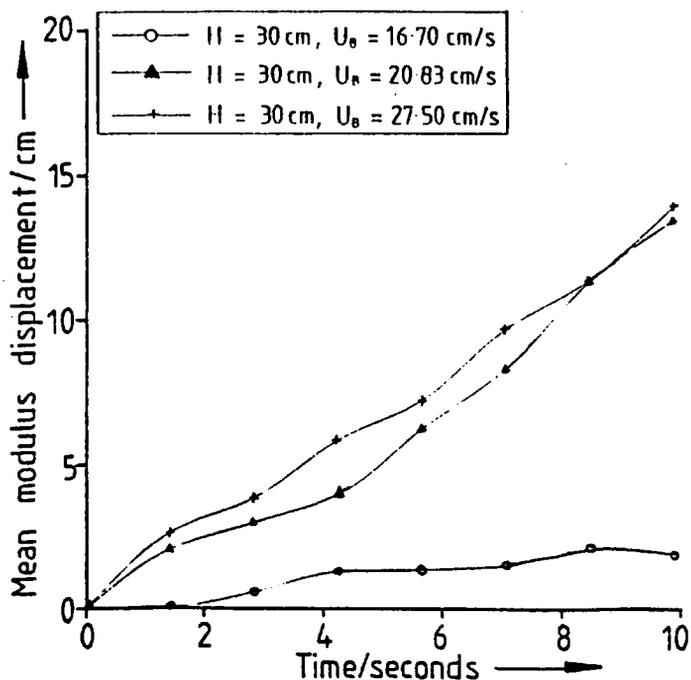
FIGURE 4·21

(a) Dispersion Data - Draft Tube
Effect of bed height on Y-dir. bandwidth.
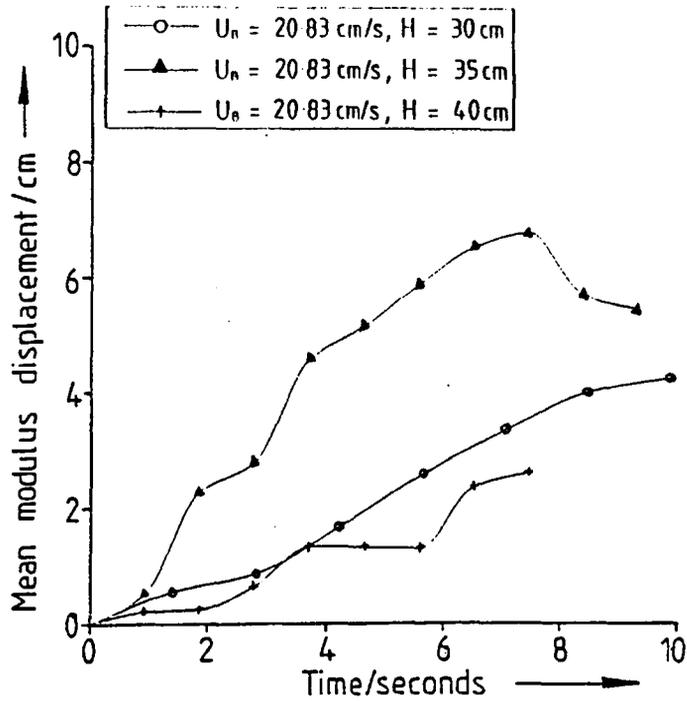
(b) Dispersion Data - Draft Tube
Effect of bed velocity on Y-dir. bandwidth.

FIGURE 4·22

(a) Dispersion Data - Draft Tube
Effect of bed height on X-dir. displacement.

(b) Dispersion Data - Draft Tube
Effect of bed velocity on X-dir. displacement.

FIGURE 4·23

(a) Dispersion Data - Draft Tube
Effect of bed height on X-dir. bandwidth.

(b) Dispersion Data - Draft Tube
Effect of bed velocity on X-dir. bandwidth.
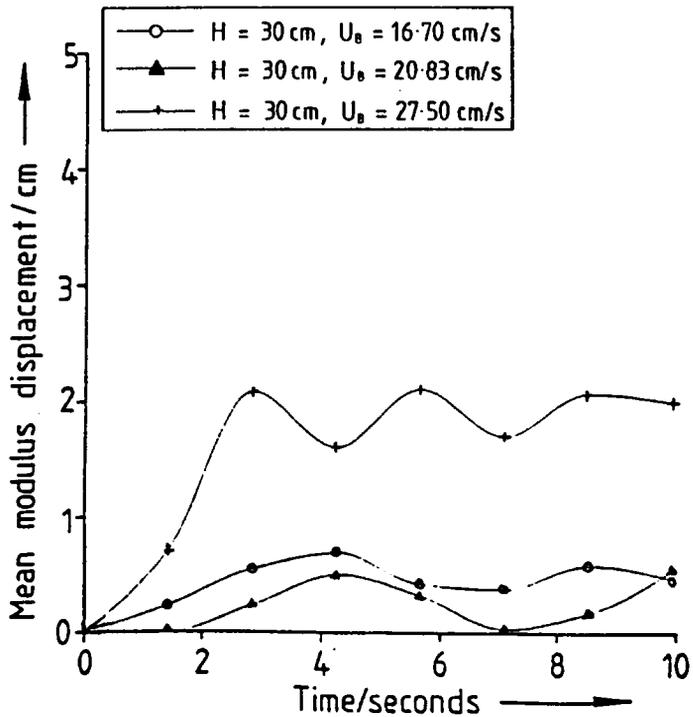
FIGURE 4·24

The effect of Y-directional bandwidth of dispersion is shown in Figure 4.22 (a) and (b). Here, as one might expect similar trends are observed as in the case of displacement, in that the optimum conditions are obtained when the bed height is equivalent to the draft tube height and when the bed velocity approaches the minimum fluidising velocity.

The X-directional displacement (sideways) given in Figure 4.23 (a) and (b), is much smaller than that for Y-direction (downwards). This could be explained by the bed geometry in that the Y-direction dimensions are greater than the X-directional dimensions. From Figure 4.23(a), it can be said that no clear trend is obtained for the displacement in the X-direction with regards to the bed height. As for the influence of bed velocity on displacement, given in Figure 4.23(b), optimum displacement rate is obtained when the bed velocity is at its maximum, in this case at 27.50cm/s. The explaination for this observation might be linked to the formation of bubbles at greater bed velocities which probably increases the movement of particles in the X-direction.

Figure 4.24 (a) and (b) shows the X-directional bandwidth with respect to the bed height and bed velocity, respectively. The X-directional bandwidth has an optimum at a bed height of 35cm (same height as the draft tube) and bed velocity of 27.50cm/s (maximum bed velocity). The

explaination for maximum bandwidth obtained for the maximum
bed velocity is the same as for the X-directional
displacement, in that the higher the bed velocity, the
greater the horizontal movement. The reason for the
maximum bandwidth obtained at a bed height of 35cm is
harder to explain.

4.5.2.2 Jet System

Figures 4.25 to 4.29 show the representation of
the digitised negatives in graphical form. Tables 4.10 to
4.14 are the corresponding numerical data obtained from the
digitised negatives. Considering Figures 4.25, 4.26 and
4.27, which show the increase in the total jet volumetric
flowrate (66, 80 and 97 lit/min respectively), while the
fluidising bed velocity is kept constant at 20.83cm/s.
Static bed height for the jet system was kept constant at
30cm throughout. Increase in jet velocity does not have a
significant effect on the dispersion rates as can be seen
from the figures. The influence of fluidising bed velocity
can be obtained by comparing Figures 4.28, 4.25 and 4.29
(16.70, 20.83 and 27.50 cm/s respectively). From the
figures it can be seen that the dispersion rate increases
with increasing bed velocity, but exact displacements and
bandwidths are not clearly obvious. To get a clear picture
graphical output was processed to yield numerical values,
which are given in Tables 4.10 to 4.14.
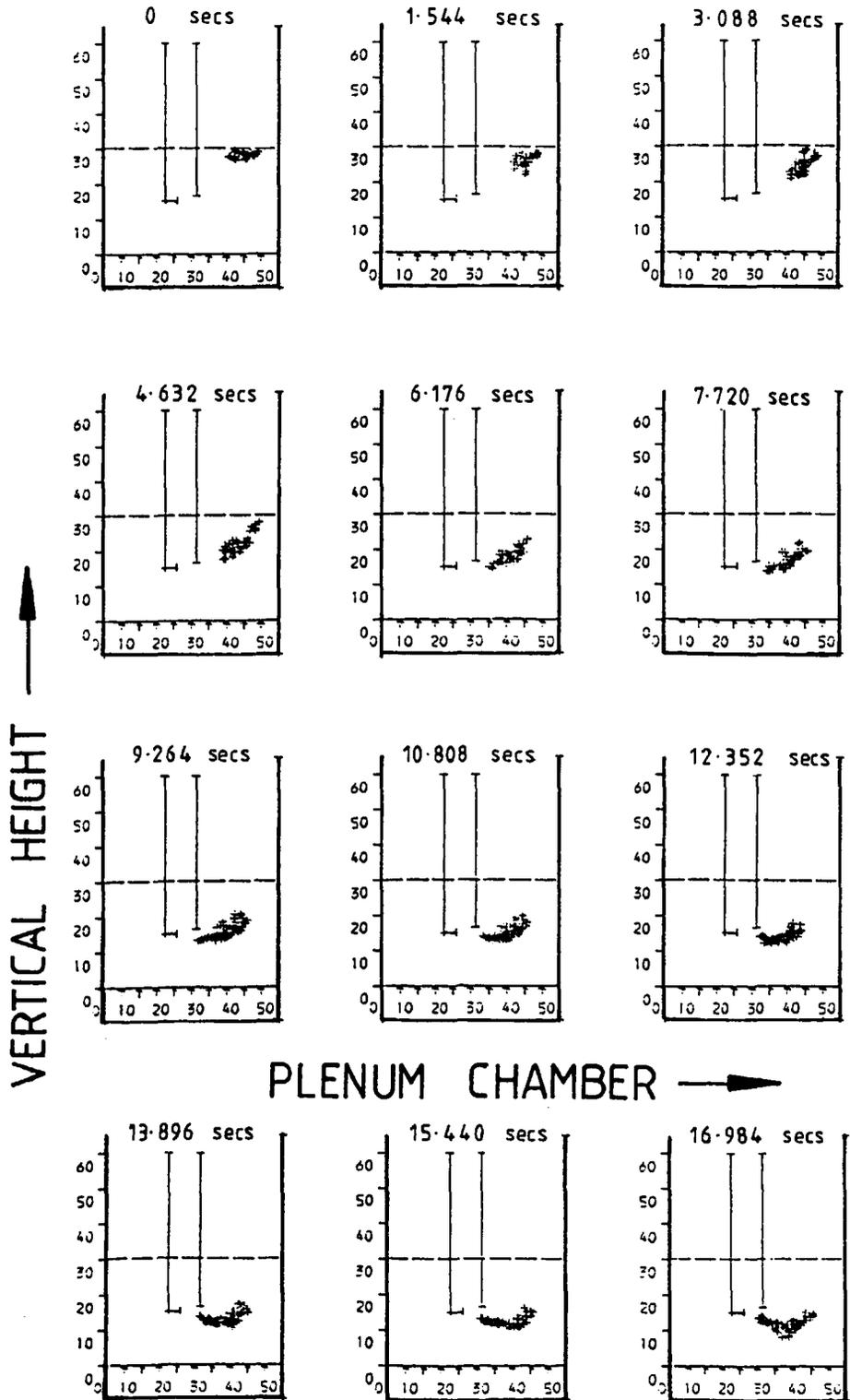
FIGURE 4·25

Jets (dispersion), $U_B = 20·83$ cm/s, $V_J = 66$ lit/min
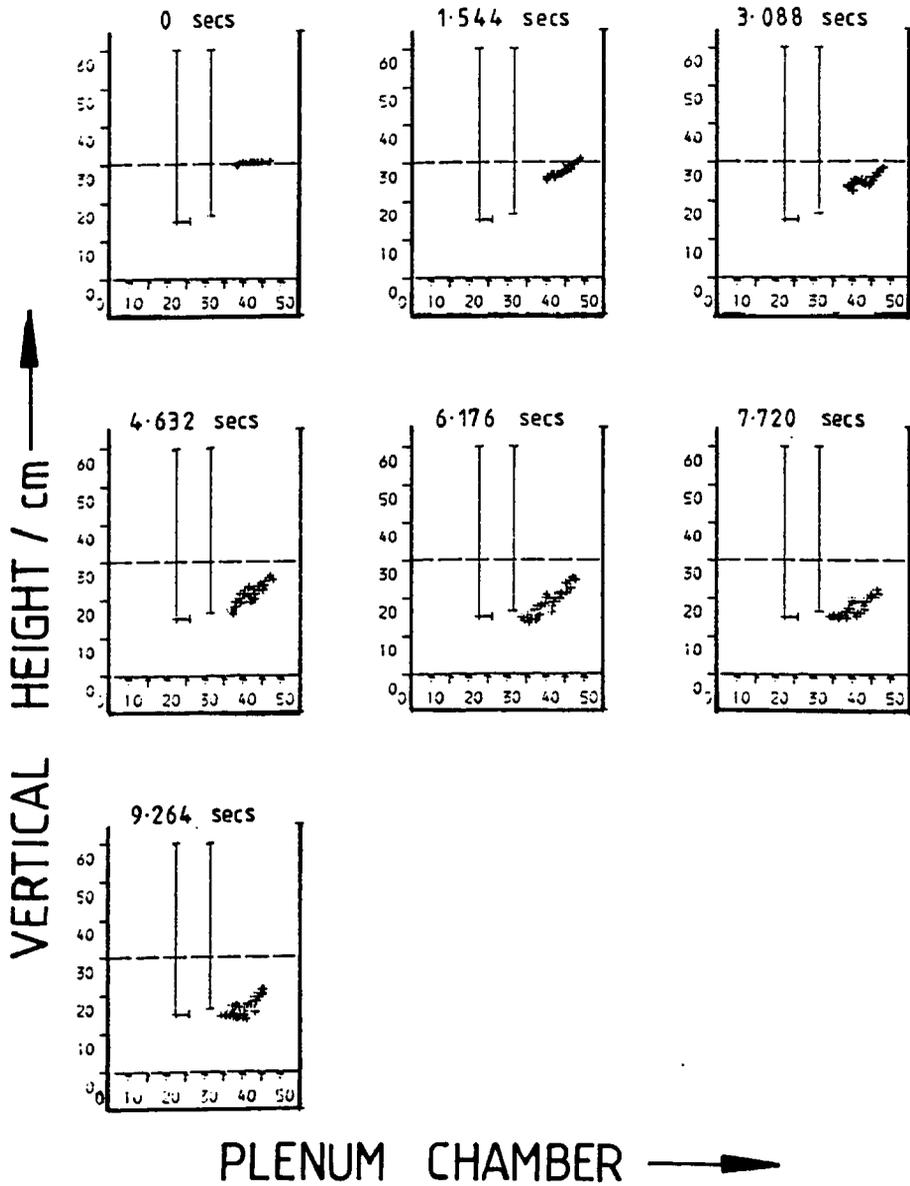
FIGURE 4·26

Jets (dispersion), $U_B = 20·83$ cm/s, $V_J = 80$ lit/min

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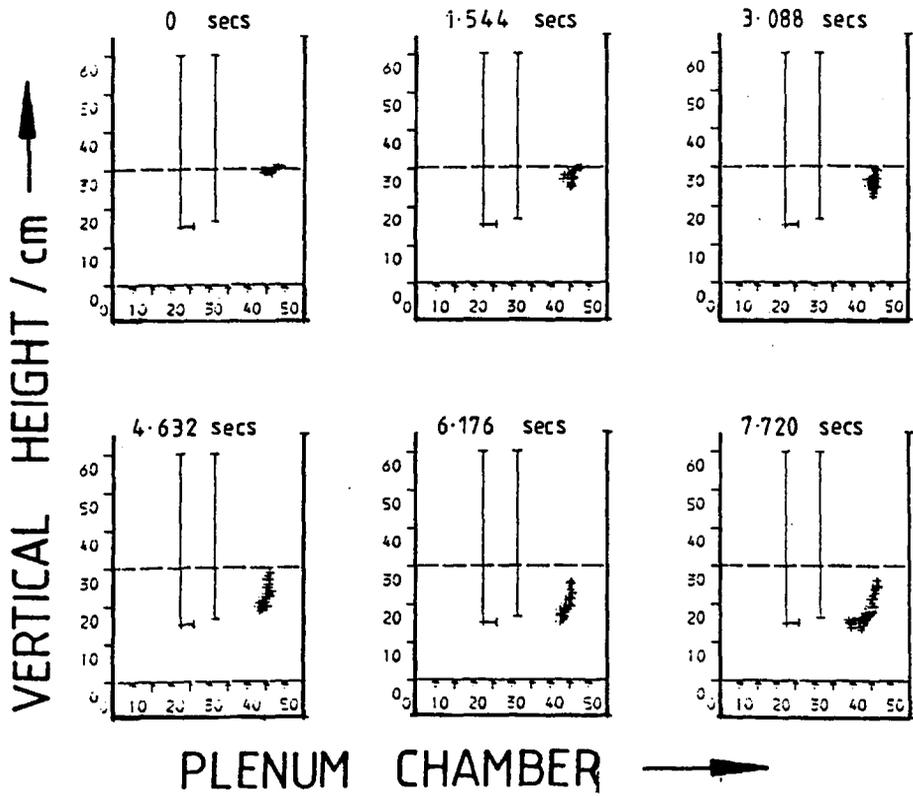
FIGURE 4·27

Jets (dispersion), U_B =20·83 cm/s, V_J =97 lit/min
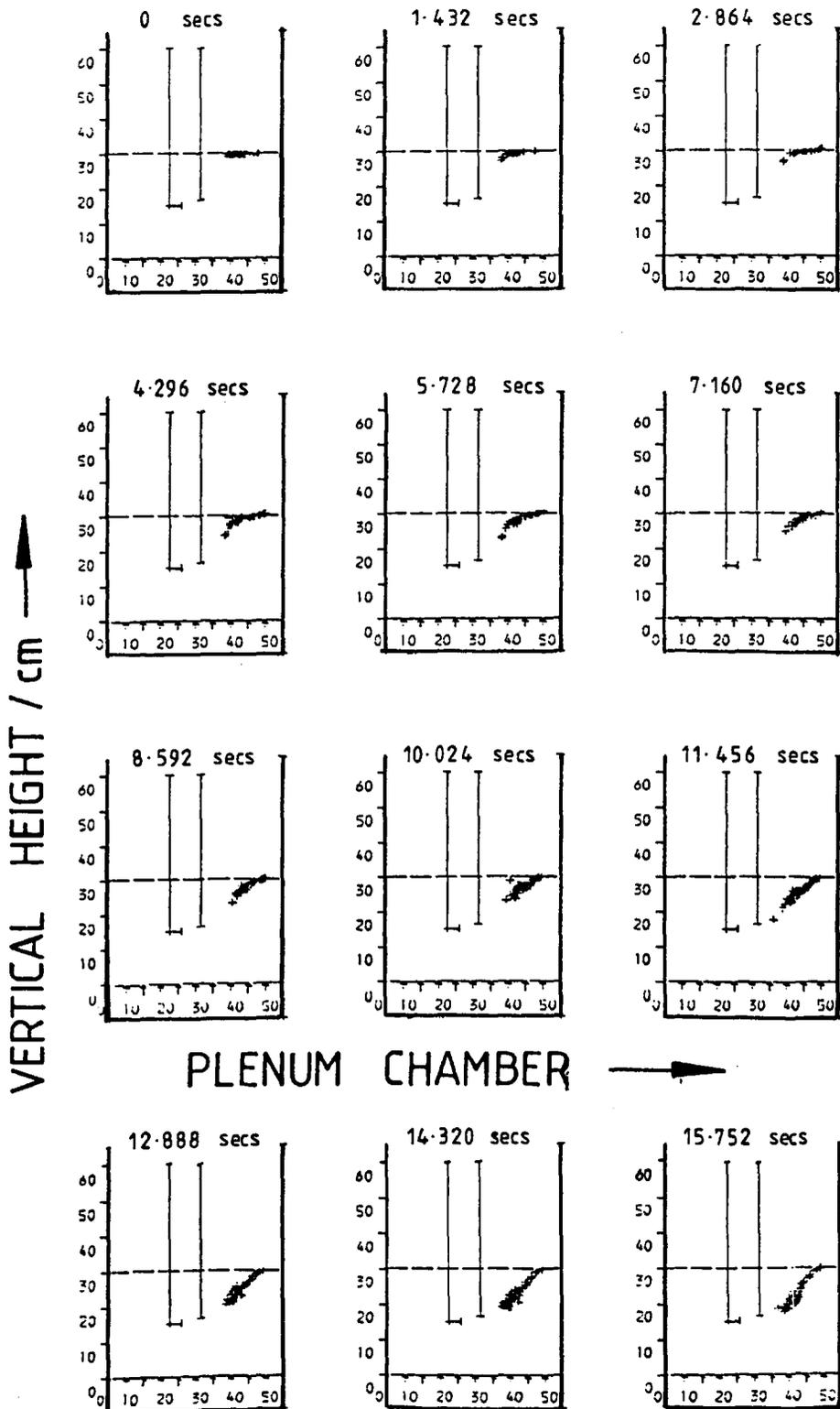
FIGURE 4·28

Jets (dispersion), $U_B = 16·70$ cm/s, $V_J = 66$ lit/min
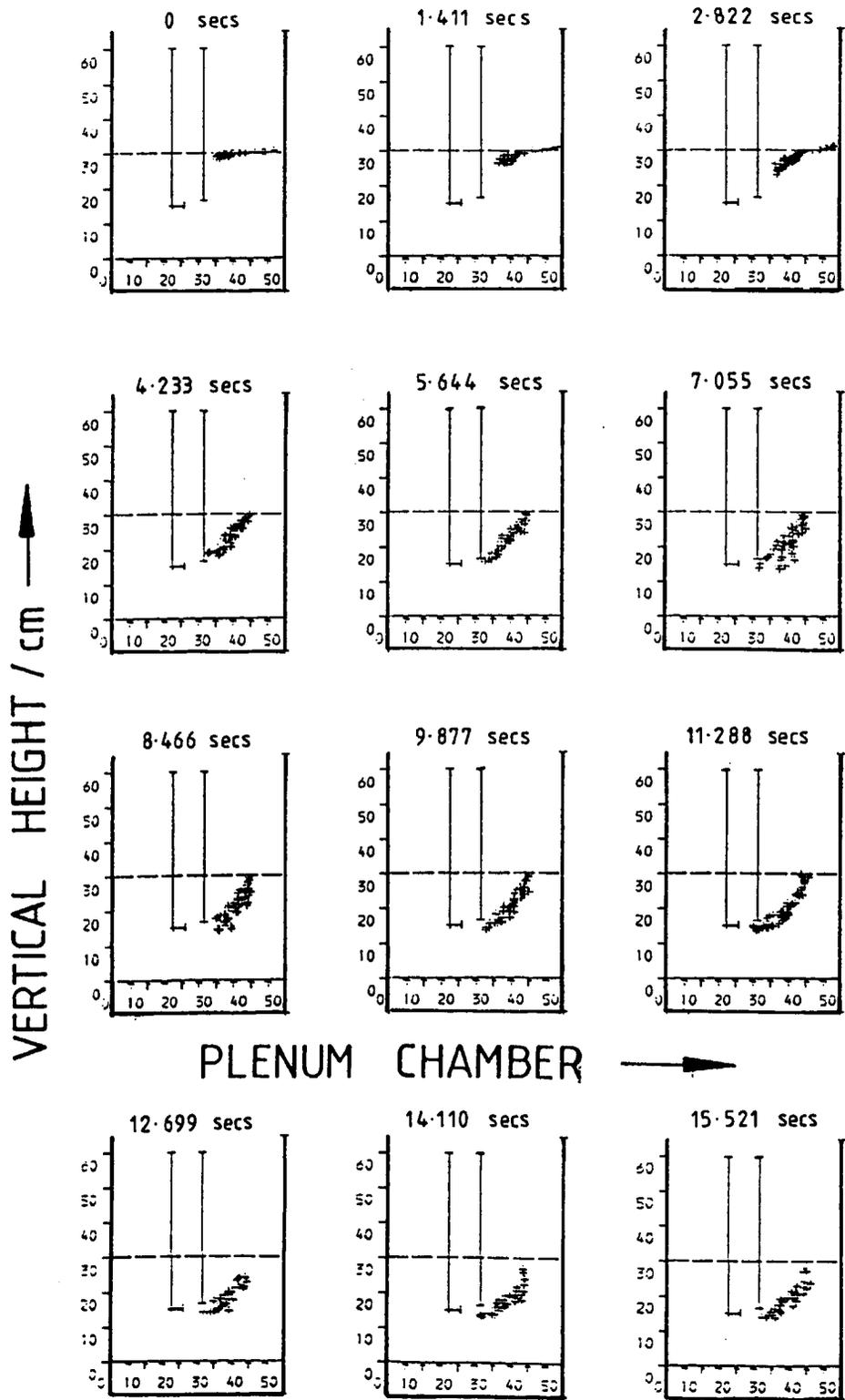
FIGURE 4·29

Jets (dispersion), $U_B = 27.50$ cm/s, $V_J = 66$ lit/min

TABLE 4·10

DISPERSION DATA-JETS(H = 30cm, U_B = 20·83cm/s, V_J = 66 lit/min)

Mean modulus values	Time/seconds						
	0·000	1·544	3·088	4·632	6·176	7·720	9·264
Displacement in X-direction, cm	0·00	0·82	0·12	1·74	3·74	3·94	4·85
Displacement in Y-direction, cm	0·00	1·79	3·45	6·39	9·61	10·91	11·93
Bandwidth in X-direction, cm	0·00	0·37	0·17	0·62	0·39	0·78	1·29
Bandwidth in Y-direction, cm	0·00	0·74	1·46	1·79	1·15	1·34	1·36

TABLE 4·11

DISPERSION DATA - JETS (H = 30cm, U_B = 20·83cm/s, V_J = 80 lit/min)

Mean modulus values	Time/seconds						
	0·000	1·544	3·088	4·632	6·176	7·720	9·264
Displacement in X-direction, cm	0·00	1·91	0·93	0·04	1·15	1·30	1·67
Displacement in Y-direction, cm	0·00	2·62	5·40	8·90	11·31	12·55	13·83
Bandwidth in X-direction, cm	0·00	0·02	0·05	0·65	1·53	0·91	0·48
Bandwidth in Y-direction, cm	0·00	1·23	1·15	2·35	3·23	1·94	2·06

TABLE 4·12

DISPERSION DATA-JETS(H = 30cm, U_B = 20·83cm/s, V_J = 97 lit/min)

Mean modulus values	Time/seconds					
	0·000	1·544	3·088	4·632	6·176	7·720
Displacement in X-direction, cm	0·00	0·47	0·64	1·42	1·79	3·02
Displacement in Y-direction, cm	0·00	1·98	3·46	7·08	9·40	11·67
Bandwidth in X-direction, cm	0·00	0·28	0·67	0·44	0·33	0·83
Bandwidth in Y-direction, cm	0·00	0·98	1·23	2·12	2·46	2·71

TABLE 4·13

DISPERSION DATA-JETS (H = 30cm, U_B = 16·70cm/s, V_J = 66 lit/min)

Mean modulus values	Time/seconds						
	0·000	1·433	2·866	4·299	5·732	7·165	8·598
Displacement in X-direction, cm	0·00	0·73	2·07	1·42	1·26	2·21	2·17
Displacement in Y-direction, cm	0·00	0·52	0·36	0·79	1·39	1·15	2·00
Bandwidth in X-direction, cm	0·00	0·07	0·94	1·12	0·70	0·34	0·35
Bandwidth in Y-direction, cm	0·00	0·35	0·72	1·23	1·39	1·15	1·55

TABLE 4·14

DISPERSION DATA-JETS (H = 30cm, U_B = 27·50cm/s, V_J = 66 lit/min)

Mean modulus values	Time/seconds						
	0·000	1·411	2·822	4·233	5·644	7·055	8·466	9·877
Displacement in X-direction, cm	0·00	1·72	2·25	0·22	0·49	0·50	0·98	0·54
Displacement in Y-direction, cm	0·00	1·22	2·11	5·47	6·25	8·08	8·07	8·33
Bandwidth in X-direction, cm	0·00	0·73	0·45	1·02	0·90	0·58	1·06	0·62
Bandwidth in Y-direction, cm	0·00·	0·89	1·45	2·79	3·04	3·84	3·46	3·77

FIGURE 4·30

Dispersion Data - Jets
H = 30 cm, U_B = 16.70 cm/s, V_J = 66 lit/min.
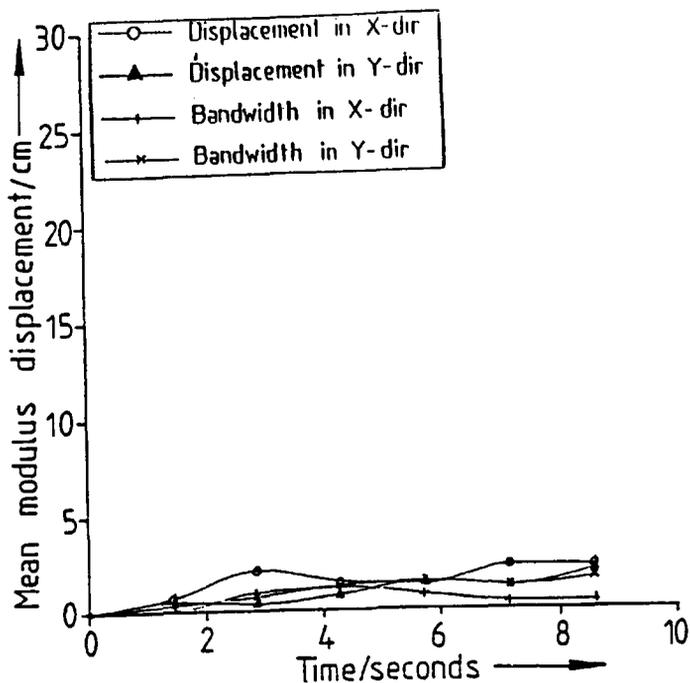
FIGURE 4·31

Dispersion Data - Jets
H = 30 cm, U_B = 20.83 cm/s, V_J = 80 lit/min.
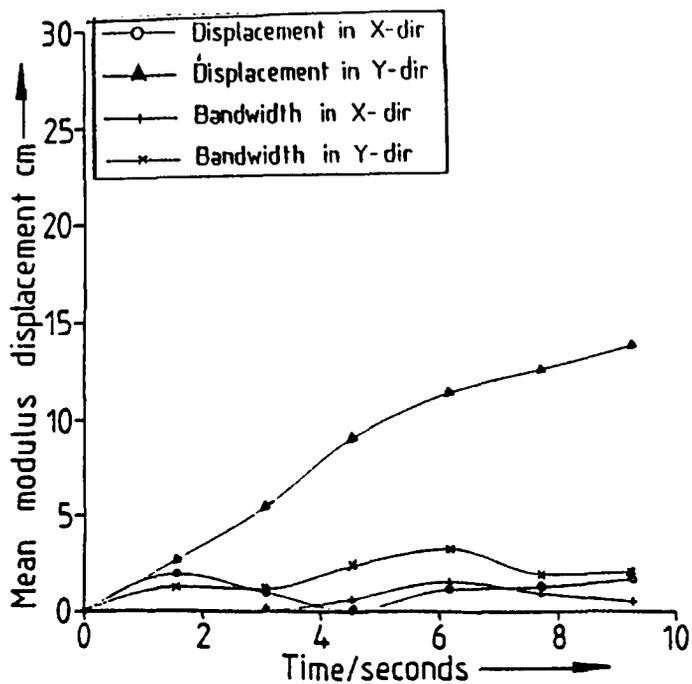
FIGURE 4·32

Dispersion Data - Jets

H = 30 cm, U$_B$ = 20.83 cm/s, V$_J$ = 97 lit/min.
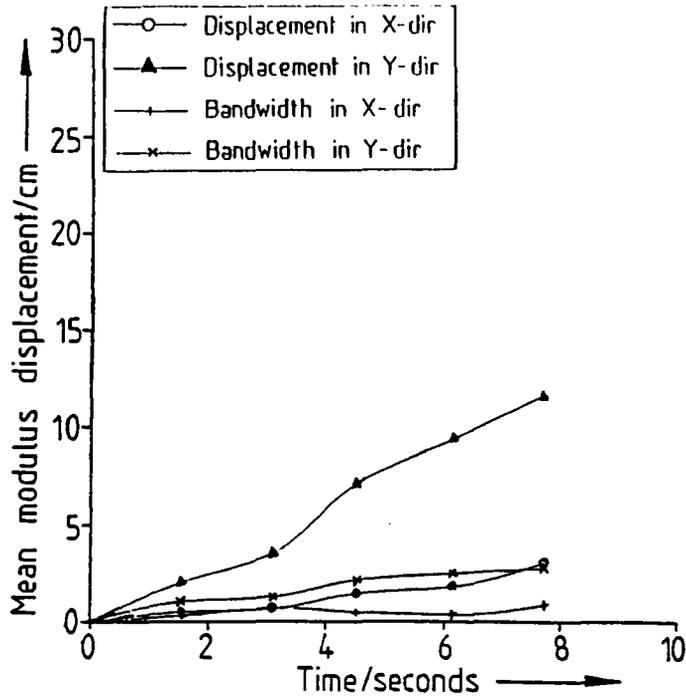
FIGURE 4·33

Dispersion Data - Jets

H = 30 cm, U$_B$ = 20.83 cm/s, V$_J$ = 66 lit/min.
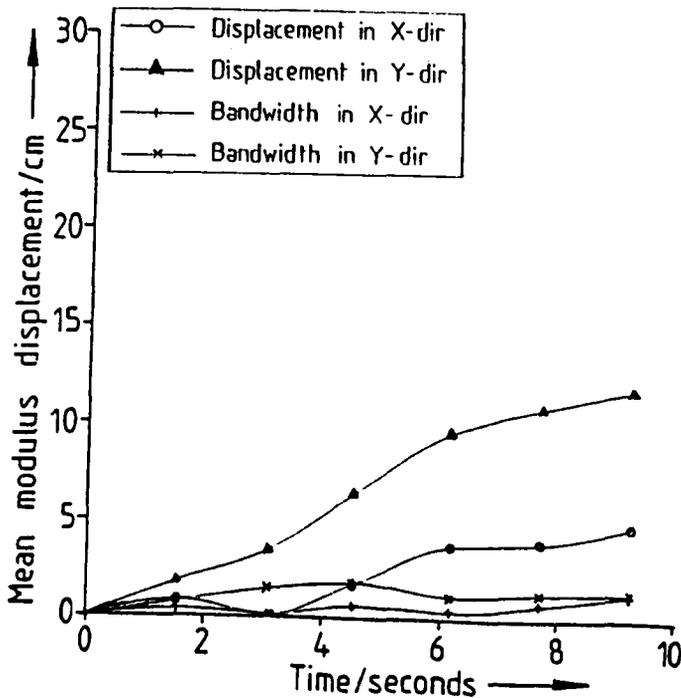
FIGURE 4·34

Dispersion Data - Jets
H = 30 cm, U$_B$ = 27.50 cm/s, V$_J$ = 66 lit/min.
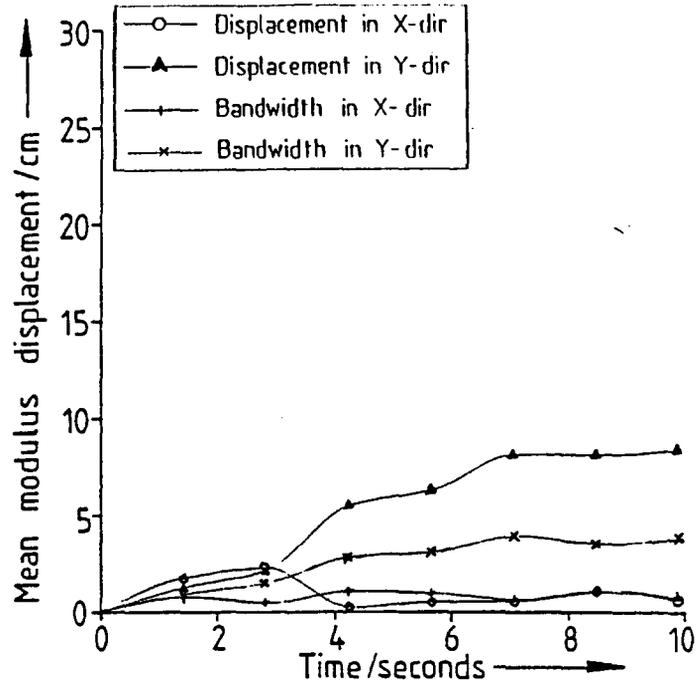
(a) Dispersion Data - Jets
Effect of bed velocity on Y-dir. displacement.
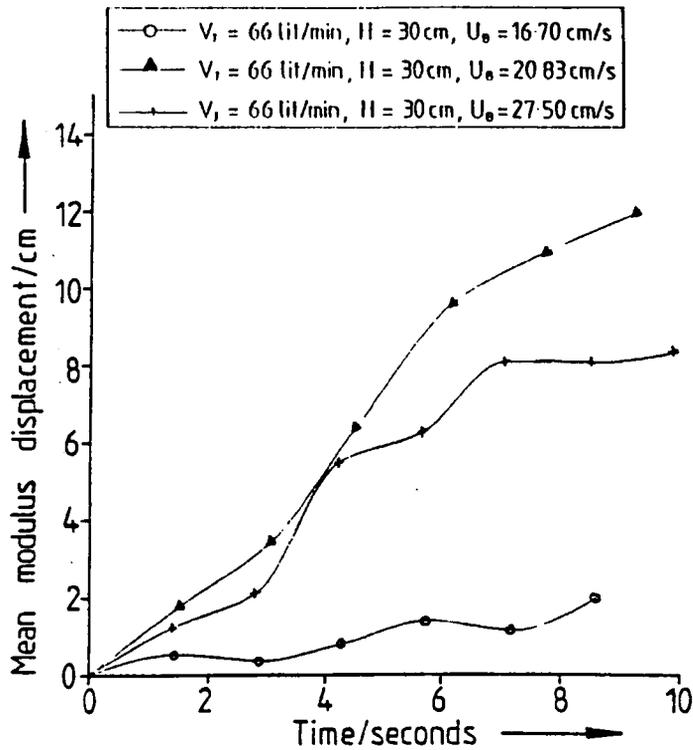
(b) Dispersion Data - Jets
Effect of jet velocity on Y-dir. displacement.
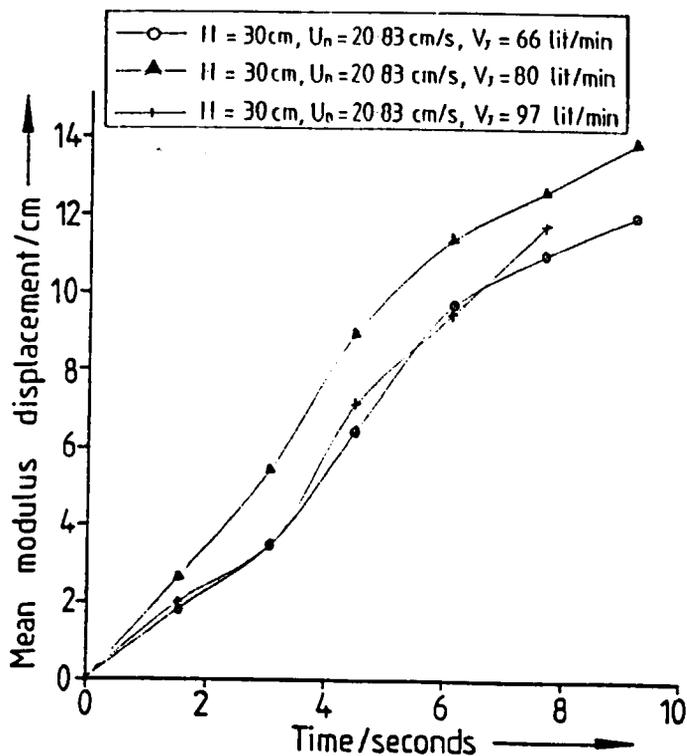
FIGURE 4·35

(a) Dispersion Data - Jets
Effect of bed velocity on Y-dir. bandwidth.
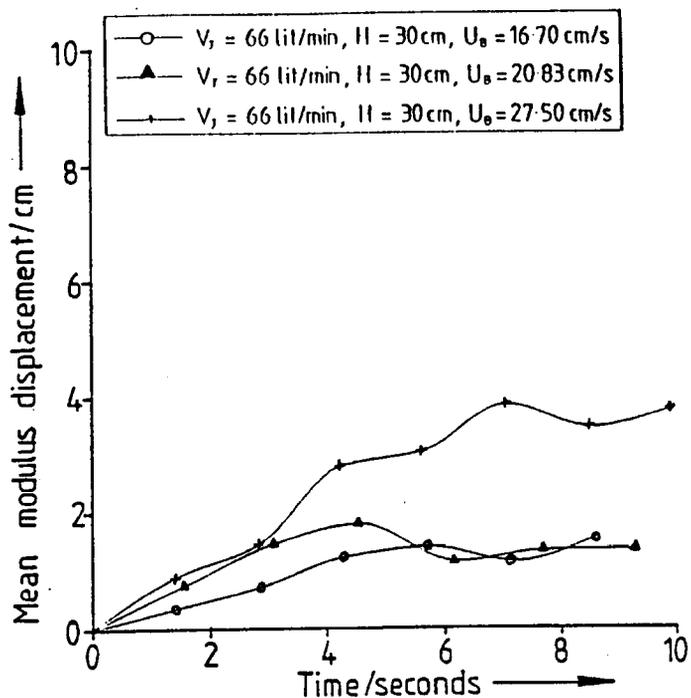
(b) Dispersion Data - Jets
Effect of jet velocity on Y-dir. bandwidth.
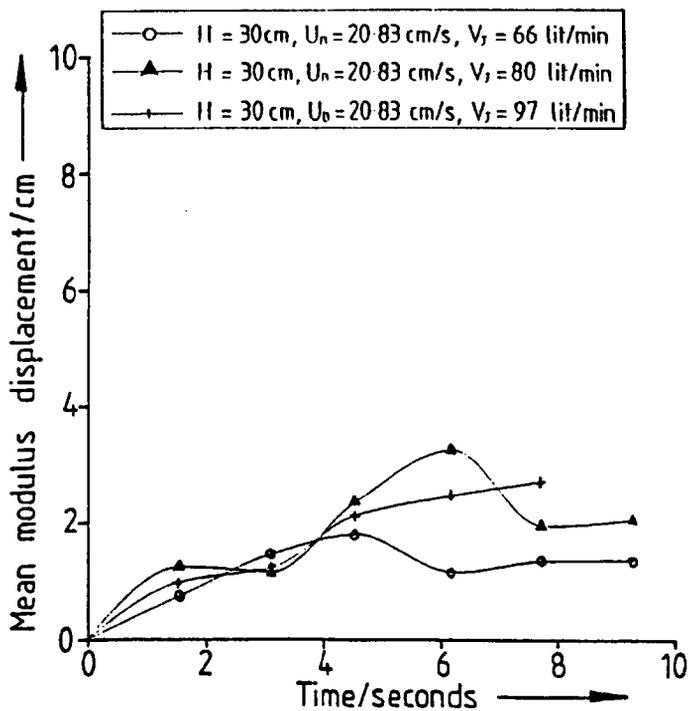
FIGURE 4·36

(a) Dispersion Data – Jets
Effect of bed velocity on X-dir. displacement.
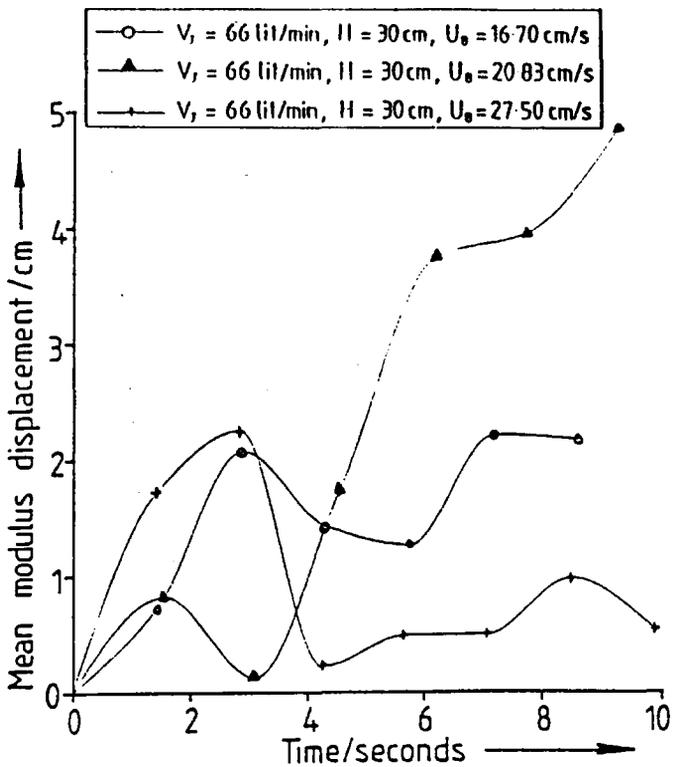
(b) Dispersion Data – Jets
Effect of jet velocity on X-dir. displacement.
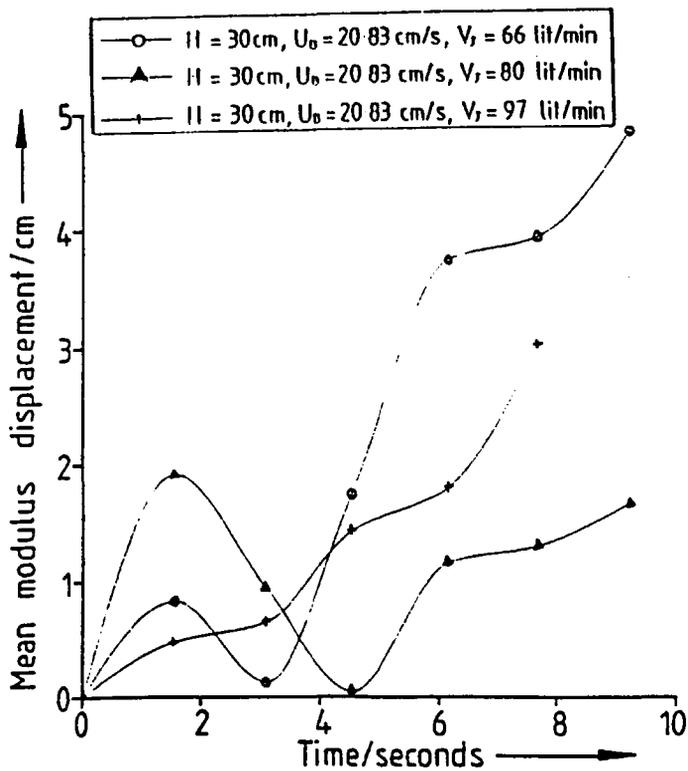
FIGURE 4·37

(a) Dispersion Data - Jets
Effect of bed velocity on X-dir. bandwidth.
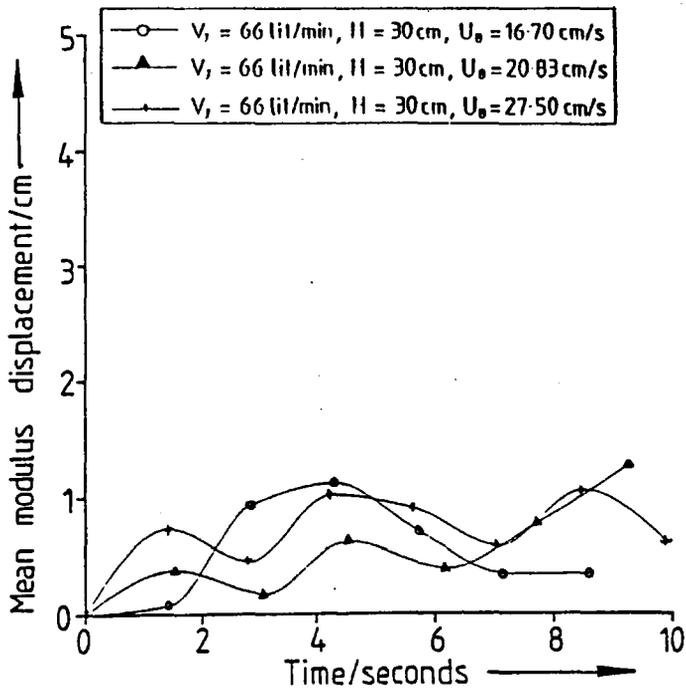
(b) Dispersion Data - Jets
Effect of jet velocity on X-dir. bandwidth.

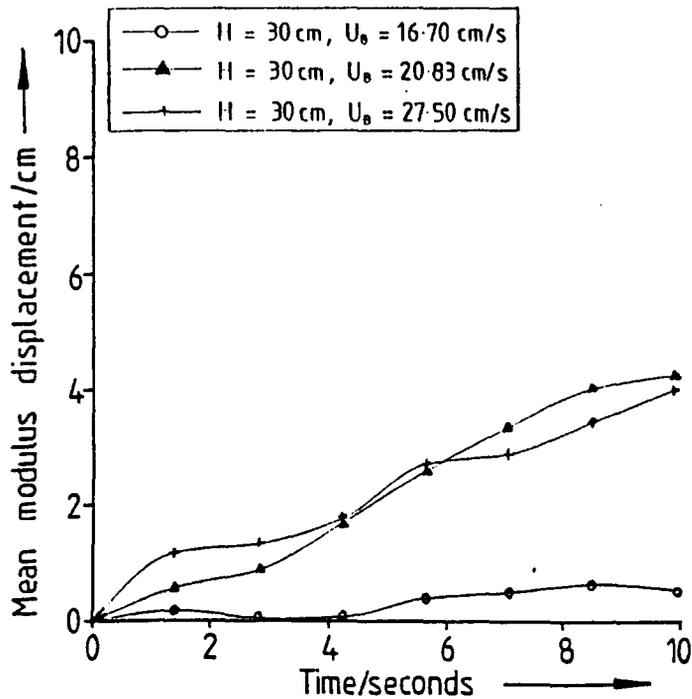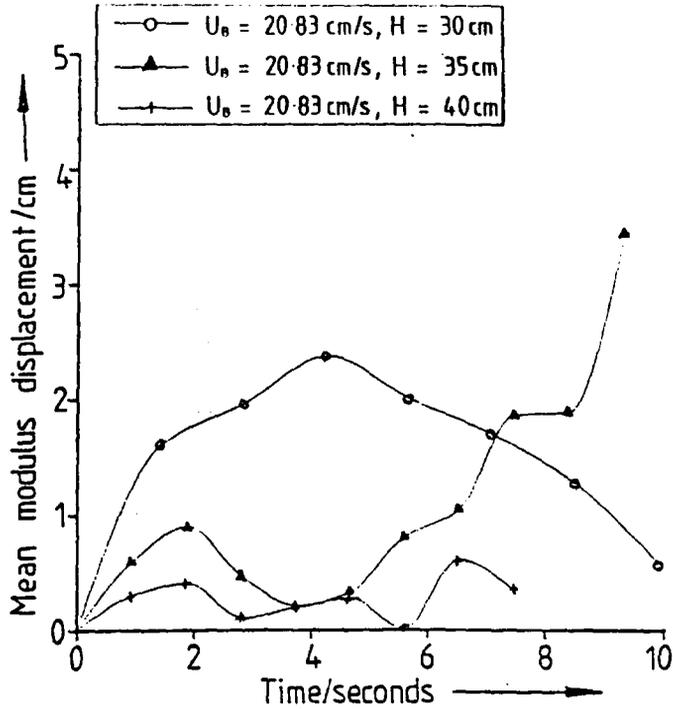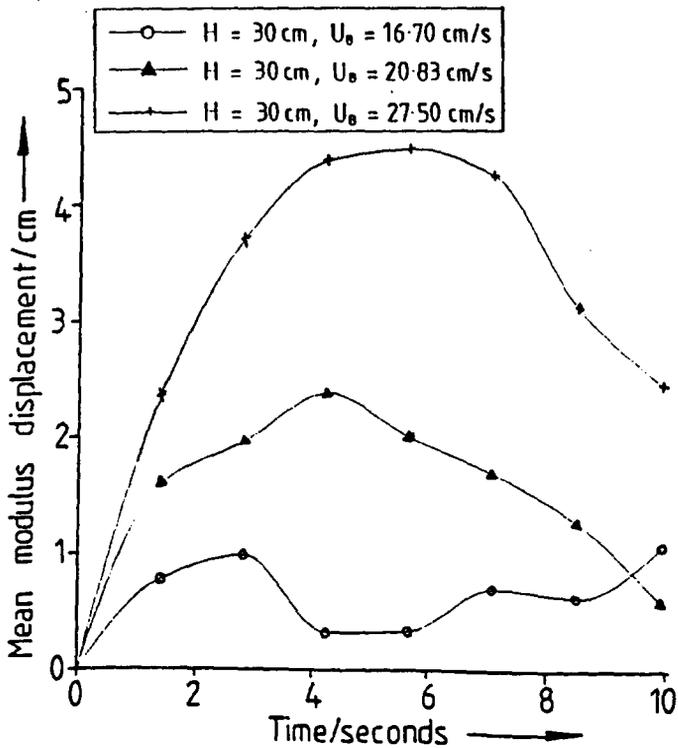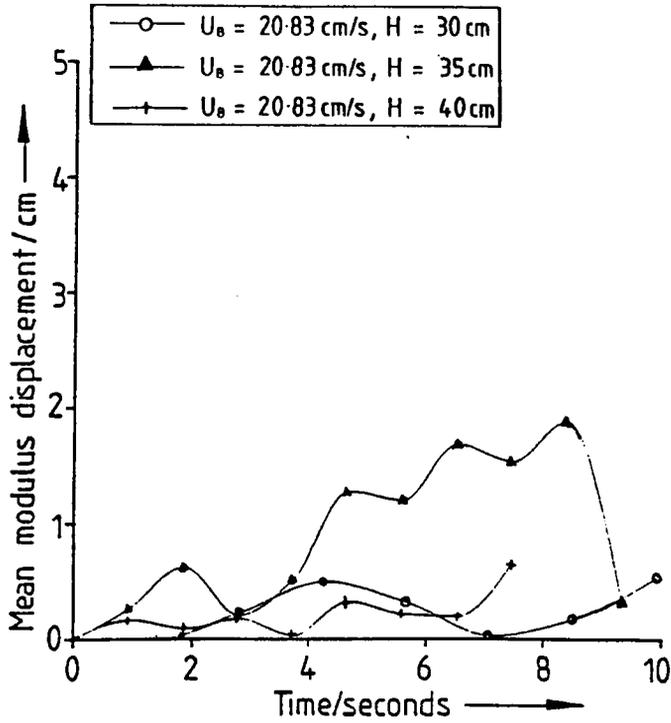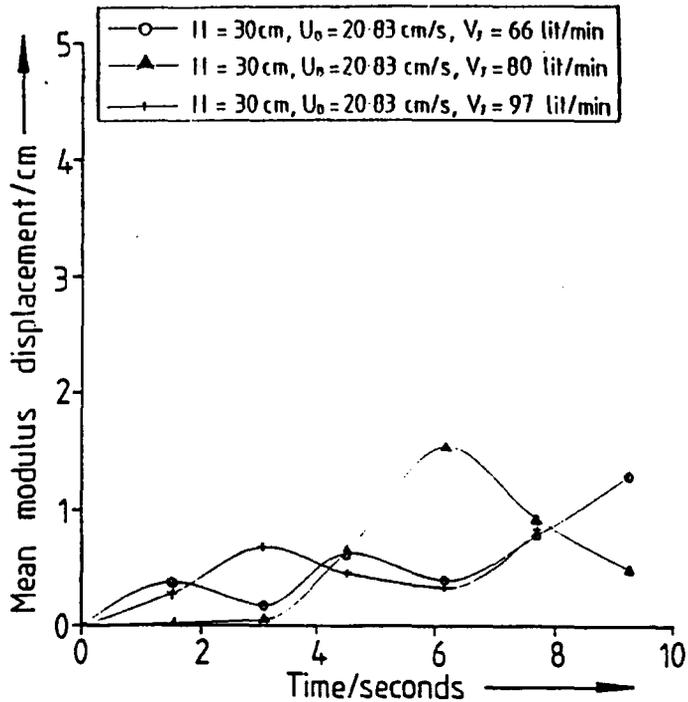FIGURE 4·38

By plotting graphs for each individual condition, as previously, we obtain Figures 4.30 to 4.34. These figures show that the highest displacement is in the Y-direction, except in the case of Figure 4.30 (U_B = 16.70cm/s and V_J = 66lit/min), where displacement and bandwidth is minimum and indistinguishable. Further splitting of the curves for displacement and bandwidth in each direction and plotting these against fluidising bed velocity and jet velocity, yields Figures 4.35 to 4.38 (a) and (b).

Considering the Y-directional displacement given in Figure 4.35 (a) and (b). The influence of fluidising bed velocity, Figure 4.35(a), shows that there is a similar trend to that observed in the case of the draft tube system, in that increasing the bed velocity, dispersion rates increase. They increase to an optimum as the bed velocity approaches the minimum fluidising velocity. Further increases have little effect on the Y-directional displacement. The reason for this is the same as previously explained. The effect of jet velocity, Figure 4.35(b), shows that increasing the total volumetric flowrate from 66 to 97 lit/min has little effect on the Y-directional displacement.

Y-directional bandwidth is given in Figure 4.36 (a) and (b). Increase in fluidising bed velocity has little or no affect until the the velocity is higher than

the minimum fluidising velocity (U_{mf} = 25cm/s). This is probably due to the increase in the movement of the bed particles as the velocity is increased above the minimum fluidising velocity.

The X-directional displacement given in Figure 4.36 (a) and (b) shows random fluctuations, no real trend is observed. The same applies to the X-directional bandwidth, Figure 4.38 (a) and (b).

4.5.2.3 Conclusions

(a) DRAFT TUBE SYSTEM

(i) Bed height is an important parameter regarding the dispersion rates. It was found from the experimental results that a maximum dispersion rate was obtained with the bed height eqviualent to the height of the draft tube.

(ii) Fluidising bed velocity was found to increase the dispersion rates but only to an optimum where it approaches the minimum fluidising velocity. Further increases of the fluidising bed velocity have no significant effect on the dispersion rates.

(b) JET SYSTEM

(i) Fluidising bed velocity for the jet system also displaced the same trend as the draft tube system, in that the optimum dispersion rates were obtained with

fluidising bed velocities approaching the minimum fluidising velocity.

(ii) The effect of jet velocity on dispersion rates is not very significant. Increasing the total jet volumetric flowrate from 66 to 97 lit/min (40% increase), there was no significant change in the dispersion rate. Perhaps increasing jet volumetric flowrate dramatically might have had an affect.

4.5.3 Particle Velocity Vectors

The particle velocity vectors are an important measurement regarding any hydrodynamic study of fluidising beds with or without draft tubes or jets. These measurements help to establish circulation rates and circulation patterns regarding solid flow.

In this study particle velocity vectors have been investigated, for the following :

(i) Free bed without internals.

(ii) Draft tube system.

(iii) Jet system.

4 .5.3.1 Free bed without internals

The particles velocity vectors were initially investigated in the case of a free bed. The purpose of

this study was two fold. Firstly to investigate the circulation rates, and secondly to show that the circulation patterns are what one would expect in a fluidisied bed. For this study the static bed height was kept constant at 30cm and the fluidising velocities were varied (30, 50, 75, 100 and 125 cm/s). Table 4.14 shows all the processed data, while Figures 4.39 to 4.42 show the particle velocity vectors for 30, 50, 100 and 125 cm/s, respectively.

From the figures it is clear that their is an almost random circulation patterns in all the cases and the only difference being that the velocity vectors get larger as the fluidising velocities are increased. This is not a new observation but only confirms the fact that this experimental bed does behave like a fluidised bed. When different circulation patterns are obtained with draft tubes and jets, the configuration of the basic experimental bed has no influence, but the only influence is from the bed internals.

Figure 4.43 shows the effect of fluidising bed velocity on average particles velocities. Increasing the fluidising velocity obviously increases the particle velocities.

TABLE 4·14

Free-Bed Particle Velocity Vectors

	H = 30 cm U_B = 0·30 m/s	H = 30 cm U_B = 0·50 m/s	H = 30 cm U_B = 0·75 m/s	H = 30 cm U_B = 1·00 m/s	H = 30 cm U_B = 1·25 m/s
Max.(av.) velocity	2·52 cm/s	3·83 cm/s	8·62 cm/s	7·19 cm/s	12·98 cm/s
Min.(av.) velocity	2·37 cm/s	2·34 cm/s	4·54 cm/s	6·10 cm/s	11·25 cm/s
Average velocity	2·44 cm/s	3·08 cm/s	6·50 cm/s	6·65 cm/s	12·11 cm/s
S.D. (av.)	± 0·08 cm/s	± 0·66 cm/s	± 1·77 cm/s	± 0·52 cm/s	± 0·41 cm/s
Up-right (av. vel.) % particles	2·36 cm/s 29·9 %	3·25 cm/s 22·4 %	6·23 cm/s 23·9 %	6·69 cm/s 28·3 %	12·07 cm/s 30·2 %
Up-left (av. vel.) % particles	2·69 cm/s 24·1 %	3·14 cm/s 42·1 %	6·71 cm/s 37·0 %	7·04 cm/s 33·7 %	12·73 cm/s 37·4 %
Down left (av. vel.) % particles	2·44 cm/s 24·1 %	2·64 cm/s 18·4 %	6·30 cm/s 23·1 %	7·37 cm/s 15·2 %	11·67 cm/s 16·2 %
Down-right (av. vel.) % particles	2·29 cm/s 21·9 %	3·17 cm/s 17·1 %	6·67 cm/s 16·0 %	5·54 cm/s 22·8 %	11·17 cm/s 16·2 %

FIGURE 4·39
Particles – Free bed, U_B = 30 cm/s

FIGURE 4·40
Particles – Free bed, U_B = 50 cm/s

FIGURE 4·41
Particles – Free bed, U_B = 100 cm/s

FIGURE 4·42
Particles – Free bed, U_B = 125 cm/s

FIGURE 4·43

Effect of Fluidising Bed Velocity on Average Particle Velocities in a Free Bed
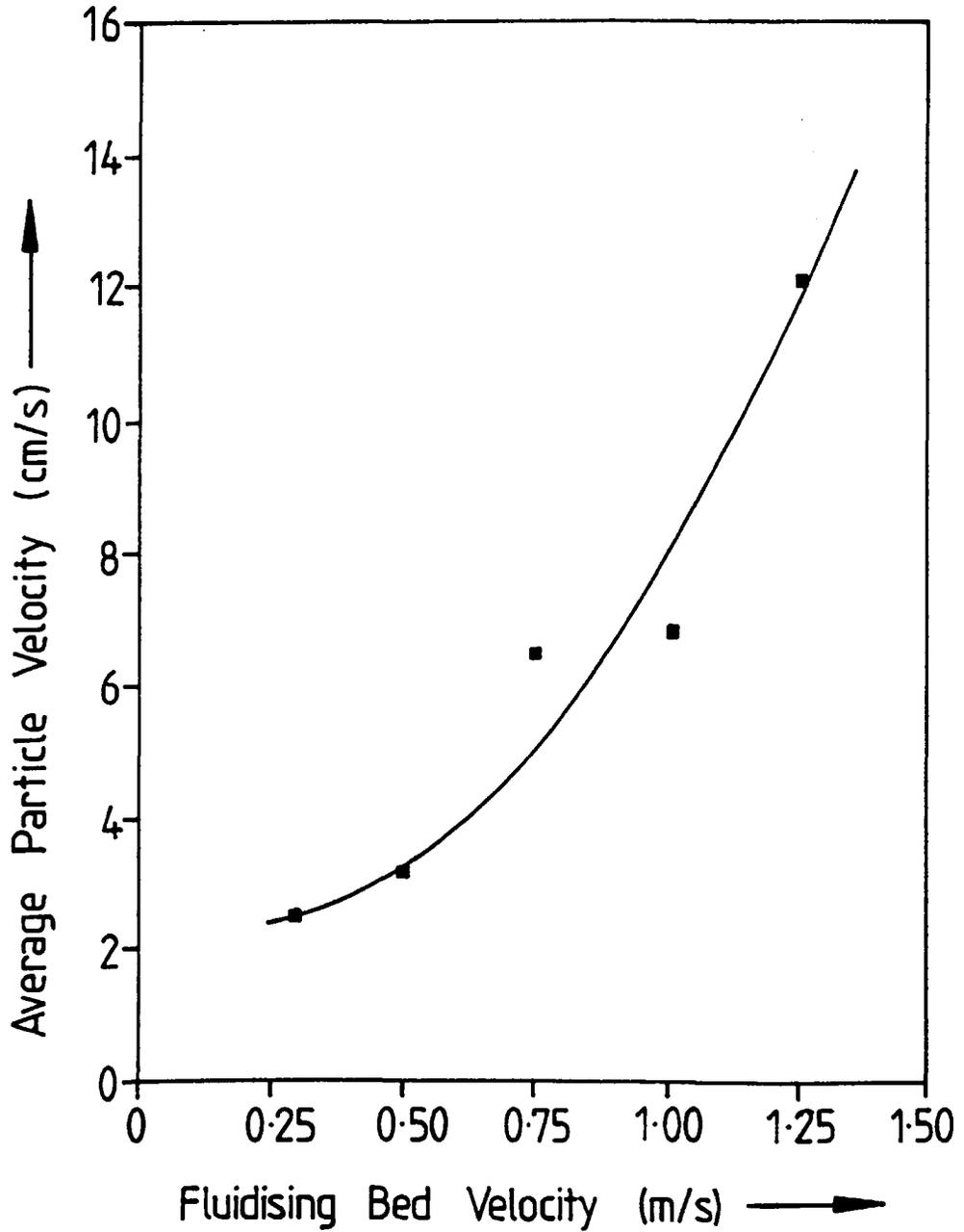
4.5.3.2 Draft tube system

This study investigated the effect of the bed height and fluidising bed velocity on the particle velocities. The results obtained from the experiments are given in Table 4.15.

Figures 4.44 to 4.48 (a, b & c) show the particle velocity vectors, concentration of the velocity vectors and the 3-D particle velocity concentration configuration for a better visual concept.

Considering the effect of bed height, Figures 4.45, 4.46 and 4.47 (a, b & c), where the fluidising velocity is kept constant at 20.83cm/s. It is obvious from the figures that there is a definite circulation pattern. the comparison of the concentration profiles and the 3-D figures shows that the bandwidth of particle circulation increases with bed height.

Bandwidth of particle circulation and patterns are more significantly altered when the fluidising bed velocity is varied keeping the bed height constant at 30cm, Figures 4.44, 4.45 and 4.48 (a, b & c). A comparison of the figures shows clearly that the bandwidth of particle circulation increases with increasing bed velocity.

Figures 4.49 and 4.50 show the effect of bed velocity and height, respectively, on the percentage of particles in the four directions. It is obvious, in

comparison to particle percentages in a free bed Figure
4.51, that the draft tube system induces a definite
particle circulation.

TABLE 4·15

Draft-Tube Particle Velocity Vectors

	H = 30 cm U_B = 16·70 cm/s	H = 30 cm U_B = 20·83 cm/s	H = 35 cm U_B = 20·83 cm/s	H = 40 cm U_B = 20·83 cm/s	H = 30 cm U_B = 27·50 cm/s
Max.(av.) velocity	4·94 cm/s	4·07 cm/s	5·33 cm/s	4·97 cm/s	4·35 cm/s
Min. (av.) velocity	4·70 cm/s	3·55 cm/s	4·62 cm/s	4·46 cm/s	3·84 cm/s
Average velocity	4·82 cm/s	3·81 cm/s	4·98 cm/s	4·73 cm/s	4·09 cm/s
S.D. (av.)	± 0·12 cm/s	± 0·26 cm/s	± 0·35 cm/s	± 0·24 cm/s	± 0·25 cm/s
Up-right (av.) % particles	15·28 cm/s 1·2 %	4·27 cm/s 7·2 %	5·40 cm/s 2·3 %	6·35 cm/s 3·5 %	4·00 cm/s 6·2 %
Up-left (av.) % particles	4·33 cm/s 2·0 %	4·08 cm/s 2·8 %	2·69 cm/s 1·0 %	4·27 cm/s 9·0 %	4·83 cm/s 2·5 %
Down-left (av.) % particles	4·74 cm/s 42·1 %	3·76 cm/s 43·9 %	4·70 cm/s 46·7 %	4·58 cm/s 38·5 %	4·53 cm/s 33·8 %
Down-right (av.) % particles	4·67 cm/s 54·7 %	3·76 cm/s 46·1 %	5·26 cm/s 50·0 %	4·83 cm/s 49·0 %	3·81 cm/s 57·5 %

FIGURE 4·44 (a)

Particle velocity vectors – Draft tube system

$U_B = 16·70 \, cm/s$, $H = 30 \, cm$

FIGURE 4·44 (b)

Particle concentrations – Draft tube system
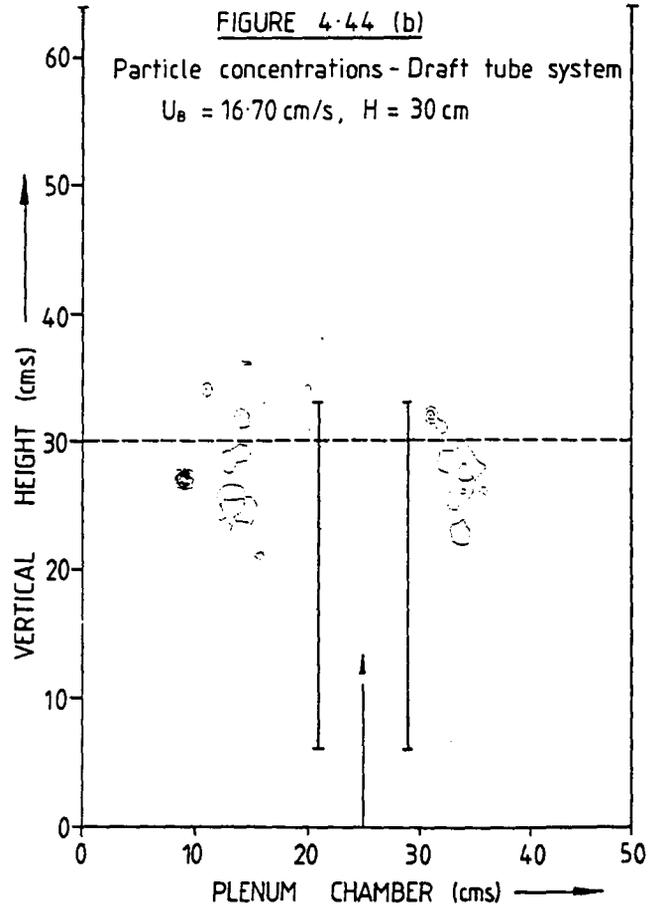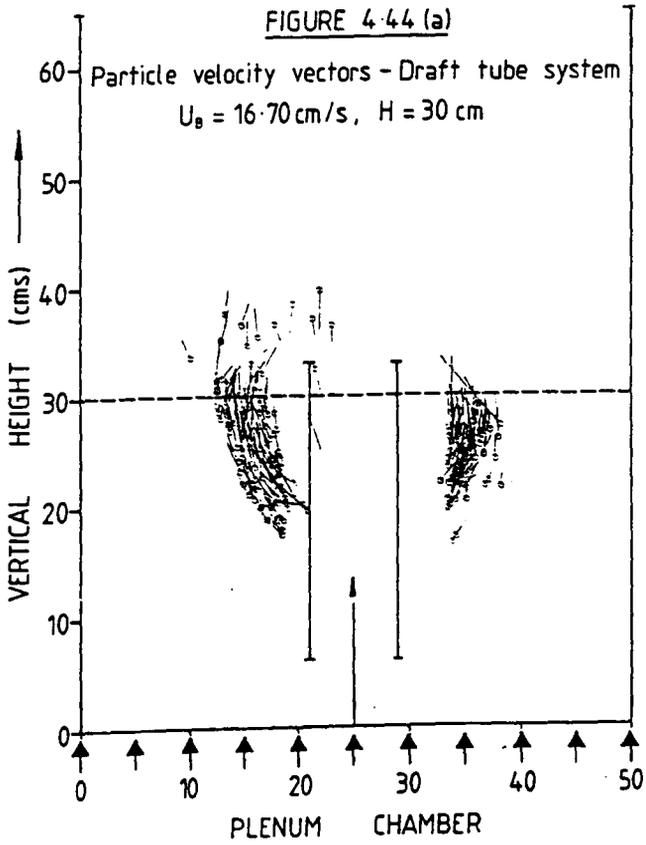
$U_B = 16·70 \, cm/s$, $H = 30 \, cm$

FIGURE 4·44 (c)

Particle velocity vectors in 3-D – Draft tube system

$U_B = 16·70 \, cm/s$, $H = 30 \, cm$

FIGURE 4·45(a)

Particle velocity vectors – Draft tube system

U_s = 20·83 cm/s, H = 30 cm

VERTICAL HEIGHT (cms)

PLENUM CHAMBER

FIGURE 4·45 (b)

Particle concentrations – Draft tube system

U_s = 20·83 cm/s, H = 30 cm

VERTICAL HEIGHT (cms)

PLENUM CHAMBER (cms)
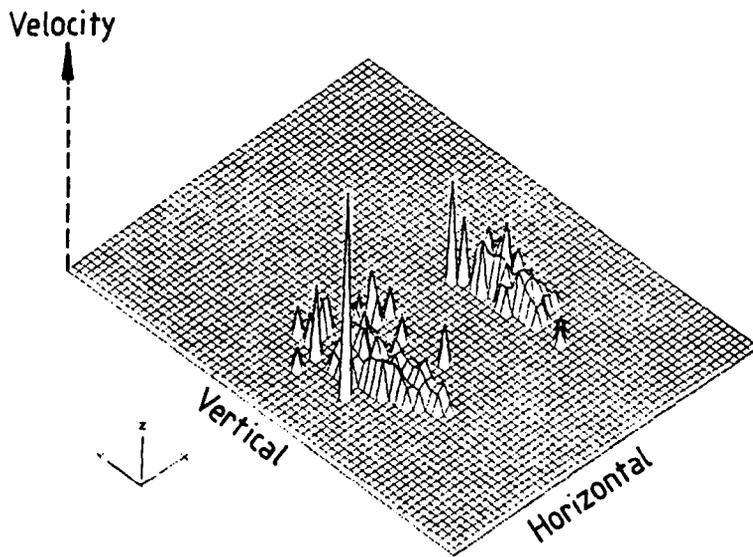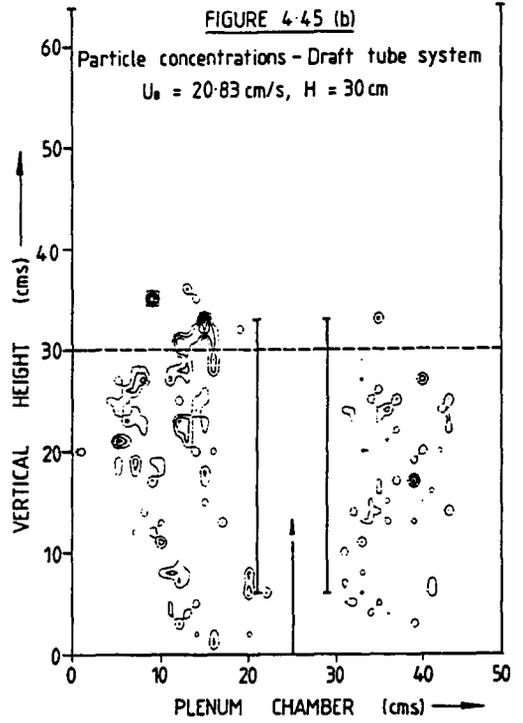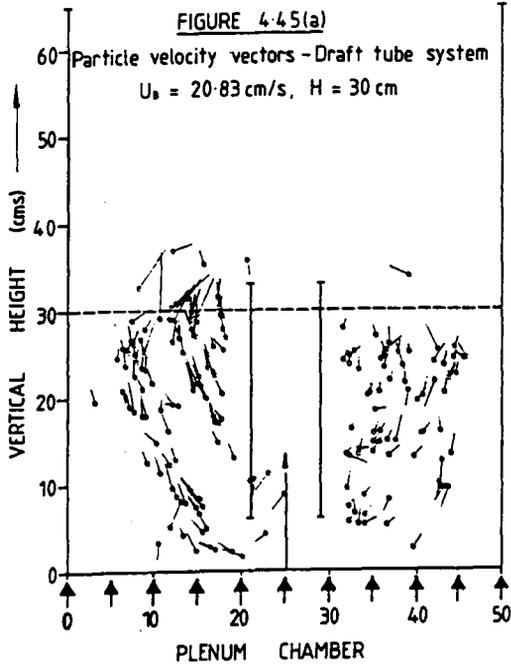
FIGURE 4·45 (c)

Particle velocity vectors in 3-D – Draft tube system

U_B = 20·83 cm/s, H = 30 cm

Velocity

Vertical

Horizontal

FIGURE 4·46 (a)
Particle velocity vectors – Draft tube system
U_B = 20·83 cm/s, H = 35 cm

PLENUM CHAMBER

FIGURE 4·46 (b)
Particle concentrations – Draft tube system
U_B = 20·83 cm/s, H = 35 cm

PLENUM CHAMBER (cms)
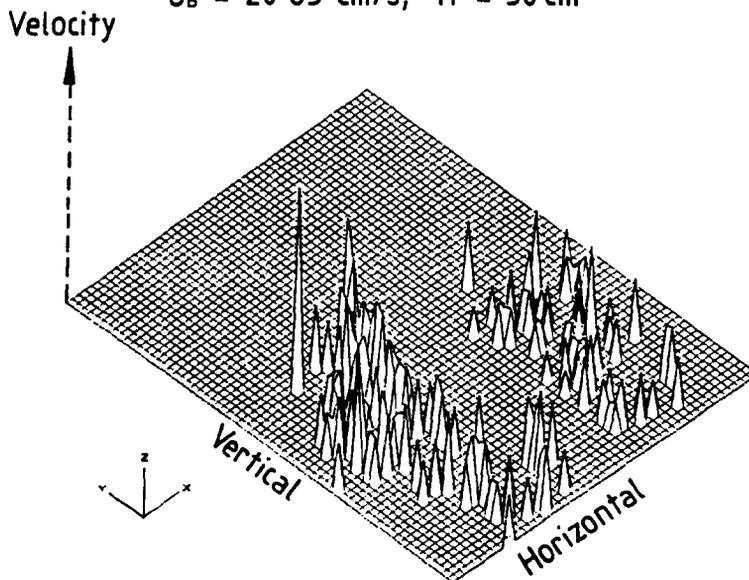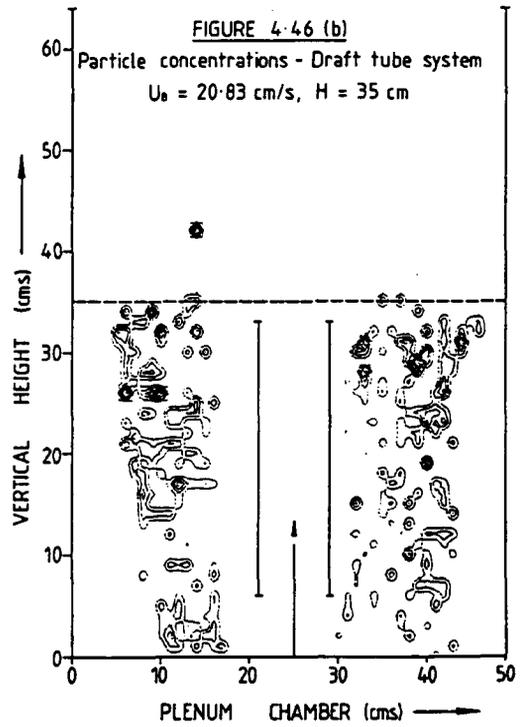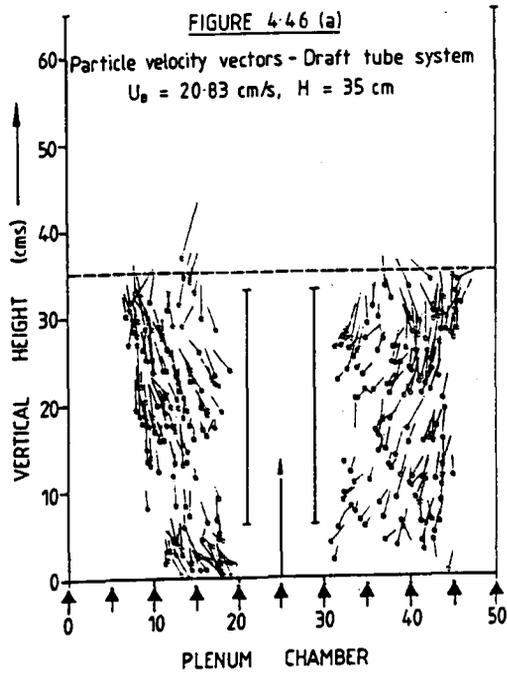
FIGURE 4·46 (c)
Particle velocity vectors in 3-D – Draft tube system
U_B = 20·83 cm/s, H = 35 cm

Velocity

Vertical

Horizontal

FIGURE 4·47 (a)
Particle velocity vectors – Draft tube system
U_0 = 20·83 cm/s, H = 40 cm

PLENUM CHAMBER

FIGURE 4·47 (b)
Particle concentrations – Draft tube system
U_0 = 20·83 cm/s, H = 40 cm

PLENUM CHAMBER (cms) ⟶
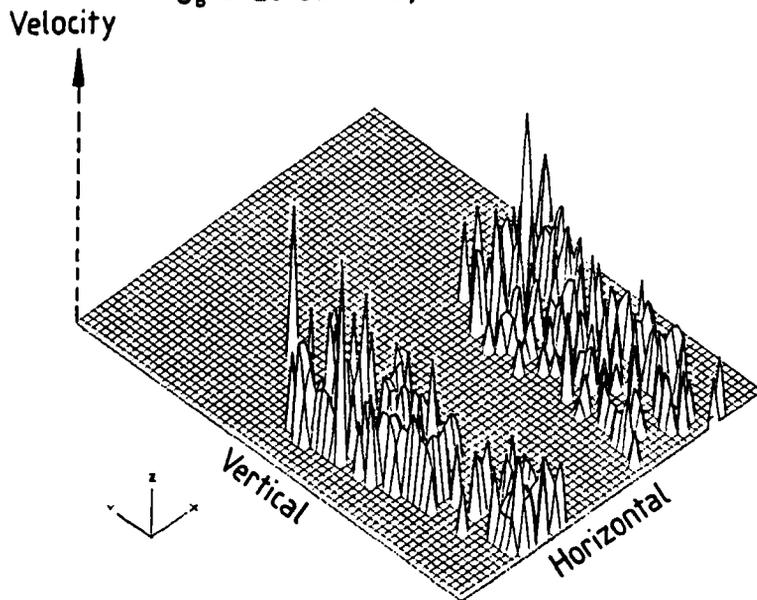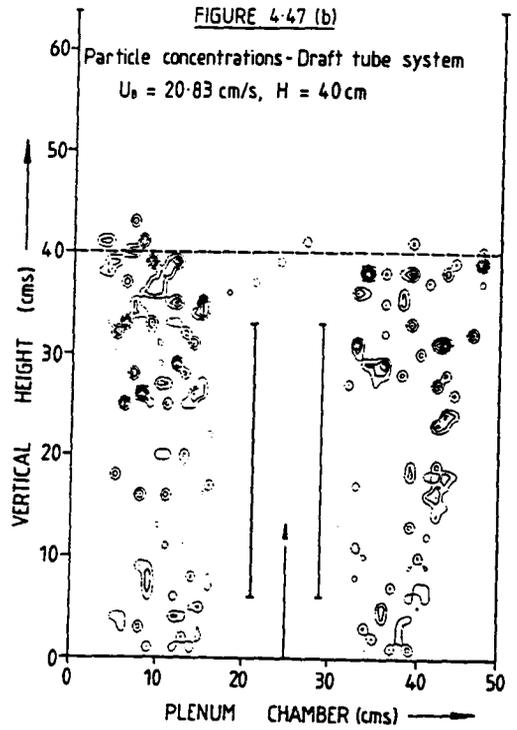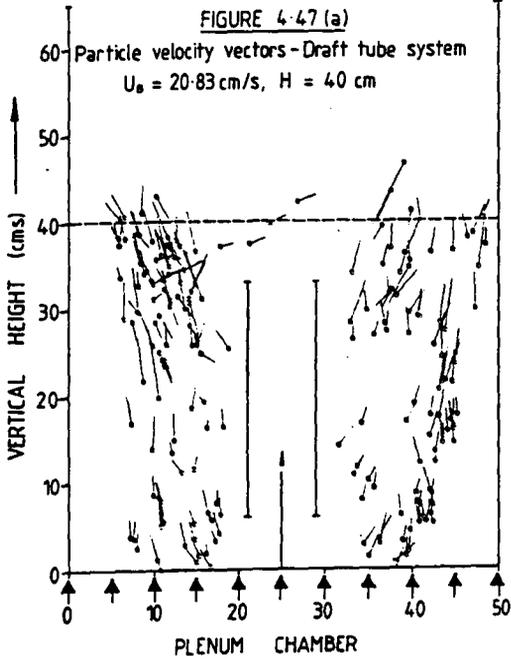
FIGURE 4·47 (c)

Particle velocity vectors in 3-D – Draft tube system
U = 20·83 cm/s, H = 40 cm

Velocity

Vertical

Horizontal

FIGURE 4·48 (a)

Particle velocity vectors - Draft tube system
$U_B = 27·50$ cm/s, $H = 30$ cm

PLENUM CHAMBER

FIGURE 4·48 (b)

Particle concentrations - Draft tube system
$U_B = 27·50$ cm/s, $H = 30$ cm

PLENUM CHAMBER (cms) ⟶
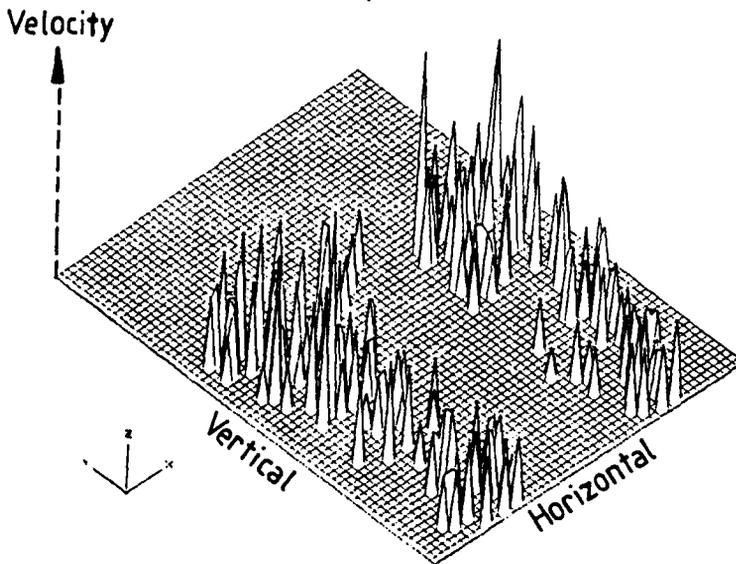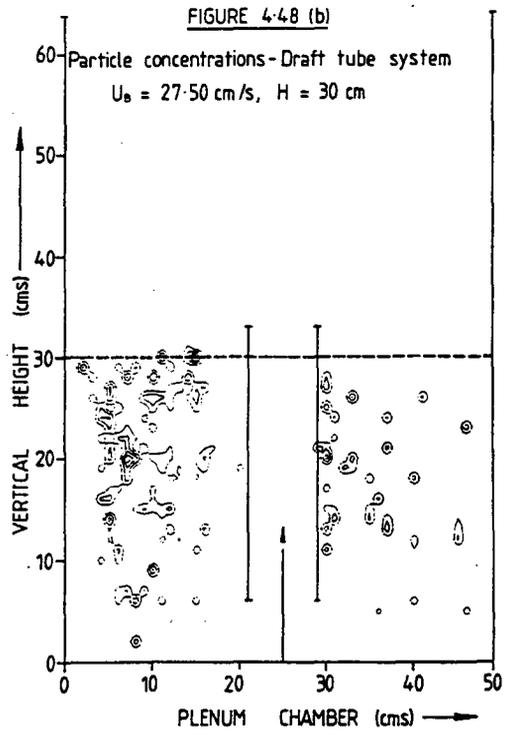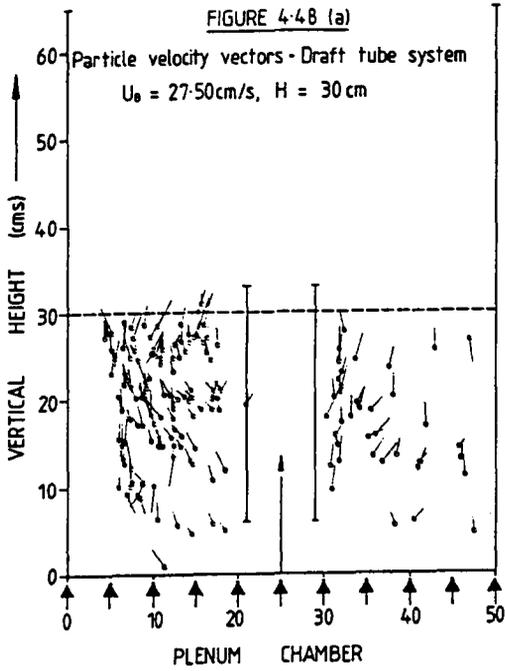
FIGURE 4·48 (c)

Particle velocity vectors in 3-D – Draft tube system
$U_B = 27·50$ cm/s, $H = 30$ cm

Velocity

Vertical

Horizontal
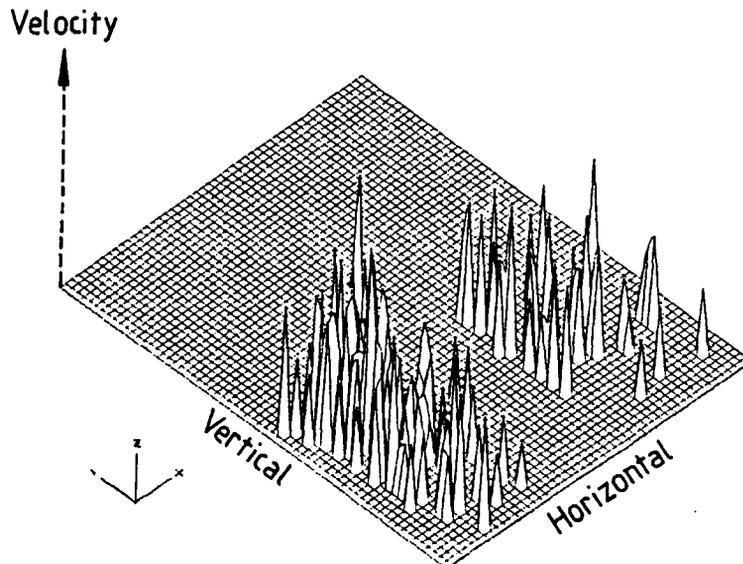
FIGURE 4·49

Effect of Fluidising Bed Velocity on Particle Percentage
in Four Directions in a Draft Tube System
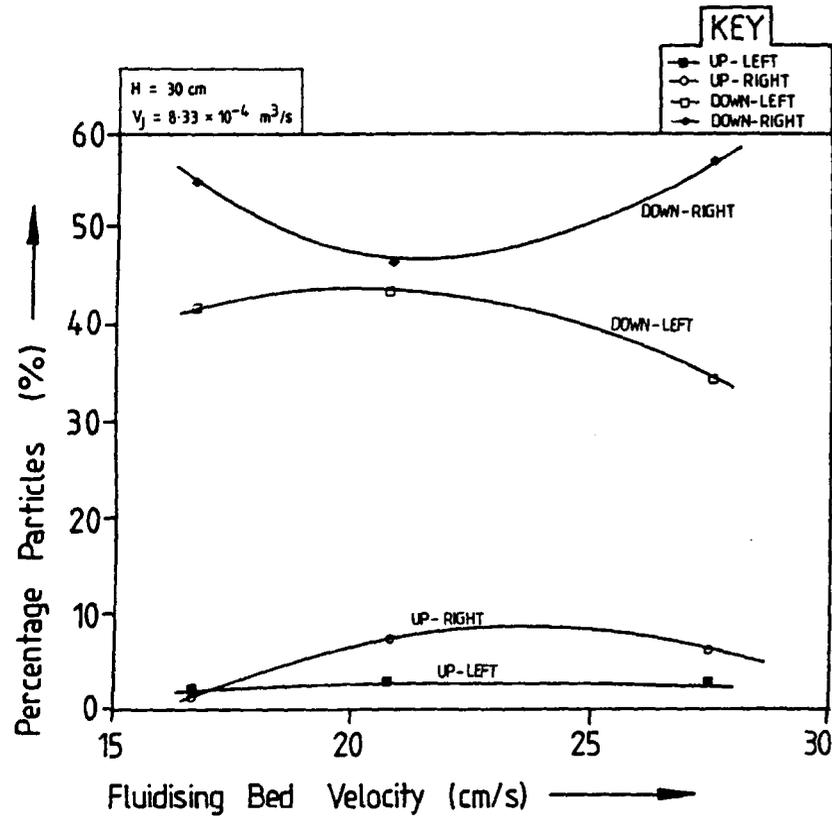
FIGURE 4·50

Effect of Bed Height on Particle Percentage in Four
Directions in a Draft Tube System
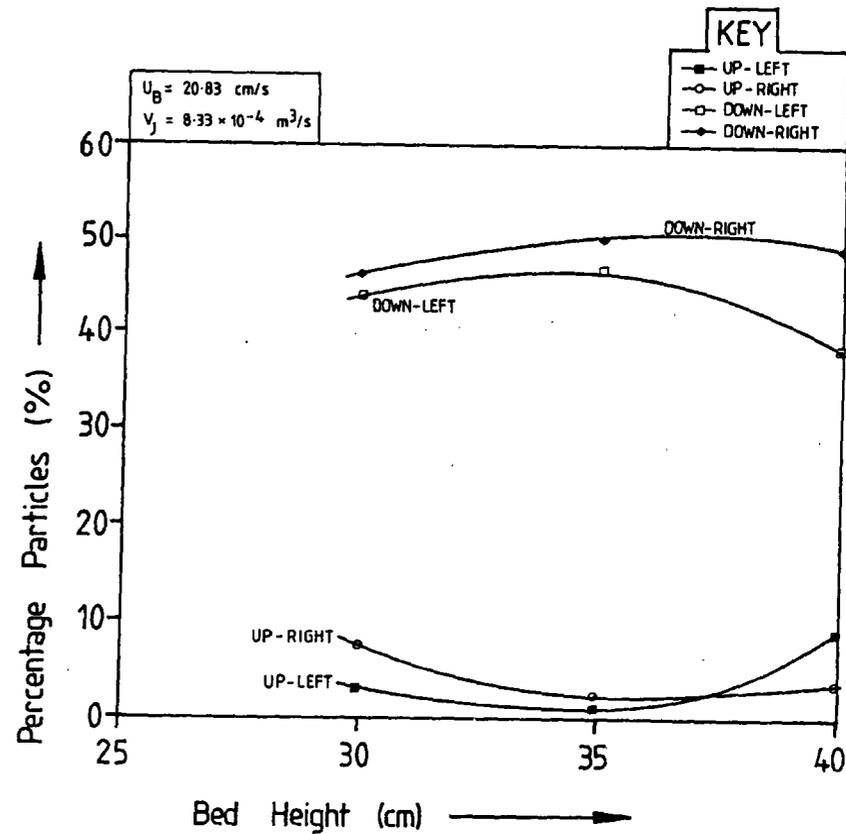
FIGURE 4·51

Effect of Fluidising Bed Velocity on Particle

Percentage in Four Directions
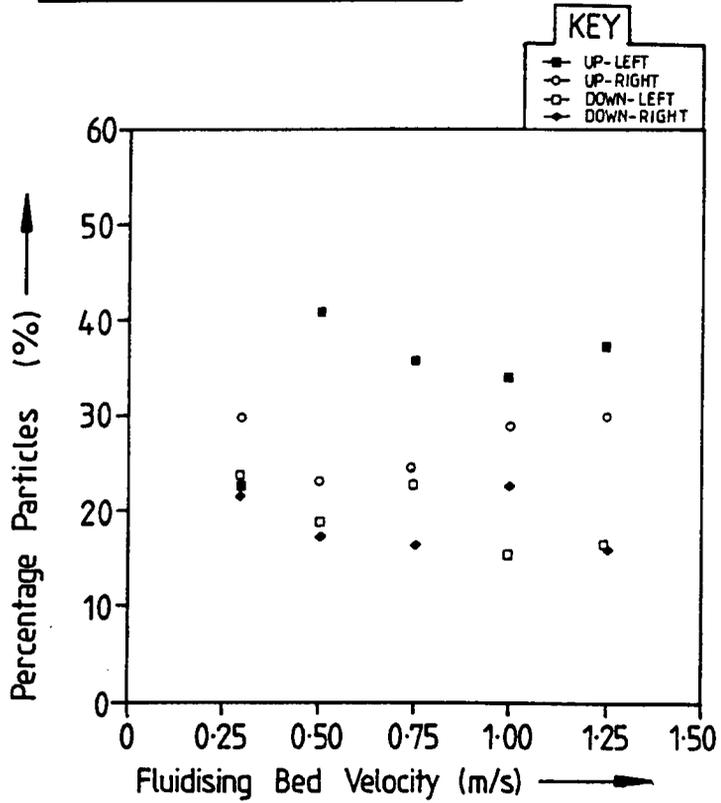
4.5.3.3 Jet system

The novel jet system incorporating two jets, one horizontal and one vertical, as shown in Figure 4.3 earlier in this chapter. THe study investigated the effect of fluidising bed velocity and total jet volumetric flowrate ($V_J = V_V + V_H$) on the particle velocities. The bed height was kept constant at 30cm. The results obtained from the experimental work are presented in Tble 4.16.

Figures 4.53, 4.54 and 4.55 (a, b & c) show the effect of the total jet volumetric flowrate (11.00, 13.33 and 16.17 x 10^{-4} m^3/s, respectively). The fluidising bed velocity is kept at 20.83 cm/s. There is a clear recirculation pattern obtained, but the effect of the jet volumetric flowrate is not obvious. The medium jet volumetric flowrate (13.33 x 10^{-4} m^3/s), Figure 4.54 (a, b & c) seems to yield the best recirulation. This is confirmed by comparing the velocity concentrations and the 3-D figures. The reason for the poor performance of the other jet volumetric flowrates is that at high V_J (16.17 x 10^{-4} m^3/s), the particles can be seen to be repelled from the mouth of the jets due to the jet penetration. Whilst at low V_J (11.00 x 10^{-4} m^3/s), there is insufficient flowrate to entrain the particles.

Considering Figures 4.52, 4.53 and 4.56 (a, b & c) which show the effect of the fluidising bed velocity

(16.70, 20.83 and 27.50 cm/s respectively). The total jet volumetric flowrate is kept constant at 11.00 x 10^{-4} m^3/s. It is clearly obvious from the figures that a recirculation pattern is obtained for the first two cases where U_B is low, but increasing U_B the recirculation patterns are distorted. The recirculation in the second case, Figure 4.53, is better than in the first case, Figure 4.52. Since a greater proportion of the bed is utilised, thus leaving negligible bed stagnant.

Figures 4.57 and 4.58 show the effect of fluidising bed velocity and jet volumetric flowrates on the percentage of particles in the four directions. The common trend in both cases is that either increasing the fluidising bed velocity or increasing the jet volumetric flowrate, the induced recirculation patterns and rates are destroyed. In other words the required recirculation is less obtainable at higher bed velocities or higher jet volumetric flowrates. In the case of jet volumetric flowrates, the best required percentage circulation of particles (Down-left) is obtained at between 13 to 14 x 10^{-4} m^3/s.

TABLE 4·16

Jets Particle Velocity Vectors

	$U_B = 16.70$ cm/s $V_H = 5.07 \times 10^{-4}$ m³/s $V_V = 5.93 \times 10^{-4}$ m³/s	$U_B = 20.83$ cm/s $V_H = 5.07 \times 10^{-4}$ m³/s $V_V = 5.93 \times 10^{-4}$ m³/s	$U_B = 20.83$ cm/s $V_H = 6.14 \times 10^{-4}$ m³/s $V_V = 7.19 \times 10^{-4}$ m³/s	$U_B = 20.83$ cm/s $V_H = 7.45 \times 10^{-4}$ m³/s $V_V = 8.72 \times 10^{-4}$ m³/s	$U_B = 27.50$ cm/s $V_H = 5.07 \times 10^{-4}$ m³/s $V_V = 5.93 \times 10^{-4}$ m³/s
Max.(av.) velocity	2·76 cm/s	2·56 cm/s	2·75 cm/s	3·39 cm/s	4·69 cm/s
Min. (av.) velocity	2·30 cm/s	2·17 cm/s	2·25 cm/s	2·49 cm/s	4·29 cm/s
Average velocity	2·53 cm/s	2·36 cm/s	2·50 cm/s	2·91 cm/s	4·47 cm/s
S.D. (av.)	± 0·21 cm/s	± 0·19 cm/s	± 0·24 cm/s	± 0·42 cm/s	± 0·22 cm/s
Up-right (av.) % particles	2·25 cm/s 9·0 %	2·33 cm/s 22·0 %	2·54 cm/s 9·0 %	2·97 cm/s 22·5 %	4·70 cm/s 24·2 %
Up-left (av.) % particles	2·32 cm/s 19·0 %	2·34 cm/s 21·0 %	2·54 cm/s 13·0 %	2·52 cm/s 13·8 %	4·49 cm/s 27·3 %
Down-left (av.) % particles	2·65 cm/s 65·0 %	2·37 cm/s 40·0 %	2·46 cm/s 60·0 %	3·29 cm/s 40·0 %	4·85 cm/s 23·5 %
Down-right (av.) % particles	2·35 cm/s 7·0 %	2·38 cm/s 17·0 %	2·57 cm/s 18·0 %	2·46 cm/s 23·7 %	3·86 cm/s 25·0 %

FIGURE 4·52 (a)

Particle velocity vectors – Jet system
$U_B = 16·70$ cm/s, $V_J = 66$ lit/min

VERTICAL HEIGHT (cms)

PLENUM CHAMBER

FIGURE 4·52 (b)

Particle concentrations – Jet system
$U_B = 16·70$ cm/s, $V_J = 66$ lit/min

VERTICAL HEIGHT (cms)

PLENUM CHAMBER (cms) →
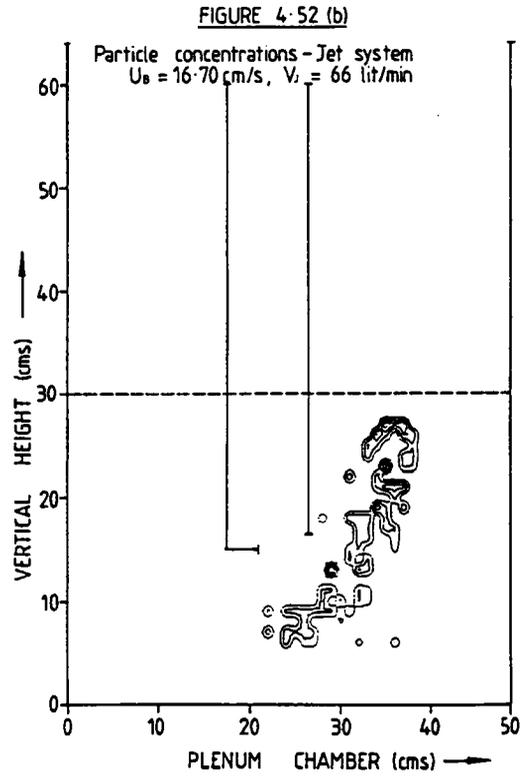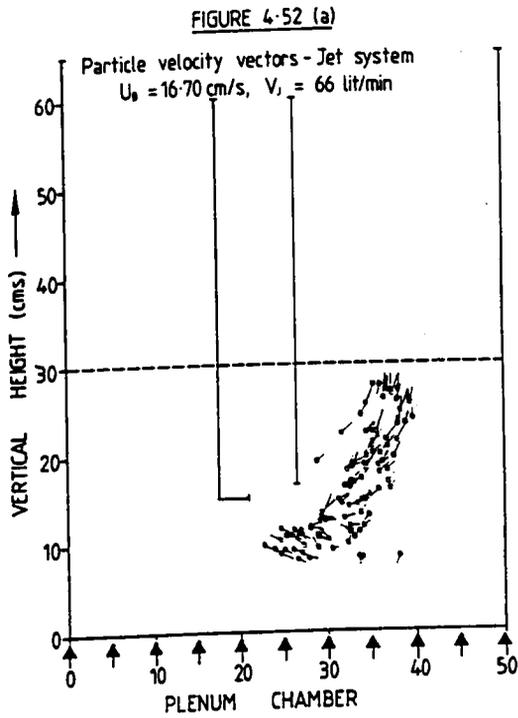
FIGURE 4·52 (c)

Particle velocity vectors in 3-D – Jet system
$U_B = 16·70$ cm/s, $V_J = 66$ lit/min

Velocity

Vertical

Horizontal

FIGURE 4·53 (a)

Particle velocity vectors – Jet system
$U_B = 20·83$ cm/s, $V_J = 66$ lit/min

PLENUM CHAMBER

FIGURE 4·53 (b)

Particle concentrations – Jet system
$U_B = 20·83$ cm/s, $V_J = 66$ lit/min

PLENUM CHAMBER (cms)
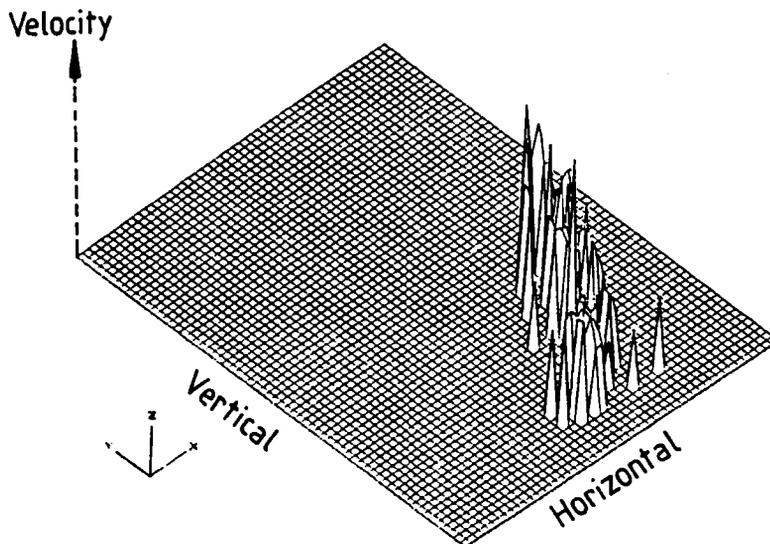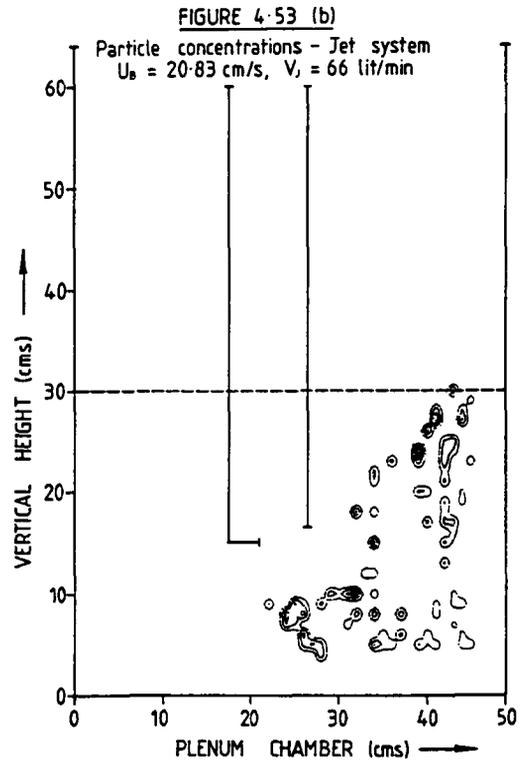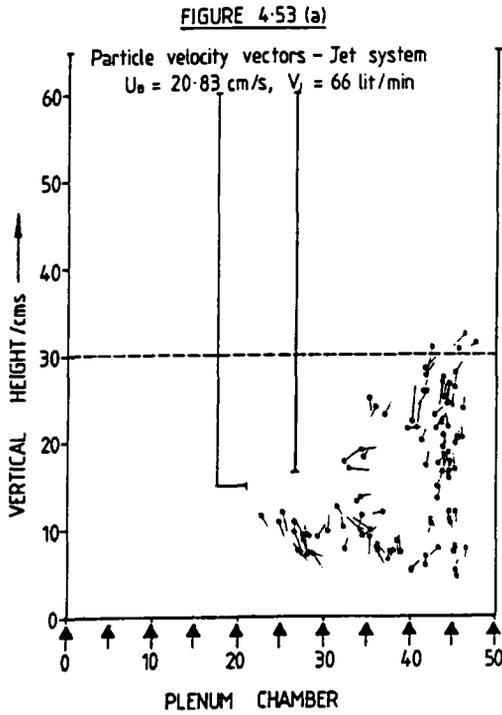
FIGURE 4·53 (c)

Particle velocity vectors in 3-D – Jet system
$U_B = 20·83$ cm/s, $V_J = 66$ lit/min

Velocity

Vertical

Horizontal

FIGURE 4·54 (a)

Particle velocity vectors – Jet system
U_B = 20·83 cm/s, V_J = 80 lit/min

FIGURE 4·54 (b)

Particle concentrations – Jet system
U_B = 20·83 cm/s, V_J = 80 lit/min
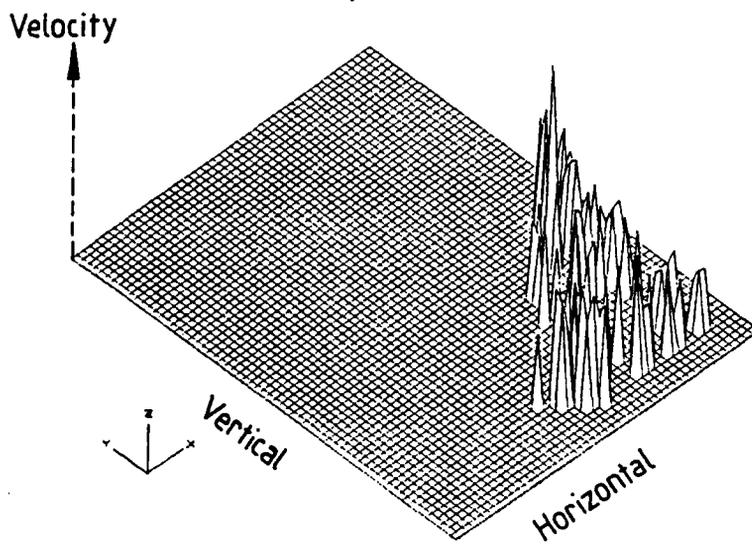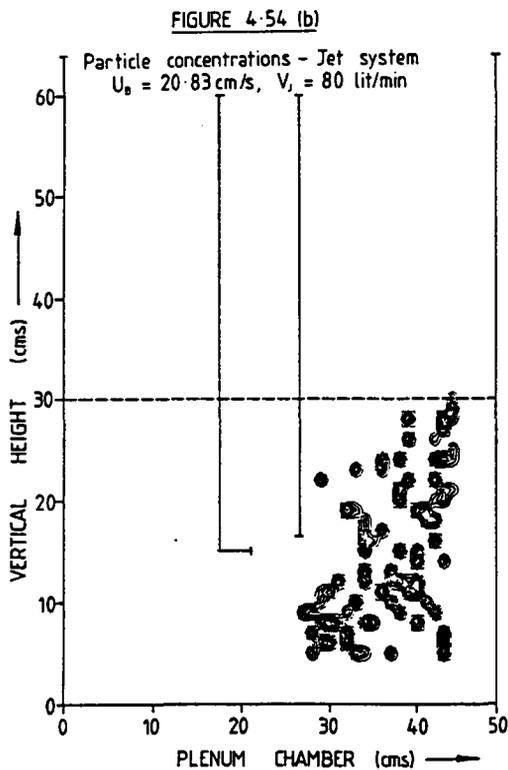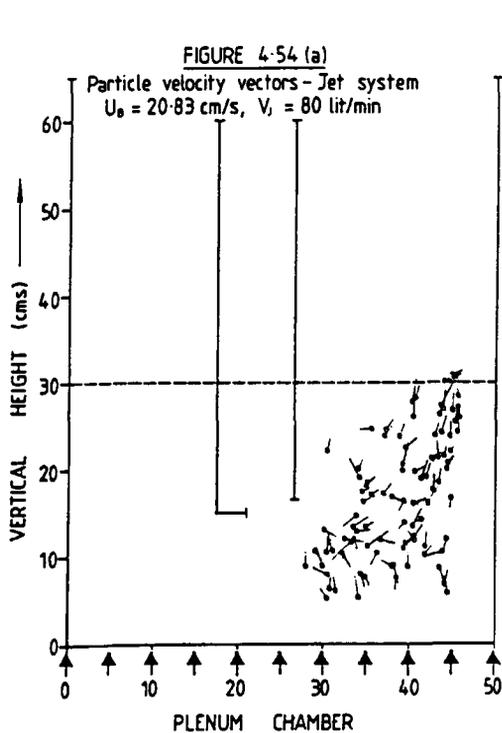
FIGURE 4·54 (c)

Particle velocity vectors in 3-D – Jet system
U_B = 20·83 cm/s, V_J = 80 lit/min

FIGURE 4·55 (a)

Particle velocity vectors - Jet system
U_B = 20·83 cm/s, V_J = 97 lit/min

VERTICAL HEIGHT (cms)

PLENUM CHAMBER

FIGURE 4·55 (b)

Particle concentrations - Jet system
U_B = 20·83 cm/s, V_J = 97 lit/min

VERTICAL HEIGHT (cms)

PLENUM CHAMBER (cms)
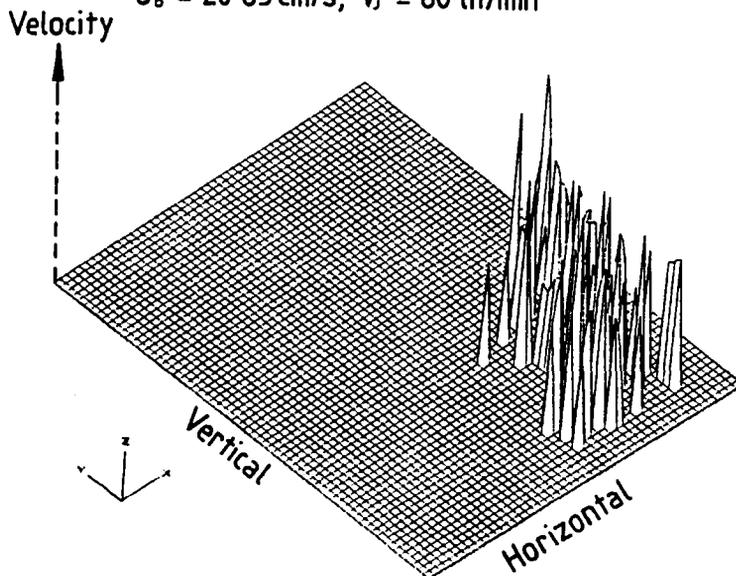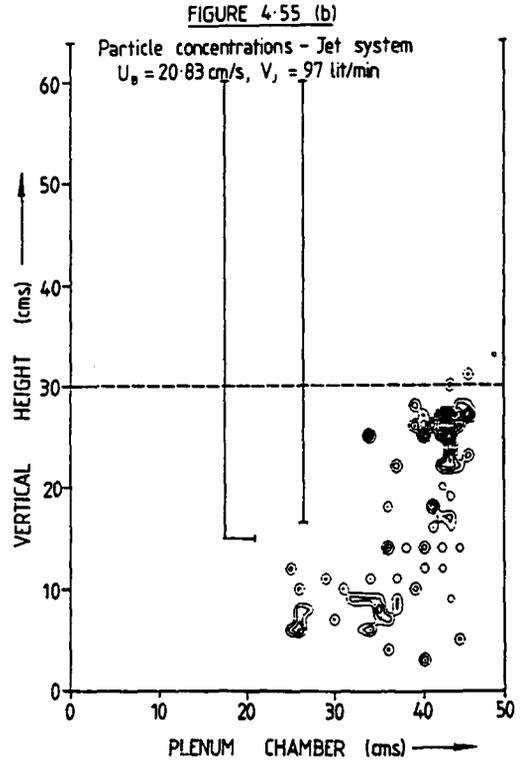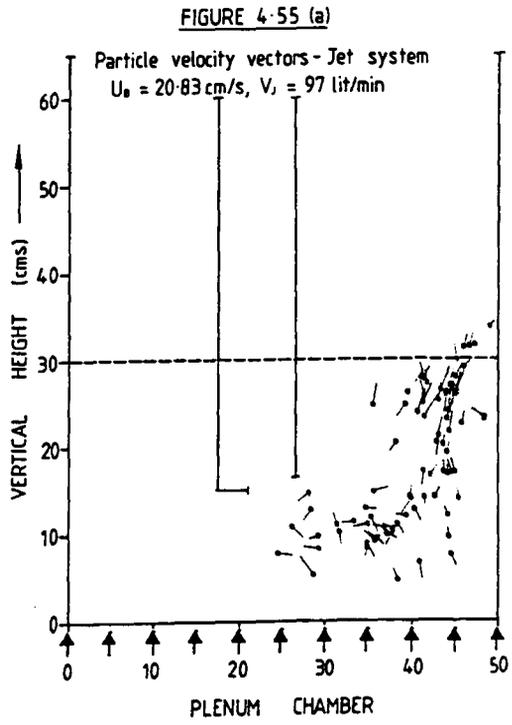
FIGURE 4·55 (c)

Particle velocity vectors in 3-D - Jet system
U_B = 20·83 cm/s, V_J = 97 lit/min

Velocity

Vertical

Horizontal

FIGURE 4·56 (a)

Particle velocity vectors - Jet system
$U_8 = 27·50$ cm/s, $V_J = 66$ lit/min

PLENUM CHAMBER

FIGURE 4·56 (b)

Particle concentrations - Jet system
$U_8 = 27·50$ cm/s, $V_J = 66$ lit/min

PLENUM CHAMBER (cms) ⟶
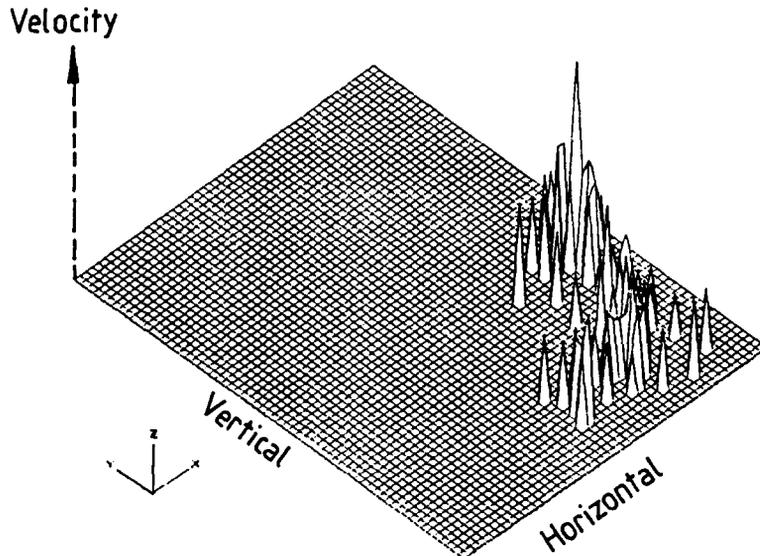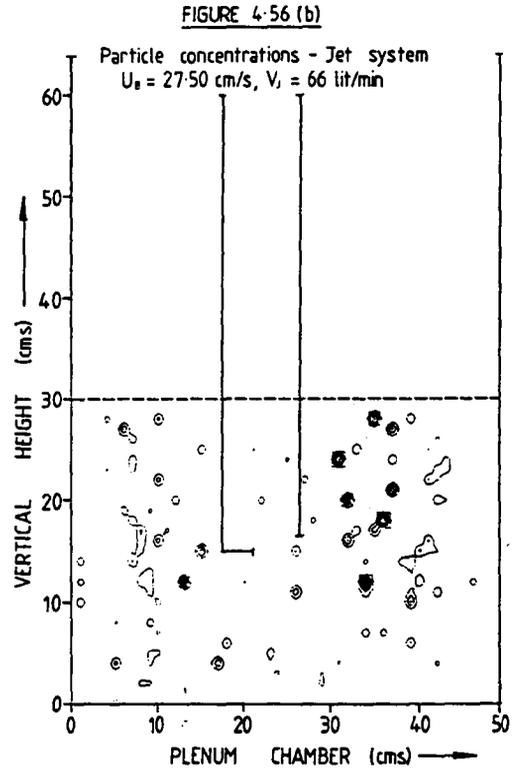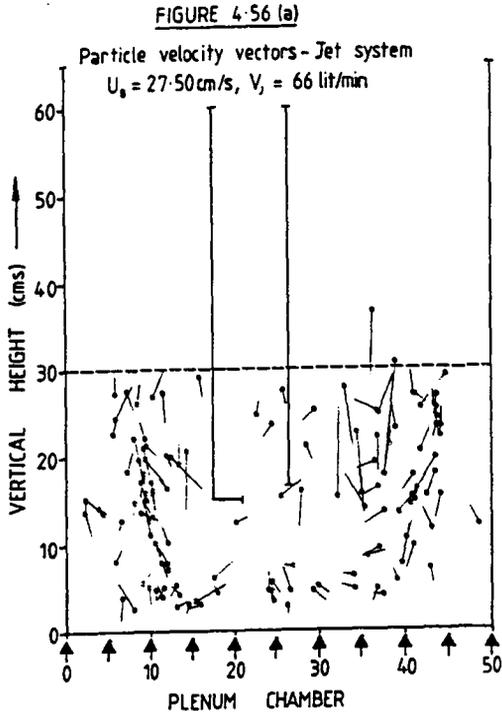
FIGURE 4·56 (c)

Particle velocity vectors in 3-D – Jet system
$U_8 = 27·50$ cm/s, $V_J = 66$ lit/min

Velocity

Vertical

Horizontal
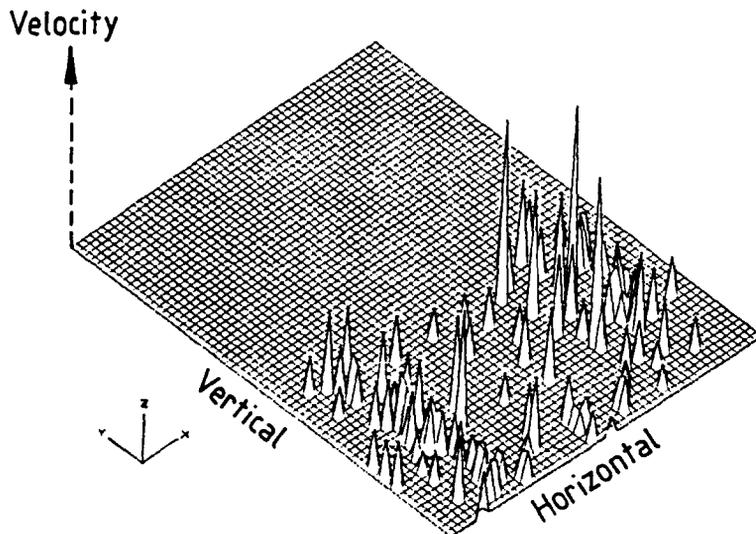
FIGURE 4·57

Effect of Fluidising Bed Velocity on Particle Percentage Four Directions in a Jet System
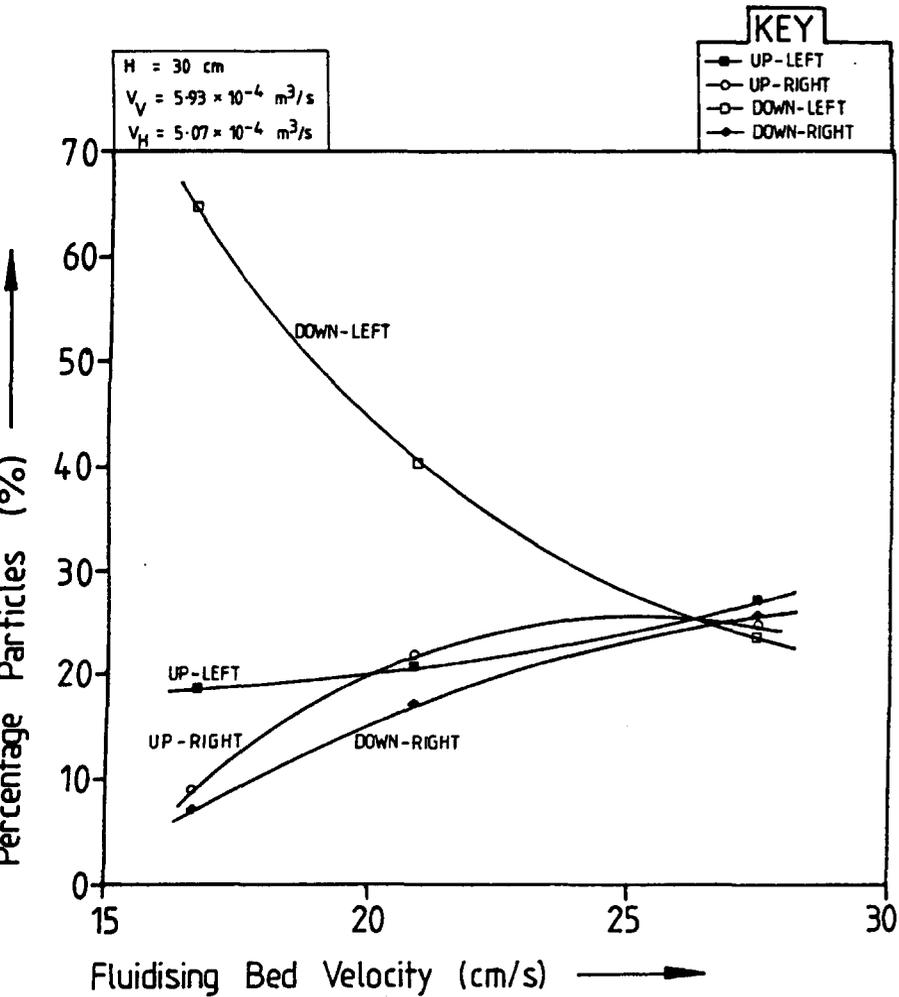
FIGURE 4·58

Effect of Jet Volumetric Flowrate (Total) on Particle Percentage in Four Directions in a Jet System

- 245 -
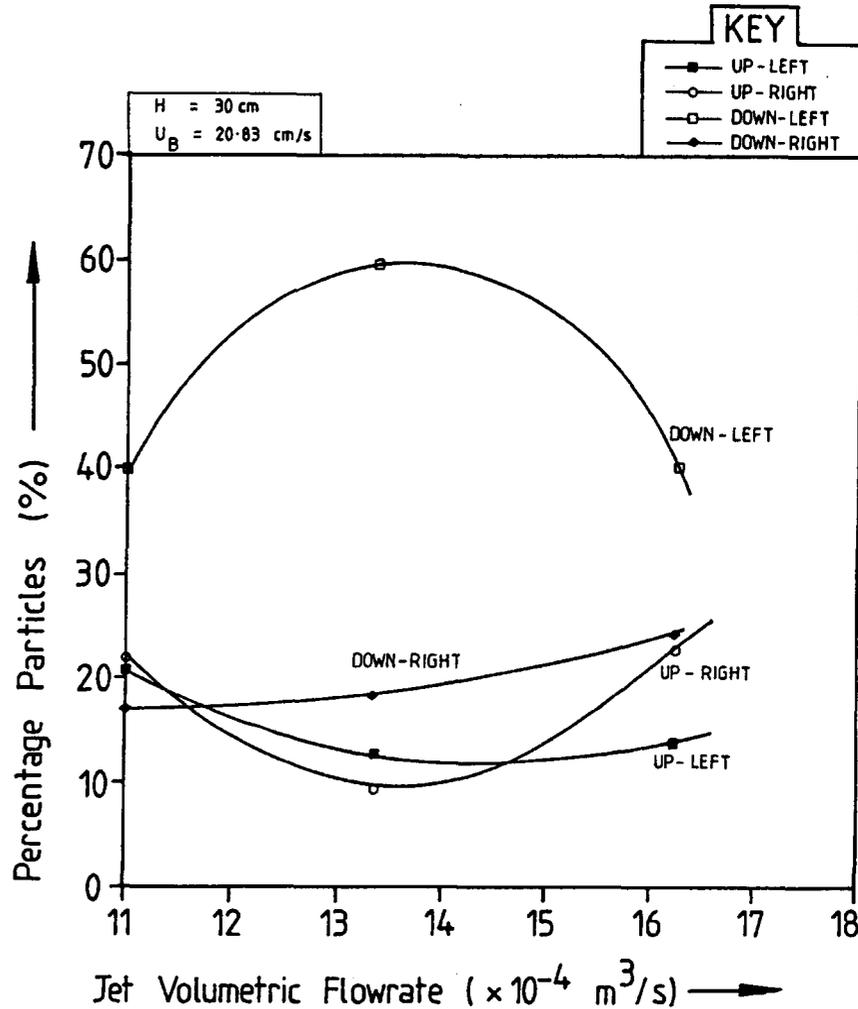
4.5.3.4 Conclusions

(A) FREE BED SYSTEM

Free bed systems are limited as far as induced recirculation requirements. It has been shown that the circulation is random and the particle velocities in all four directions increases with increasing bed velocity.

(B) DRAFT TUBE SYSTEM

In this system required recirculation patterns and rates can be obtained as shown. Increasing the bed height improves the particle circulation bandwidth by utilising a greater proportion of the bed geometry and leaving very little dead space. The same kind of trend, but of greater magnitude, was also seen by increasing the fluidising bed velocity. From the particle percentage observations, both the bed height and fluidising bed velocity have an optimum value where the required circulation (down-right and down-left) are maximum. For the bed height it is 35cm (same as the draft tube height) and for the fluidising bed velocity it is approximately 20cm/s (just under U_{mf}).

(C) JET SYSTEM

This type of system can also produce the required circulation patterns and rates. In the case of the influence of jet volumetric flowrates, it was found that best required circulation rates (down-left) were produced

at medium jet flowrates (13 to 14 x 10^{-4} m^3/s), for both the circulation patterns and particle percentages. The low and high jet flowrates seem to either have insufficient entrainment or repelling effect, respectively. In the case of fluidising bed velocity, it was found that low U_B (less than U_{mf}) produced the best circulation patterns while the lowest U_B (approximately 15cm/s) produced the best particle percentage in the required direction (down-left).

The significance of this study with respect to the gasification/pyrolysis of biomass is that it has been shown that induced recirculation rates can be obtained using either the draft tube or the jet system. From the conclusions of the experimental chapter it was found that separating the combustion reactions from the gasification and pyrolysis reactions would yield a better quality produced. In this chapter it has been shown that circulation rates can be induced and hence distinct zones can be created within the reactor bed using either of the draft tube or jet system, giving a better quality product gas. Furthermore since the rates of induced circulation can be controlled by a number of parameters it would be possible to use different feedstocks employing the same reactor bed system.

CHAPTER FIVE

SUMMARY AND CONCLUSIONS

5 SUMMARY AND CONCLUSIONS

It is the aim of this chapter to summarise the findings of the studies undertaken, gasification/pyrolysis of woochips and the cold modelling. Full discussion of results is included in the relevant chapters, therefore only brief conclusions will be presented here along with suggestions for further work.

5.1 THE GASIFICATION/PYROLYSIS OF WOODCHIPS

Woodchips and other forms of wood wastes have been shown to have the potential to be an important supplement to conventional fuels if efficiently utilised. Using air as the gasifying medium, woodchips were converted into a low CV fuel gas of about 6 MJ/Nm^3, at fuel feedrates in excess of about 3.5 x S, and at bed temperatures in the range of 700 - 800 $^\circ$C.

5.1.1 Fuel Feedrate

The results of the experimental work showed a marked improvement in the quality of the product gas as fuel feedrate was increased. Although there was a limiting factor, the gasification efficiency, which decreased with

very high fuel feedrates (> 4 x S). The drop in efficiency
at higher fuel feedrates is due to the 'diminishing
returns' effect, whereby an increase in the fuel feedrate
is not matched by a corresponding rise in the calorific
value of the fuel gas produced. In other words there is a
limit to how much the fuel gas quality can be improved with
respect to fuel feedrate.

5.1.2 Temperature

The quality of the product gas showed marked
improvement with increase in temperature. The only
limiting factors are the materials of construction, the ash
fusion temperature and the energy loss from hot product
gases. The latter can be minimised by utilising the
sensible heat of the product gases either by using heta
exchangers to remove the heat energy, or by using the
product gas immediately after production in an adjacent
plant.

5.1.3 Bed Height

Increasing the static bed height produced a better
quality fuel gas. The reason for this was that there was a
better separation of the combustion and
gasification/pyrolysis zones. This could further be
improved by having an induced recirculation within the

system. For this purpose a cold modelling study was conducted to enable the determination of the induced recirculations within two different systems, namely the draft tube and the jet system.

5.2 COLD MODELLING

5.2.1 Free Bed System

The empty fluidised bed without any bed internals demonstrated random recirculation, with little or none induced circulation patterns.

5.2.2 Draft Tube And Jet Systems

In these systems the required recirculation patterns and rates can be obtained. They could be employed to separate the combustion and gasification/pyrolysis zones with respect to woodchip conversion processes. This would yield a better quality product gas. Additionally since the rates of induced recirculation can be controlled by a number of parameters, as shown in the cold modelling chapter, it would be possible to use different feedstocks employing the same reactor bed configuration.

5.3 FURTHER WORK

The following suggestions for further work in this field are :

(1) The yield and characterisation of the oils and tars formed during gasification/pyrolysis with view to assessing by-product recovery.

(2) Experimental tests employing either or both the draft tube system and the jet system, using air as the gasifying medium, and woodchips as the feedstock to obtain comparative results.

(3) The processing of woodchips with steam, steam/air or steam/oxygen mixtures for the production of synthesis gas, rich in carbon-monoxide and hydrogen, especially using the the draft tube or jet system.

CHAPTER SIX

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6 BIBLIOGRAPHY AND PAPERS PUBLISHED BY THE AUTHOR

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APPENDICES

APPENDIX 1

SIGNIFICANCE OF REACTOR BED GAS PROFILES

The Tables, 1 to 9, given here represents the
information collected regarding the concentration of
gaseous species at several distances above the distributor
plate. The results obtained are in the form of a
concentration profile through the bed.

TABLE 1

BED PROFILE (RUN 1)

Bed depth = 10 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0320 Kg/min (2·0 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0·73	2·23	6·21	8·95
H_2	0	0·18	0·49	1·29	1·86
CH_4	0	0·05	0·29	0·87	1·09
C_2H_2	0	0	0	0	0
C_2H_4 ⟩ C_2H_6	0	0	0·16	0·44	0·57
C_3H_6	0	0	0·02	0·08	0·11
C_3H_8	0	0	0	0·01	0·01
O_2	21·00	9·46	5·19	1·22	0·66
N_2	79·00	87·81	82·02	75·20	70·52
CO_2	0	1·77	9·60	14·78	16·23
Temp. K	298	1037	1040	771	503

TABLE 2

BED PROFILE (RUN 2)

Bed depth = 10 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0499 Kg/min (3·1 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0	2·34	7·52	14·13
H_2	0	0	0·53	1·73	3·41
CH_4	0	0	0·31	1·91	3·08
C_2H_2	0	0	0·01	0·09	0·26
C_2H_4 C_2H_6	0	0	0·20	0·95	1·66
C_3H_6	0	0	0·03	0·09	0·24
C_3H_8	0	0	0	0	0·01
O_2	21·00	9·23	2·15	1·19	0·45
N_2	79·00	87·80	81·74	71·44	59·21
CO_2	0	2·97	12·69	15·09	17·52
Temp. K	298	1059	1060	700	598

TABLE 3

BED PROFILE (RUN 3)

Bed depth = 10 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0499 Kg/min (3·1 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0·60	5·77	9·17	13·45
H_2	0	0·13	1·02	1·21	3·18
CH_4	0	0·07	1·16	1·96	2·79
C_2H_2	0	0	0·03	0·11	0·30
C_2H_4 C_2H_6	0	0	0·75	0·91	1·35
C_3H_6	0	0	0·09	0·15	0·23
C_3H_8	0	0	0	0	0·01
O_2	21·00	10·95	3·85	1·32	0·91
N_2	79·00	86·71	75·11	70·58	60·74
CO_2	0	1·54	12·22	14·69	17·04
Temp. K	298	1064	1019	725	524

TABLE 4

BED PROFILE (RUN 4)

Bed depth = 10 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0656 Kg/min (4·1 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0	5·34	12·31	14·37
H_2	0	0·01	1·10	2·01	3·20
CH_4	0	0	1·14	2·33	3·58
C_2H_2	0	0	0·11	0·15	0·17
C_2H_4 C_2H_6	0	0	1·01	1·09	1·14
C_3H_6	0	0	0·06	0·11	0·21
C_3H_8	0	0	0	0·01	0·02
O_2	21·00	12·56	1·91	1·45	1·12
N_2	79·00	87·05	74·10	64·79	60·10
CO_2	0	0·38	15·23	15·75	16·09
Temp. K	298	1051	1054	751	566

TABLE 5

BED PROFILE (RUN 5)

Bed depth = 10 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0368 Kg/min (2·3 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0·13	4·51	5·23	6·15
H_2	0	0·05	0·97	1·07	1·27
CH_4	0	0	0·72	0·89	1·12
C_2H_2	0	0	0	0	0
C_2H_4 / C_2H_6	0	0	0·44	0·56	0·76
C_3H_6	0	0	0·10	0·13	0·15
C_3H_8	0	0	0·01	0·01	0·01
O_2	21·00	12·73	6·46	5·36	4·50
N_2	79·00	86·58	79·00	78·34	75·91
CO_2	0	0·51	7·79	8·41	9·37
Temp. K	298	955	880	678	512

TABLE 6

BED   PROFILE (RUN 6)

Bed   depth = 10 cm

Air   feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel   feedrate = 0·0624 Kg/min (3·9 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0	8·71	10·57	12·42
H_2	0	0·01	1·43	1·75	2·20
CH_4	0	0	1·42	1·51	1·66
$C_2 H_2$	0	0	0·09	0·10	0·12
$C_2 H_4$ $C_2 H_6$	0	0	0·69	0·71	0·83
$C_3 H_6$	0	0	0·17	0·18	0·18
$C_3 H_8$	0	0	0·03	0·04	0·06
O_2	21·00	12·55	5·99	3·71	2·61
N_2	79·00	87·40	73·01	70·69	67·11
CO_2	0	0·04	8·47	10·74	12·81
Temp. K	298	942	947	703	544

TABLE 7

BED PROFILE (RUN 7)

Bed depth = 10 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0416 Kg/min (2·6 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0·25	2·02	6·43	9·83
H_2	0	0·09	0·46	1·01	1·94
CH_4	0	0	0·28	0·95	1·76
C_2H_2	0	0	0	0	0
C_2H_4 C_2H_6	0	0	0·19	0·87	1·16
C_3H_6	0	0	0·03	0·15	0·25
C_3H_8	0	0	0	0·01	0·02
O_2	21·00	12·27	5·70	3·91	3·00
N_2	79·00	86·46	82·38	76·84	71·12
CO_2	0	0·93	8·94	9·83	10·92
Temp. K	298	959	935	751	585

TABLE 8

BED PROFILE (RUN 8)

Bed depth = 15 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0376 Kg/min (2·4 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0	0·03	1·09	3·87
H_2	0	0·01	0·06	0·21	0·51
CH_4	0	0	0	0·12	0·44
C_2H_2	0	0	0	0·01	0·03
C_2H_4 C_2H_6 >	0	0	0	0·09	0·20
C_3H_6	0	0	0	0·03	0·07
C_3H_8	0	0	0	0·01	0·03
O_2	21·00	18·99	18·33	7·94	1·54
N_2	79·00	80·20	80·16	79·29	75·42
CO_2	0	0·80	1·42	11·21	17·82
Temp. K	298	738	726	623	462

TABLE 9

BED PROFILE (RUN 9)

Bed depth = 5 cm

Air feedrate (15°C) = 80 lit/min (0·0976 Kg/min)

Fuel feedrate = 0·0564 Kg/min (3·6 × S)

	DISTANCE ABOVE DISTRIBUTOR PLATE				
	0 cm	2 cm	8 cm	40 cm	130 cm
CO	0	0	0	14·45	17·06
H_2	0	0	0·08	1·91	2·52
CH_4	0	0	0·02	1·27	1·79
C_2H_2	0	0	0	0·28	0·30
C_2H_4 C_2H_6	0	0	0·01	0·54	0·65
C_3H_6	0	0	0	0·06	0·08
C_3H_8	0	0	0	0·01	0·01
O_2	21·00	19·89	17·26	2·47	2·02
N_2	79·00	80·19	77·15	69·26	65·45
CO_2	0	0·02	5·48	9·75	10·12
Temp. K	298	875	825	583	404

APPENDIX 2

THE EFFECT OF VARYING PROCESS CONDITIONS

ON PRODUCT GAS QUALITY AND YIELD

The experimental results derived to investigate the effect of varying process conditions on product gas quality and yield are presented here, Tables 1 to 15 (a-e). Tables b and c represent the yield (Nm^3/Kg) and output (Nm^3/min), respectively, of the major gaseous components. While Table e gives the various process efficiencies. Tables a and d are the resulting derived data which have been employed to investigate the effect of varying process conditions.

TABLE 1

(a)

Bed depth = 10 cm
Temperature = 941–970 K
Air feedrate(15°C) = 80 lit/min (0·0976 Kg/min)

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
CO	6·91	9·18	8·32	7·55	9·84	11·00	9·83	14·20	12·32
H_2	1·27	1·62	0·96	1·44	1·90	2·28	1·94	2·16	2·27
CH_4	1·12	1·20	1·53	1·29	1·32	1·45	1·76	1·82	1·28
C_2H_2	0	0·12	0	0	0·14	0	0	0·10	0·05
C_2H_4 / C_2H_6	0·76	0·71	0·55	0·86	0·79	0·85	1·16	0·86	0·60
C_3H_6	0·15	0·12	0·08	0·18	0·14	0·19	0·25	0·22	0·17
C_3H_8	0·01	0·05	0·01	0·01	0·11	0·01	0·02	0·02	0·02
O_2	4·50	2·15	5·14	5·26	2·28	0·38	3·00	2·98	3·04
N_2	75·91	75·15	69·79	73·64	73·46	66·13	71·12	65·41	65·63
CO_2	9·37	9·70	13·62	9·77	10·02	17·71	10·92	12·23	14·62
CV MJ/Nm³(dry)									
Gross	2·15	2·56	2·24	2·41	2·88	3·02	3·23	3·67	2·98
Net	2·03	2·43	2·13	2·28	2·73	2·87	3·05	3·49	2·84
Fuel feedrate Kg/min	0·0368	0·0277	0·0322	0·0368	0·0288	0·0336	0·0368	0·0624	0·0512
	2·3×S	1·7×S	2·0×S	2·3×S	1·8×S	2·1×S	2·3×S	3·9×S	3·2×S
Temp. K	957	957	950	955	967	966	959	942	954
Run duration mins	90	90	75	90	90	115	90	80	60

(c)

Bed depth = 10 cm, Temperature range = 941–970 K

$Nm^3/min \times 10$	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
CO	5·48	7·33	7·16	6·17	8·04	10·00	8·32	13·14	11·33
H_2	1·01	1·29	0·83	1·18	1·55	2·07	1·64	2·00	2·09
CH_4	0·89	0·96	1·32	1·05	1·08	1·32	1·49	1·68	1·18
C_2H_2	0·00	0·10	0·00	0·00	0·11	0·00	0·00	0·09	0·05
C_2H_4 / C_2H_6	0·60	0·57	0·47	0·70	0·65	0·77	0·98	0·80	0·55
C_3H_6	0·12	0·10	0·07	0·15	0·11	0·17	0·21	0·20	0·16
C_3H_8	0·01	0·04	0·01	0·01	0·09	0·01	0·02	0·02	0·02
O_2	3·57	1·72	4·43	4·30	1·86	0·35	2·54	2·76	2·80
N_2	60·16	60·04	60·10	60·16	60·05	60·12	60·16	60·51	60·36
CO_2	7·43	7·75	11·73	7·98	8·19	16·10	9·24	11·31	13·44
Total	79·25	79·89	86·11	81·69	81·75	90·91	84·59	92·51	91·96

(b)

Bed depth = 10 cm, Temperature range = 941–970 K

$Nm^3/Kg \times 10$	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
CO	11·89	26·46	22·24	16·77	27·92	29·76	22·61	21·06	22·13
H_2	2·74	4·66	2·58	3·21	5·38	6·16	4·46	3·20	4·08
CH_4	2·42	3·46	4·10	2·85	3·75	3·93	4·05	2·69	2·30
C_2H_2	0·00	0·36	0·00	0·00	0·38	0·00	0·00	0·14	0·10
C_2H_4 / C_2H_6	1·63	2·06	1·46	1·90	2·26	2·29	2·66	1·28	1·07
C_3H_6	0·33	0·36	0·22	0·41	0·38	0·50	0·57	0·32	0·31
C_3H_8	0·03	0·14	0·03	0·03	0·31	0·03	0·05	0·03	0·04
O_2	9·70	6·21	13·76	11·68	6·46	1·04	6·90	4·42	5·47
N_2	163·48	216·75	186·64	163·48	208·51	178·93	163·48	96·97	117·89
CO_2	20·19	27·98	36·43	21·68	28·44	47·92	25·11	18·12	26·25
Total	215·35	288·41	267·42	221·98	283·85	270·56	229·86	148·25	179·61

(d)

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
Prod. gas flow									
Dry (Nm³/min)	0·0792	0·0799	0·0861	0·0817	0·0818	0·0909	0·0846	0·0925	0·0920
Wet (Nm³/min)	0·0849	0·0837	0·0905	0·0867	0·0857	0·0955	0·0896	0·1010	0·0990
Per Kg fuel									
Dry (Nm³/Kg)	2·1522	2·8844	2·6739	2·2201	2·8403	2·7054	2·2989	1·6186	1·7961
Wet (Nm³/Kg)	2·3071	3·0216	2·8105	2·3560	2·9757	2·8423	2·4348	1·7673	1·9330
Steam content of wet gas (%)	6·0	4·5	4·9	5·8	4·6	4·8	5·6	8·5	7·1
Carry-over (g/min)	0·089	0·075	0·091	0·109	0·099	0·067	0·119	0·019	0·169
Bed-char + ash (g/min)	0·303	0·111	0·129	0·147	0·115	0·134	0·147	0·250	0·205

(e)

Bed depth = 10 cm, Temperature range = 941 – 970 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	46·7	18·9	56·3	100·0	22·9	35·7
RUN 2	71·7	27·8	59·0	100·0	36·7	53·8
RUN 3	74·3	22·6	84·7	100·0	29·8	45·6
RUN 4	51·9	22·3	63·0	100·0	26·5	39·8
RUN 5	75·5	31·7	62·5	100·0	40·5	57·6
RUN 6	98·2	31·9	90·7	100·0	40·6	57·8
RUN 7	65·9	31·4	65·4	100·0	36·8	51·2
RUN 8	51·2	19·4	68·1	100·0	27·1	36·5
RUN 9	60·4	18·6	79·1	100·0	26·7	37·9

TABLE 2

(a)

Bed depth = 10 cm
Temperature = 971— 990 K
Air feedrate(15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8
CO	12.42	9.68	12.66	12.53	9.81	12.24	13.50	11.81
H_2	2.20	1.98	2.20	1.70	1.92	2.41	1.82	2.88
CH_4	1.66	1.27	1.68	1.65	1.70	1.58	1.63	1.70
C_2H_2	0.12	0.06	0.12	0.19	0.00	0.18	0.06	0
C_2H_4 / C_2H_6	0.82	0.70	0.83	0.77	1.43	0.96	0.89	0.50
C_3H_6	0.18	0.18	0.19	0.19	0.10	0.18	0.25	0
C_3H_8	0.05	0.04	0.06	0.02	0.01	0.02	0.06	0
O_2	2.61	5.33	2.66	1.58	0.87	1.73	4.04	1.12
N_2	67.13	74.04	66.76	68.23	68.12	69.16	68.60	68.59
CO_2	12.81	6.72	12.84	13.14	16.04	11.54	9.15	13.40
CV MJ/Nm³(dry)								
Gross	3.36	2.70	3.43	3.29	3.23	3.43	3.52	2.88
Net	3.20	2.56	3.26	3.14	3.05	3.26	3.36	2.74
Fuel feedrate Kg/min	0.0624	0.0376	0.0624	0.0459	0.0417	0.0288	0.0384	0.0480
	3.9×S	2.4×S	3.9×S	2.9×S	2.6×S	1.8×S	2.4×S	3.0×S
Temp. K	989	978	990	981	983	971	980	978
Run duration mins	80	70	80	80	120	90	70	60

(b)

Bed depth = 10 cm, Temperature range = 971-990 K

Nm³/Kg ×10²	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8
CO	19.24	20.93	18.38	24.12	20.79	36.91	30.83	21.62
H_2	3.17	4.28	3.19	3.27	4.08	7.26	4.17	5.27
CH_4	2.40	2.74	2.44	3.18	3.60	4.76	3.72	3.10
C_2H_2	0.18	0.13	0.18	0.37	0.00	0.56	0.13	0.00
C_2H_4 / C_2H_6	1.18	1.52	1.20	1.48	3.02	2.88	2.03	0.92
C_3H_6	0.26	0.40	0.27	0.37	0.22	0.56	0.57	0.00
C_3H_8	0.08	0.08	0.08	0.04	0.02	0.07	0.13	0.00
O_2	3.77	11.52	3.86	3.05	1.85	5.21	9.22	2.04
N_2	96.97	160.03	96.97	131.33	144.44	208.51	156.72	125.64
CO_2	18.51	14.52	18.65	25.29	34.00	34.79	20.91	24.54
Total	145.74	216.14	145.22	192.14	212.01	301.49	228.44	183.15

(c)

Bed depth = 10 cm, Temperature range = 971–990 K

Nm³/min×10⁻³	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8
CO	11.99	7.87	11.47	11.07	8.67	10.63	11.84	10.38
H_2	1.98	1.61	1.99	1.50	1.70	2.09	1.60	2.53
CH_4	1.50	1.03	1.52	1.46	1.50	1.37	1.43	1.49
C_2H_2	0.11	0.05	0.11	0.17	0.00	0.16	0.05	0.00
C_2H_4 / C_2H_6	0.74	0.57	0.75	0.68	1.26	0.83	0.78	0.44
C_3H_6	0.16	0.15	0.17	0.17	0.09	0.16	0.22	0.00
C_3H_8	0.05	0.03	0.05	0.02	0.01	0.02	0.05	0.00
O_2	2.35	4.33	2.41	1.40	0.77	1.50	3.54	0.98
N_2	60.51	60.17	60.51	60.28	60.23	60.05	60.18	60.31
CO_2	11.55	5.46	11.64	11.61	14.18	10.02	8.03	11.78
Total	90.94	81.27	90.62	88.36	88.41	86.83	87.72	87.91

(d)

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8
Prod. gas flow								
Dry(Nm/min)	0.0901	0.0813	0.0906	0.0884	0.0884	0.0868	0.0877	0.0879
Wet(Nm/min)	0.0987	0.0864	0.0992	0.0946	0.0941	0.0908	0.0930	0.0945
Per Kg fuel								
Dry(Nm/Kg)	1.4446	2.1614	1.4524	1.9248	2.1201	3.0149	2.2846	1.8319
Wet(Nm/Kg)	1.5814	2.2984	1.4740	2.0619	2.2571	3.1517	2.4216	1.9688
Steam content of wet gas(%)	8.7	6.0	8.6	6.6	6.1	4.3	5.7	7.0
Carry-over (g/min)	0.175	0.211	0.297	0.223	0.195	0.095	0.178	0.201
Bed-char +ash (g/min)	0.250	0.150	0.250	0.184	0.167	0.115	0.154	0.192

(e)

Bed depth = 10 cm, Temperature range = 971–990 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	47.6	18.1	64.4	100.0	24.2	33.8
RUN 2	48.0	22.6	55.9	100.0	29.0	42.3
RUN 3	48.4	18.5	65.3	100.0	24.8	34.4
RUN 4	64.3	22.2	70.0	100.0	31.6	44.1
RUN 5	72.9	29.1	75.6	100.0	33.9	48.0
RUN 6	95.2	39.3	74.8	100.0	51.4	70.4
RUN 7	69.2	28.4	70.7	100.0	40.2	54.7
RUN 8	57.2	20.5	66.6	100.0	26.2	38.0

TABLE 3

Temperature = 991—1010 K
Air feedrate (15°C) = 80 lit/min

(d)

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10	RUN 11	RUN 12
CO	10.56	11.33	13.00	13.37	12.00	11.55	10.73	11.09	15.04	14.07	12.05	13.98
H_2	2.14	1.94	2.51	2.89	2.38	1.94	1.82	2.28	2.47	2.55	2.98	3.13
CH_4	1.19	2.06	1.68	2.86	1.32	2.10	2.08	1.61	2.00	2.98	3.05	2.56
C_2H_2	0	0	0.14	0.30	0	0	0	0.09	0.08	0.31	0.28	0.30
C_2H_4 / C_2H_6	0.74	1.42	0.87	1.64	0.85	1.47	1.75	0.95	1.10	1.75	1.55	1.63
C_3H_6	0.12	0.17	0.16	0.20	0.14	0.21	0.12	0.17	0.30	0.20	0.19	0.22
C_3H_8	0.01	0.01	0.02	0.02	0.01	0.02	0.01	0.17	0.04	0.02	0.02	0.02
O_2	0.96	3.08	5.99	1.66	1.23	3.99	2.89	2.25	3.55	2.01	1.98	1.00
N_2	67.85	66.71	64.21	61.61	65.61	65.13	68.10	72.18	66.16	60.38	63.25	61.18
CO_2	16.43	13.28	11.42	15.45	16.46	13.59	12.50	9.21	9.26	15.73	14.65	15.98
CV MJ/Nm³(dry)												
Gross	2.71	3.63	3.48	4.70	3.07	3.76	3.72	3.37	4.13	4.87	4.53	4.70
Net	2.58	3.43	3.31	4.42	2.92	3.55	3.51	3.18	3.92	4.59	4.26	4.43
Fuel feedrate Kg/min	0.0320	0.0417	0.0558	0.0525	0.0336	0.0417	0.0417	0.0398	0.0384	0.0576	0.0576	0.0576
	2.0× S	2.6× S	3.5× S	3.3× S	2.1× S	2.6× S	2.6× S	2.5× S	2.4× S	3.6× S	3.6× S	3.6× S
Temp. K	1003	999	1008	1004	1003	999	1001	993	997	1003	1006	1002
Run duration mins	115	120	60	60	115	120	120	90	70	60	60	60

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10	RUN 11	RUN 12
Prod gas flow												
Dry (Nm/min)	0.0886	0.0903	0.0941	0.0980	0.0916	0.0925	0.0884	0.0832	0.0909	0.1010	0.0956	0.0988
Wet (Nm/min)	0.0930	0.0960	0.1017	0.1052	0.0962	0.0982	0.0942	0.0871	0.0962	0.1080	0.1034	0.1067
Per Kg fuel												
Dry (Nm/Kg)	2.7678	2.1650	1.6862	1.8665	2.7271	2.2175	2.1209	2.0904	2.3688	1.7535	1.6590	1.7151
Wet (Nm/Kg)	2.9047	2.3019	1.8233	2.0034	2.8640	2.3544	2.2578	2.1894	2.5057	1.8748	1.7960	1.8521
Steam content of wet gas (%)	4.7	5.9	7.5	6.8	4.8	5.8	6.1	4.5	5.5	7.3	7.6	7.4
Carry-over (g/min)	0.088	0.117	0.267	0.259	0.301	0.248	0.321	0.311	0.233	0.315	0.355	0.341
Bed-char +ash (g/min)	0.128	0.167	0.223	0.210	0.134	0.167	0.178	0.115	0.154	0.230	0.230	0.243

(e)

$Nm^3/Kg\times10^3$	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10	RUN 11	RUN 12
CO	29.22	24.53	21.92	24.95	32.74	25.61	22.76	23.19	35.62	24.44	20.00	23.98
H_2	5.94	4.20	4.23	5.39	6.49	4.29	3.86	4.77	5.86	4.43	4.95	5.36
CH_4	3.28	4.46	2.83	5.33	3.60	4.65	4.41	3.37	4.74	5.17	5.05	4.39
C_2H_2	0.00	0.00	0.23	0.55	0.00	0.00	0.00	0.18	0.18	0.54	0.47	0.52
C_2H_4 / C_2H_6	2.06	3.07	1.47	3.07	2.32	3.26	3.72	1.98	2.60	3.04	2.57	2.80
C_3H_6	0.34	0.36	0.27	0.38	0.39	0.46	0.26	0.35	0.70	0.35	0.31	0.38
C_3H_8	0.03	0.02	0.04	0.04	0.03	0.05	0.02	0.35	0.10	0.03	0.03	0.03
O_2	2.66	6.67	10.11	3.10	3.36	8.85	6.14	4.70	8.41	3.49	3.28	1.72
N_2	187.78	144.44	108.28	114.99	178.93	144.44	144.44	130.88	156.72	104.93	104.93	104.93
CO_2	45.47	28.75	19.26	28.84	44.88	30.14	26.50	19.25	21.93	27.34	24.30	27.41
Total	276.78	216.50	168.62	186.65	272.71	221.75	212.09	209.04	236.88	175.35	165.90	171.51

Bed depth = 10 cm, Temperature range = 991–1010 K

(c)

	MASS BALANCE EFFICIENCIES %				GASIFICATION	THERMAL
	Carbon	Hydrogen	Oxygen	Nitrogen	EFFICIENCY %	EFFICIENCY %
RUN 1	93.0	27.9	86.3	100.0	37.3	55.3
RUN 2	72.7	32.7	78.0	100.0	38.9	53.6
RUN 3	54.0	21.9	78.2	100.0	29.2	40.7
RUN 4	75.6	38.0	83.3	100.0	43.2	56.9
RUN 5	97.3	30.8	91.9	100.0	41.6	59.6
RUN 6	76.5	34.7	84.7	100.0	41.2	56.4
RUN 7	69.2	33.9	72.0	100.0	39.0	53.6
RUN 8	80.8	39.3	62.8	100.0	48.1	66.8
RUN 9	78.6	36.5	74.7	100.0	48.6	64.5
RUN 10	73.0	35.9	85.0	100.0	41.8	54.6
RUN 11	63.1	34.3	74.2	100.0	37.0	49.3
RUN 12	71.1	34.1	81.2	100.0	39.8	52.4

Bed depth = 10 cm, Temperature range = 991 – 1010 K

$Nm^3/min\times10^3$	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10	RUN 11	RUN 12
CO	9.35	10.23	12.23	13.10	11.00	10.68	9.49	9.23	13.68	14.08	11.52	13.81
H_2	1.90	1.75	2.36	2.83	2.18	1.79	1.61	1.90	2.25	2.55	2.85	3.09
CH_4	1.05	1.86	1.58	2.80	1.21	1.94	1.84	1.34	1.82	2.98	2.91	2.53
C_2H_2	0.00	0.00	0.13	0.29	0.00	0.00	0.00	0.07	0.07	0.31	0.27	0.30
C_2H_4 / C_2H_6	0.66	1.28	0.82	1.61	0.78	1.36	1.55	0.79	1.00	1.75	1.48	1.61
C_3H_6	0.11	0.15	0.15	0.20	0.13	0.19	0.11	0.14	0.27	0.20	0.18	0.22
C_3H_8	0.01	0.01	0.02	0.02	0.01	0.02	0.01	0.14	0.04	0.02	0.02	0.02
O_2	0.85	2.78	5.64	1.63	1.13	3.69	2.56	1.87	3.23	2.01	1.89	0.99
N_2	60.09	60.23	60.42	60.37	60.12	60.23	60.23	60.05	60.18	60.44	60.44	60.44
CO_2	14.55	11.99	10.75	15.14	15.08	12.57	11.05	7.66	8.42	15.75	14.00	15.79
Total	88.57	90.28	94.09	97.99	91.63	92.47	88.44	83.20	90.96	100.10	95.56	98.79

TABLE 4

(a)

Bed depth = 10 cm

Temperature = 1011−1020 K

Air feedrate(15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4
CO	10·38	10·71	13·46	16·80
H₂	1·42	1·45	2·07	4·30
CH₄	1·40	1·43	2·41	1·85
C₂H₂	0·17	0·17	0	0·19
C₂H₄ / C₂H₆	0·66	0·68	1·07	1·00
C₃H₆	0·15	0·16	0·29	0·17
C₃H₈	0·02	0·02	0·02	0·01
O₂	3·24	3·48	5·49	3·88
N₂	72·94	71·83	63·89	59·04
CO₂	9·62	10·07	11·30	12·76
CV MJ/Nm³(dry)				
Gross	2·76	2·84	3·94	4·38
Net	2·63	2·71	3·74	4·15
Fuel feedrate Kg/min	0·0459	0·0459	0·0417	0·0560
	2·9 × S	2·9 × S	2·6 × S	3·5 × S
Temp. K	1018	1018	1013	1015
Run duration mins	80	80	120	60

(b)

Bed depth = 10 cm, Temperature range = 1011−1020 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4
CO	18·69	19·59	30·43	30·70
H₂	2·55	2·66	4·68	7·86
CH₄	2·53	2·61	5·44	3·38
C₂H₂	0·30	0·30	0·00	0·34
C₂H₄ / C₂H₆	1·20	1·24	2·42	1·82
C₃H₆	0·26	0·28	0·65	0·30
C₃H₈	0·04	0·04	0·05	0·02
O₂	5·84	6·36	12·42	7·09
N₂	131·33	131·33	144·44	107·89
CO₂	17·32	18·41	25·54	23·32
Total	180·06	182·83	226·07	182·75

(c)

Bed depth = 10 cm, Temperature range = 1011−1020 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3	RUN 4
CO	8·58	8·99	12·69	17·19
H₂	1·17	1·22	1·95	4·40
CH₄	1·16	1·20	2·27	1·89
C₂H₂	0·14	0·14	0·00	0·19
C₂H₄ / C₂H₆	0·55	0·57	1·01	1·02
C₃H₆	0·12	0·13	0·27	0·17
C₃H₈	0·02	0·02	0·02	0·01
O₂	2·68	2·92	5·18	3·97
N₂	60·28	60·28	60·23	60·42
CO₂	7·95	8·45	10·65	13·06
Total	82·65	83·92	94·27	102·34

(d)

	RUN 1	RUN 2	RUN 3	RUN 4
Prod. gas flow				
Dry (Nm/min)	0·0826	0·0839	0·0943	0·1023
Wet (Nm/min)	0·0889	0·0902	0·1000	0·1100
Per Kg fuel				
Dry (Nm/Kg)	1·8006	1·8283	2·2607	1·8275
Wet (Nm/Kg)	1·9375	1·9654	2·3976	1·9645
Steam content of wet gas(%)	7·1	7·0	5·7	7·0
Carry-over (g/min)	0·501	0·565	0·273	0·309
Bed-char +ash (g/min)	0·184	0·184	0·167	0·224

(e)

Bed depth = 10 cm, Temperature range = 1011 − 1020 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	47·4	17·6	56·3	100·0	24·8	36·6
RUN 2	50·0	18·3	59·9	100·0	25·9	38·0
RUN 3	76·4	35·4	86·9	100·0	44·2	59·8
RUN 4	70·1	30·6	89·0	100·0	39·7	52·8

TABLE 5

(a)

Bed depth = 10 cm
Temperature = 1021-1040 K
Air feedrate(15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
CO	8.95	10.84	13.24	10.64	11.61	12.00	9.84	13.06	13.09
H_2	1.86	1.94	2.65	2.24	2.85	1.55	1.59	1.88	2.75
CH_4	1.09	1.41	2.55	1.50	2.53	1.56	1.66	1.94	3.05
C_2H_2	0	0	0.19	0.13	0.19	0.20	0.19	0.15	0.29
C_2H_4 / C_2H_6	0.57	0.98	1.32	0.79	1.27	0.76	0.88	1.01	1.52
C_3H_6	0.11	0.14	0.23	0.13	0.21	0.18	0.13	0.19	0.22
C_3H_8	0.01	0.01	0.02	0.08	0.03	0.02	0.01	0.04	0.02
O_2	0.66	1.94	1.35	2.24	2.50	3.57	8.04	9.00	1.25
N_2	70.52	67.18	61.37	68.86	66.36	71.54	68.98	63.09	61.68
CO_2	16.23	15.56	17.08	13.39	12.45	8.62	8.68	9.64	16.13
CV MJ/Nm³(dry) Gross	2.30	2.99	4.27	3.05	4.04	3.16	2.95	3.66	4.65
Net	2.19	2.84	4.03	2.89	3.80	3.02	2.80	3.48	4.38
Fuel feedrate Kg/min	0.0320	0.0320	0.0499	0.0624	0.0682	0.0459	0.0512	0.0560	0.0592
	2.0×S	2.0×S	3.1×S	3.9×S	4.3×S	2.9×S	3.2×S	3.5×S	3.7×S
Temp. K	1040	1032	1024	1036	1040	1037	1031	1033	1035
Run duration mins	115	115	95	80	70	80	60	60	120

(b)

Bed depth = 10 cm, Temperature range = 1021-1040 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
CO	23.84	30.31	26.09	14.98	15.54	22.03	16.82	22.34	21.67
H_2	4.97	5.44	5.23	3.16	3.81	2.85	2.71	3.21	4.56
CH_4	2.91	3.94	5.03	2.12	3.39	2.85	2.83	3.32	5.05
C_2H_2	0.00	0.00	0.38	0.18	0.25	0.37	0.33	0.25	0.47
C_2H_4 / C_2H_6	1.53	2.75	2.60	1.10	1.70	1.39	1.50	1.73	2.52
C_3H_6	0.28	0.41	0.46	0.18	0.28	0.33	0.21	0.32	0.37
C_3H_8	0.03	0.03	0.04	0.11	0.04	0.04	0.02	0.07	0.03
O_2	1.75	5.44	2.66	3.16	3.34	6.56	13.73	15.39	2.08
N_2	187.78	187.78	120.92	96.97	88.84	131.33	117.89	107.89	102.13
CO_2	43.22	43.50	33.65	18.86	16.67	15.82	14.82	16.48	26.71
Total	266.31	279.53	197.03	140.82	133.87	183.59	170.90	171.02	165.59

(c)

Bed depth = 10 cm, Temperature range = 1021-1040 K

Nm³/min × 10³	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
CO	7.63	9.70	13.02	9.35	10.60	10.11	8.61	12.51	12.83
H_2	1.59	1.74	2.61	1.97	2.60	1.31	1.39	1.80	2.70
CH_4	0.93	1.26	2.51	1.32	2.31	1.31	1.45	1.86	2.99
C_2H_2	0.00	0.00	0.19	0.11	0.17	0.17	0.17	0.14	0.28
C_2H_4 / C_2H_6	0.49	0.88	1.30	0.69	1.16	0.64	0.77	0.97	1.49
C_3H_6	0.09	0.13	0.23	0.11	0.19	0.15	0.11	0.18	0.22
C_3H_8	0.01	0.01	0.02	0.07	0.03	0.02	0.01	0.04	0.02
O_2	0.56	1.74	1.33	1.97	2.28	3.01	7.03	8.62	1.23
N_2	60.09	60.09	60.34	60.51	60.59	60.28	60.36	60.42	60.46
CO_2	13.83	13.92	16.79	11.77	11.37	7.26	7.59	9.23	15.81
Total	85.22	89.45	98.32	87.87	91.30	84.27	87.50	95.77	98.03

(d)

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9
Prod. gas flow Dry(Nm/min)	0.0852	0.0894	0.0983	0.0879	0.0913	0.0843	0.0875	0.0980	0.0980
Wet(Nm/min)	0.0896	0.0938	0.1052	0.0964	0.1006	0.0906	0.0945	0.1034	0.1061
Per Kg fuel Dry(Nm/Kg)	2.6631	2.7953	1.9703	1.4082	1.3387	1.8359	1.7090	1.7102	1.6559
Wet(Nm/Kg)	2.8000	2.9322	2.1072	1.5452	1.4756	1.9728	1.8459	1.8471	1.7929
Steam content of wet gas(%)	4.9	4.7	6.5	8.9	9.3	6.9	7.4	7.4	7.6
Carry-over (g/min)	0.207	0.199	0.127	0.357	0.402	0.397	0.369	0.315	0.299
Bed-char + ash (g/min)	0.128	0.128	0.200	0.250	0.273	0.184	0.205	0.224	0.237

(e)

Bed depth = 10 cm, Temperature range = 1021-1040 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	82.7	23.1	78.3	100.0	30.5	48.4
RUN 2	94.4	31.9	88.1	100.0	41.5	60.5
RUN 3	80.7	35.2	89.9	100.0	41.6	56.1
RUN 4	44.1	16.9	60.8	100.0	21.3	31.3
RUN 5	45.2	23.9	59.9	100.0	26.6	36.8
RUN 6	50.7	20.2	57.9	100.0	29.0	41.4
RUN 7	43.4	19.6	68.4	100.0	25.0	36.6
RUN 8	52.9	23.3	83.7	100.0	31.1	43.1
RUN 9	67.8	33.7	79.4	100.0	37.9	50.7

TABLE 6

(a)

Bed depth = 10 cm

Temperature = 1041-1060 K

Air feedrate(15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10
CO	13.23	17.46	13.68	13.34	18.44	10.96	17.52	18.78	14.45	13.15
H_2	2.99	3.17	3.01	3.03	3.19	2.52	3.41	3.15	3.31	2.93
CH_4	2.78	2.55	2.79	2.79	2.55	2.83	3.08	2.49	2.94	2.50
C_2H_2	0.23	0.20	0.25	0.24	0.20	0.18	0.29	0.18	0.25	0.20
C_2H_4 / C_2H_6	1.30	1.43	1.39	1.33	1.44	1.01	1.66	1.45	1.31	1.25
C_3H_6	0.22	0.33	0.22	0.22	0.36	0.18	0.24	0.32	0.23	0.19
C_3H_8	0.01	0.04	0.02	0.02	0.04	0.01	0.01	0.04	0.01	0.01
O_2	0.98	2.47	1.70	1.06	2.50	1.06	0.45	2.78	1.05	2.13
N_2	61.53	60.08	60.95	61.20	59.01	64.47	59.21	58.54	59.22	62.01
CO_2	16.73	12.27	15.99	16.77	12.27	16.78	14.13	12.27	17.23	15.63
CV MJ/Nm³(dry) Gross	4.40	5.07	4.54	4.46	5.23	3.81	5.41	5.20	4.68	4.19
Net	4.14	4.81	4.28	4.20	4.96	3.58	5.12	4.94	4.42	3.95
Fuel feedrate Kg/min	0.0499	0.0682	0.0512	0.0499	0.0682	0.0499	0.0499	0.0682	0.0592	0.0560
	3.1×S	4.3×S	3.2×S	3.1×S	4.3×S	3.1×S	3.1×S	4.3×S	3.7×S	3.5×S
Temp. K	1051	1052	1056	1051	1052	1054	1060	1056	1053	1060
Run duration mins	95	70	120	100	80	95	100	70	120	120

(c)

Bed depth = 10 cm, Temperature range = 1041-1060 K

Nm³/Kg ×10²	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10
CO	25.99	25.82	26.46	26.35	27.76	20.56	35.77	28.50	24.92	22.88
H_2	5.87	4.69	5.82	5.99	4.81	4.73	6.95	4.78	5.71	5.09
CH_4	5.47	3.77	5.39	5.51	3.84	5.31	6.29	3.78	5.07	4.36
C_2H_2	0.46	0.29	0.49	0.48	0.31	0.34	0.60	0.28	0.44	0.34
C_2H_4 / C_2H_6	2.54	2.11	2.70	2.62	2.17	1.90	3.39	2.20	2.26	2.18
C_3H_6	0.44	0.48	0.43	0.44	0.54	0.34	0.48	0.48	0.39	0.34
C_3H_8	0.02	0.06	0.04	0.04	0.06	0.02	0.02	0.06	0.02	0.02
O_2	1.92	3.65	3.28	2.10	3.77	1.98	0.92	4.22	1.81	3.71
N_2	120.92	88.84	117.89	120.92	88.84	120.92	120.92	88.84	102.13	107.89
CO_2	32.88	18.14	30.92	33.13	18.48	31.46	28.86	18.62	29.71	27.20
Total	196.51	147.86	193.40	197.58	150.54	187.56	204.21	151.76	172.47	174.00

(e)

Bed depth = 10 cm, Temperature range = 1041-1060 K

Nm³/min ×10³	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10
CO	12.97	17.61	13.55	13.15	18.93	10.26	17.85	19.44	14.75	12.81
H_2	2.93	3.20	2.98	2.99	3.28	2.36	3.47	3.26	3.38	2.85
CH_4	2.73	2.57	2.76	2.75	2.62	2.65	3.14	2.58	3.00	2.44
C_2H_2	0.23	0.20	0.25	0.24	0.21	0.17	0.30	0.19	0.26	0.19
C_2H_4 / C_2H_6	1.27	1.44	1.38	1.31	1.48	0.95	1.69	1.50	1.34	1.22
C_3H_6	0.22	0.33	0.22	0.22	0.37	0.17	0.24	0.33	0.23	0.19
C_3H_8	0.01	0.04	0.02	0.02	0.04	0.01	0.01	0.04	0.01	0.01
O_2	0.96	2.49	1.68	1.05	2.57	0.99	0.46	2.88	1.07	2.08
N_2	60.34	60.59	60.36	60.34	60.59	60.34	60.34	60.59	60.46	60.42
CO_2	16.41	12.37	15.83	16.53	12.60	15.70	14.40	12.70	17.59	15.23
Total	98.06	100.84	99.02	98.59	102.67	93.59	101.90	103.50	102.10	97.44

(d)

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7	RUN 8	RUN 9	RUN 10
Prod gas flow Dry (Nm/min)	0.0981	0.1008	0.0990	0.0986	0.1027	0.0936	0.1019	0.1035	0.1021	0.0974
Wet (Nm/min)	0.1049	0.1102	0.1060	0.1054	0.1120	0.1004	0.1087	0.1128	0.1102	0.1051
Per Kg fuel Dry (Nm/Kg)	1.9651	1.4786	1.9340	1.9758	1.5054	1.8756	2.0421	1.5176	1.7247	1.7400
Wet (Nm/Kg)	2.1022	1.6155	2.0711	2.1126	1.6424	2.0124	2.1792	1.6545	1.8616	1.8770
Steam content of wet gas (%)	6.5	8.5	6.6	6.5	8.3	6.8	6.3	8.3	7.4	7.3
Carry-over (g/min)	0.300	0.377	0.409	0.435	0.459	0.371	0.404	0.389	0.278	0.431
Bed-char + ash (g/min)	0.200	0.273	0.205	0.200	0.273	0.200	0.200	0.273	0.237	0.224

Bed depth = 10 cm, Temperature range = 1041-1060 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	80.2	37.2	87.1	100.0	42.6	57.7
RUN 2	60.6	29.0	74.8	100.0	37.2	48.6
RUN 3	78.8	37.6	87.7	100.0	43.4	58.3
RUN 4	81.2	38.0	88.2	100.0	43.4	58.6
RUN 5	63.5	29.8	77.9	100.0	39.1	50.8
RUN 6	70.3	32.0	79.7	100.0	35.1	49.3
RUN 7	89.9	44.9	86.8	100.0	54.7	70.6
RUN 8	64.3	29.5	80.0	100.0	39.2	51.0
RUN 9	74.2	34.5	88.1	100.0	39.9	53.4
RUN 10	67.7	30.7	82.3	100.0	36.0	49.4

TABLE 7

(a)

Bed depth = 10 cm

Temperature = 1061 – 1080 K

Air feedrate (15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4
CO	13.45	21.98	14.95	12.78
H_2	3.18	3.50	3.47	2.57
CH_4	2.79	2.62	3.34	2.14
C_2H_2	0.30	0.22	0.33	0.20
C_2H_4 / C_2H_6	1.35	1.58	1.78	1.09
C_3H_6	0.23	0.46	0.27	0.18
C_3H_8	0.01	0.05	0.03	0.01
O_2	0.91	2.14	0.97	3.08
N_2	60.74	55.37	58.21	63.64
CO_2	17.04	12.08	16.65	14.31
CV MJ/Nm³(dry)				
Gross	4.54	5.95	5.35	3.83
Net	4.28	5.67	5.04	3.63
Fuel feedrate	0.0499	0.0682	0.0656	0.0512
Kg/min	3.1× S	4.3× S	4.1× S	3.2× S
Temp. K	1067	1061	1064	1070
Run duration mins	95	70	120	120

(b)

Bed depth = 10 cm, Temperature range = 1061–1080 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4
CO	26.77	35.26	23.70	23.67
H_2	6.33	5.62	5.50	4.76
CH_4	5.55	4.21	5.29	3.96
C_2H_2	0.60	0.35	0.52	0.37
C_2H_4 / C_2H_6	2.68	2.55	2.82	2.01
C_3H_6	0.46	0.73	0.43	0.33
C_3H_8	0.02	0.07	0.04	0.02
O_2	1.80	3.43	1.54	5.70
N_2	120.92	88.84	92.30	117.89
CO_2	33.93	19.38	26.40	26.50
Total	199.08	160.44	158.67	185.23

(c)

Bed depth = 10 cm, Temperature range = 1061–1080 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3	RUN 4
CO	13.36	24.05	15.55	12.12
H_2	3.16	3.83	3.61	2.44
CH_4	2.77	2.87	3.47	2.03
C_2H_2	0.30	0.24	0.34	0.19
C_2H_4 / C_2H_6	1.34	1.74	1.85	1.03
C_3H_6	0.23	0.50	0.28	0.17
C_3H_8	0.01	0.05	0.03	0.01
O_2	0.90	2.34	1.01	2.92
N_2	60.34	60.59	60.55	60.36
CO_2	16.93	13.22	17.32	13.57
Total	99.34	109.42	104.02	94.84

(d)

	RUN 1	RUN 2	RUN 3	RUN 4
Prod. gas flow				
Dry (Nm³/min)	0.0993	0.1094	0.1040	0.0948
Wet (Nm³/min)	0.1062	0.1188	0.1130	0.1018
Per Kg fuel				
Dry (Nm³/Kg)	1.9908	1.6044	1.5857	1.8523
Wet (Nm³/Kg)	2.1276	1.7413	1.7227	1.9892
Steam content of wet gas (%)	6.4	7.6	7.9	6.9
Carry-over (g/min)	0.079	0.137	0.205	0.211
Bed-char + ash (g/min)	0.200	0.273	0.262	0.205

(e)

Bed depth = 10 cm, Temperature range = 1061 – 1080 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	83.0	39.0	89.5	100.0	44.6	60.2
RUN 2	75.0	34.6	87.2	100.0	47.6	60.3
RUN 3	71.0	37.4	84.1	100.0	41.8	54.8
RUN 4	67.0	28.5	81.4	100.0	35.2	49.1

TABLE 8

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TABLE 3·20(a)

Bed depth = 15 cm

Temperature = 749–850 K

Air feedrate = 80 lit/min

	RUN 1	RUN 2	RUN 3
CO	13·58	11·98	4·44
H_2	1·11	1·23	0·29
CH_4	0·91	0·94	0·47
C_2H_2	0·15	0·17	0·11
C_2H_6 C_2H_4	0·42	0·45	0·28
C_3H_6	0·01	0·01	0·03
C_3H_8	0	0	0
O_2	2·82	5·76	9·58
N_2	73·27	69·95	69·90
CO_2	7·73	9·51	14·90
CV MJ/Nm³(dry)			
Gross	2·61	2·47	1·07
Net	2·53	2·38	1·03
Fuel feedrate Kg/min	0·0400	0·0400	0·0454
	2·5 × S	2·5 × S	2·8 × S
Temp. K	823	813	749
Run duration mins	60	60	60

TABLE 3·20 (b)

Bed depth = 15 cm, Temperature = 749–850 K

Nm³/Kg × 10⁷	RUN 1	RUN 2	RUN 3
CO	27·90	25·78	8·44
H_2	2·28	2·65	0·55
CH_4	1·88	2·02	0·90
C_2H_2	0·30	0·38	0·20
C_2H_6 C_2H_4	0·88	0·98	0·53
C_3H_6	0·02	0·02	0·07
C_3H_8	0·00	0·00	0·00
O_2	5·80	12·40	18·19
N_2	150·50	150·50	132·78
CO_2	15·88	20·45	28·30
Total	205·42	215·18	189·93

TABLE 3·20 (c)

Bed depth = 15 cm, Temperature = 759–850 K

Nm³/min × 10³	RUN 1	RUN 2	RUN 3
CO	11·16	10·31	3·83
H_2	0·91	1·06	0·25
CH_4	0·75	0·81	0·41
C_2H_2	0·12	0·15	0·09
C_2H_6 C_2H_4	0·35	0·39	0·24
C_3H_6	0·01	0·01	0·03
C_3H_8	0·00	0·00	0·00
O_2	2·32	4·96	8·26
N_2	60·20	60·20	60·28
CO_2	6·35	8·18	12·85
Total	82·17	86·07	86·23

TABLE 3·20 (d)

Bed depth = 15 cm, Temperature = 749–850 K

	RUN 1	RUN 2	RUN 3
Prod. gas flow Dry(Nm³/min)	0·0822	0·0861	0·0862
Wet(Nm³/min)	0·0876	0·0915	0·0924
Per Kg fuel Dry(Nm³/Kg)	2·054	2·152	1·899
Wet(Nm³/Kg)	2·191	2·289	2·036
Steam content of wet gas(%)	6·3	6·0	6·7
Carry-over (g/min)	0·028	0·031	0·024
Bed char + ash (g/min)	0·160	0·160	0·182

TABLE 3·20 (e)

Bed depth = 15 cm, Temperature range = 749 – 850 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	53·7	12·7	56·7	100·0	27·2	39·2
RUN 2	57·0	14·2	72·9	100·0	25·1	39·1
RUN 3	43·9	·6·0	87·3	100·0	10·2	19·1

TABLE 9

(a)

Bed depth = 15 cm
Temperature = 851-950 K
Air feedrate = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4
CO	12.94	9.60	8.67	8.23
H_2	1.56	0.66	1.02	0.88
CH_4	1.29	0.44	1.38	1.50
C_2H_2	0.16	0.01	0.12	0.02
C_2H_4 C_2H_6	0.68	0.10	0.66	0.19
C_3H_6	0.13	0.01	0.16	0.10
C_3H_8	0.01	0	0.02	0.03
O_2	2.37	9.04	2.07	1.80
N_2	62.11	75.61	68.20	68.65
CO_2	18.75	4.53	17.70	18.60
CV MJ/Nm³(dry)				
Gross	3.04	1.56	2.46	2.02
Net	2.91	1.53	2.34	1.92
Fuel feedrate Kg/min	0.0427	0.0400	0.0457	0.0390
	2.7 × S	2.5 × S	2.9 × S	2.4 × S
Temp. K	938	873	875	863
Run duration mins	60	60	60	35

(b)

Bed depth = 15 cm, Temperature range = 851 – 950 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4
CO	29.39	19.10	16.76	18.51
H_2	3.54	1.32	1.97	1.97
CH_4	2.93	0.88	2.67	3.38
C_2H_2	0.37	0.02	0.24	0.05
C_2H_4 C_2H_6	1.54	0.20	1.27	0.44
C_3H_6	0.30	0.02	0.31	0.23
C_3H_8	0.02	0.00	0.04	0.08
O_2	5.39	18.00	4.00	4.05
N_2	141.08	150.50	131.90	154.33
CO_2	42.60	9.02	34.22	41.36
Total	227.14	199.05	193.41	224.82

(c)

Bed depth = 15 cm, Temperature = 851 – 950 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3	RUN 4
CO	12.55	7.64	7.66	7.22
H_2	1.51	0.53	0.90	0.77
CH_4	1.25	0.35	1.22	1.32
C_2H_2	0.16	0.01	0.11	0.02
C_2H_4 C_2H_6	0.66	0.08	0.58	0.17
C_3H_6	0.13	0.01	0.14	0.09
C_3H_8	0.01	0.00	0.02	0.03
O_2	2.30	7.20	1.83	1.58
N_2	60.24	60.20	60.28	60.19
CO_2	18.19	3.61	15.64	16.31
Total	96.99	79.62	88.39	87.68

(d)

Bed depth = 15 cm, Temperature = 851 – 950 K

	RUN 1	RUN 2	RUN 3	RUN 4
Prod. gas flow				
Dry(Nm³/min)	0.0970	0.0796	0.0884	0.0877
Wet(Nm³/min)	0.1028	0.0851	0.0946	0.0930
Per Kg fuel				
Dry(Nm³/Kg)	2.271	1.991	1.934	2.248
Wet(Nm³/Kg)	2.408	2.128	2.071	2.385
Steam content of wet gas (%)	5.7	6.4	6.6	5.7
Carry-over (g/min)	0.024	0.044	0.018	0.020
Bed char + ash (g/min)	0.171	0.160	0.183	0.156

(e)

Bed depth = 15 cm, Temperature range = 851 – 950 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	89.1	21.7	104.0	100.0	34.6	51.3
RUN 2	33.0	5.4	58.2	100.0	15.9	27.1
RUN 3	64.5	17.6	80.6	100.0	23.7	36.8
RUN 4	73.2	16.0	86.4	100.0	22.6	36.5

TABLE 10

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(a)

Bed depth = 15 cm

Temperature = 951–1050 K

Air feedrate = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7
CO	6.81	14.99	11.06	7.22	10.42	18.17	21.17
H_2	1.06	1.92	1.45	1.30	0.85	3.10	3.18
CH_4	1.10	2.75	1.86	0.57	2.28	2.80	3.20
C_2H_2	0.10	0.15	0.14	0.02	0.08	0.09	0.11
C_2H_4 / C_2H_6	0.51	1.32	0.93	0.22	0.92	1.09	1.19
C_3H_6	0.15	0.19	0.17	0.07	0.32	0.32	0.44
C_3H_8	0.02	0.02	0.02	0.01	0.03	0.04	0.09
O_2	2.01	1.14	0.88	1.52	4.67	2.10	2.00
N_2	76.80	61.44	67.69	70.77	71.64	56.47	52.81
CO_2	11.44	16.08	15.80	18.30	8.79	15.82	15.81
CV MJ/Nm³(dry)							
Gross	2.00	4.42	3.22	1.55	3.33	4.94	5.74
Net	1.90	4.19	3.06	1.48	3.16	4.70	5.45
Fuel feedrate Kg/min	0.0457	0.0457	0.0457	0.0543	0.0543	0.0543	0.0688
	2.9×S	2.9×S	2.9×S	3.4×S	3.4×S	3.4×S	4.3×S
Temp. K	963	973	984	992	1029	1006	1000
Run duration mins	40	40	40	35	50	35	35

(b)

Bed depth = 15 cm, Temperature range = 951–1050 K

$Nm^3/Kg \times 10^3$	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7
CO	11.71	32.19	21.55	11.34	16.17	35.78	35.31
H_2	1.82	4.11	2.82	2.04	1.32	6.11	5.30
CH_4	1.88	5.91	3.63	0.90	3.54	5.51	5.33
C_2H_2	0.18	0.33	0.26	0.04	0.13	0.18	0.19
C_2H_4 / C_2H_6	0.88	2.84	1.82	0.35	1.44	2.15	1.99
C_3H_6	0.26	0.42	0.33	0.11	0.50	0.63	0.73
C_3H_8	0.04	0.04	0.04	0.02	0.06	0.07	0.14
O_2	3.46	2.45	1.71	2.39	7.26	4.14	3.33
N_2	131.90	131.90	131.90	111.23	111.23	111.23	88.07
CO_2	19.65	34.53	30.79	28.77	13.65	31.16	26.37
Total	171.75	214.68	194.86	157.16	155.28	196.98	166.77

(c)

Bed depth = 15 cm, Temperature = 951–1050 K

$Nm^3/min \times 10^{-3}$	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7
CO	5.35	14.71	9.85	6.16	8.78	19.43	24.29
H_2	0.83	1.88	1.29	1.11	0.72	3.32	3.65
CH_4	0.86	2.70	1.66	0.49	1.92	2.99	3.67
C_2H_2	0.08	0.15	0.12	0.02	0.07	0.10	0.13
C_2H_4 / C_2H_6	0.40	1.30	0.83	0.19	0.78	1.17	1.37
C_3H_6	0.12	0.19	0.15	0.06	0.27	0.34	0.50
C_3H_8	0.02	0.02	0.02	0.01	0.03	0.04	0.10
O_2	1.58	1.12	0.78	1.30	3.94	2.25	2.29
N_2	60.28	60.28	60.28	60.40	60.40	60.40	60.59
CO_2	8.98	15.78	14.07	15.62	7.41	16.92	18.14
Total	78.49	98.11	89.05	85.34	84.31	106.96	114.74

(d)

Bed depth = 15 cm, Temperature = 951–1050 K

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6	RUN 7
Prod. gas flow Dry(Nm³/min)	0.0785	0.0981	0.0891	0.0853	0.0843	0.1070	0.1147
Wet(Nm³/min)	0.0848	0.1044	0.0953	0.0928	0.0917	0.1144	0.1242
Per Kg fuel Dry(Nm³/Kg)	1.718	2.147	1.949	1.572	1.553	1.970	1.668
Wet(Nm³/Kg)	1.854	2.284	2.086	1.709	1.690	2.107	1.805
Steam content of wet gas (%)	7.4	6.0	6.6	8.0	8.1	6.5	7.6
Carry-over (g/min)	0.018	0.018	0.021	0.025	0.038	0.014	0.028
Bed char + ash (g/min)	0.183	0.183	0.183	0.217	0.217	0.217	0.275

(e)

Bed depth = 15 cm, Temperature range = 951–1050 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION	THERMAL
	Carbon	Hydrogen	Oxygen	Nitrogen	EFFICIENCY %	EFFICIENCY %
RUN 1	40.5	13.2	50.1	100.0	17.1	29.6
RUN 2	89.7	36.8	91.8	100.0	47.1	65.7
RUN 3	68.5	23.8	74.8	100.0	31.2	47.1
RUN 4	47.1	7.6	70.4	100.0	12.2	24.0
RUN 5	42.6	20.5	55.4	100.0	25.7	39.6
RUN 6	88.6	37.0	101.6	100.0	48.4	67.1
RUN 7	82.7	35.6	102.6	100.0	47.6	65.0

TABLE 11

(a)

Bed depth = 20 cm

Temperature = 751-800 K

Air feedrate (15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4
CO	3·14	5·73	8·18	2·72
H_2	0·12	0·10	0·30	0·22
CH_4	0·62	0·24	0·56	0·20
C_2H_2	0·01	0	0	0·01
C_2H_4 / C_2H_6	0·31	0·05	0·09	0·05
C_3H_6	0·05	0·01	0·02	0·01
C_3H_8	0·01	0	0·01	0
O_2	6·83	7·79	4·28	9·32
N_2	83·35	76·22	73·03	77·98
CO_2	5·56	9·86	13·53	9·49
CV MJ/Nm³(dry)				
Gross	0·93	0·88	1·39	0·50
Net	0·89	0·86	1·36	0·49
Fuel feedrate Kg/min	0·0643	0·0495	0·0495	0·0328
	4·0× S	3·1× S	3·1× S	2·1× S
Temp. K	785	751	794	793
Run duration mins	45	40	40	60

(b)

Bed depth = 20 cm, Temperature range = 751-800 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4
CO	3·54	9·17	13·66	6·40
H_2	0·14	0·16	0·50	0·52
CH_4	0·70	0·38	0·93	0·46
C_2H_2	0·02	0·00	0·00	0·03
C_2H_4 / C_2H_6	0·36	0·08	0·14	0·12
C_3H_6	0·06	0·02	0·04	0·03
C_3H_8	0·02	0·00	0·02	0·00
O_2	7·71	12·46	7·15	21·89
N_2	94·14	121·88	121·88	183·26
CO_2	6·28	15·76	22·58	22·29
Total	112·95	159·92	166·89	235·00

(c)

Bed depth = 20 cm, Temperature = 751-800 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3	RUN 4
CO	2·28	4·54	6·76	2·10
H_2	0·09	0·08	0·25	0·17
CH_4	0·45	0·19	0·46	0·15
C_2H_2	0·01	0·00	0·00	0·01
C_2H_4 / C_2H_6	0·23	0·04	0·07	0·04
C_3H_6	0·04	0·01	0·02	0·01
C_3H_8	0·01	0·00	0·01	0·00
O_2	4·96	6·17	3·54	7·18
N_2	60·53	60·33	60·33	60·11
CO_2	4·04	7·80	11·18	7·31
Total	72·63	79·16	82·61	77·08

(d)

Bed depth = 20 cm, Temperature = 751-800 K

	RUN 1	RUN 2	RUN 3	RUN 4
Prod. gas flow				
Dry (Nm³/min)	0·0726	0·0792	0·0826	0·0771
Wet (Nm³/min)	0·0814	0·0859	0·0894	0·0816
Per Kg fuel				
Dry (Nm³/Kg)	1·129	1·599	1·669	2·350
Wet (Nm³/Kg)	1·266	1·736	1·806	2·487
Steam content of wet gas (%)	10·8	7·9	7·6	5·5
Carry-over (g/min)	0·280	0·500	0·500	0·268
Bed-char + ash (g/min)	0·257	0·198	0·198	0·131

(e)

Bed depth = 20 cm, Temperature range = 751-800 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	12·8	3·9	33·0	100·0	5·2	11·3
RUN 2	28·5	1·7	59·5	100·0	7·2	14·5
RUN 3	42·1	4·3	66·3	100·0	11·8	20·6
RUN 4	33·0	2·7	66·3	100·0	6·0	16·4

TABLE 12

(a)

Bed depth = 20 cm
Temperature = 801–900 K
Air feedrate (15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6
CO	7.34	8.46	9.50	5.40	7.38	8.09
H_2	0.49	0.63	1.05	0.56	0.78	0.88
CH_4	0.91	0.80	1.30	0.68	1.01	1.18
C_2H_2	0.02	0	0.01	0.02	0.03	0.03
C_2H_4 / C_2H_6	0.40	0.13	0.24	0.22	0.37	0.41
C_3H_6	0.10	0.03	0.07	0.07	0.13	0.15
C_3H_8	0.01	0.01	0.02	0.01	0.01	0.03
O_2	2.24	1.79	2.57	3.51	3.34	2.73
N_2	71.83	72.87	69.52	74.78	71.25	71.24
CO_2	16.66	15.28	15.72	14.75	15.70	15.26
CV MJ/Nm³(dry) Gross	1.74	1.60	2.11	1.26	1.84	2.07
Net	1.67	1.55	2.02	1.21	1.75	1.98
Fuel feedrate Kg/min	0.0643	0.0495	0.0495	0.0328	0.0328	0.0328
	4.0× S	3.1× S	3.1× S	2.1× S	2.1× S	2.1× S
Temp. K	853	831	856	855	879	893
Run duration mins	45	40	40	60	60	60

(b)

Bed depth = 20 cm, Temperature range = 801 – 900 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6
CO	9.63	14.14	16.65	13.23	18.99	20.82
H_2	0.64	1.05	1.84	1.37	2.01	2.26
CH_4	1.20	1.33	2.28	1.68	2.59	3.05
C_2H_2	0.03	0.00	0.02	0.06	0.09	0.09
C_2H_4 / C_2H_6	0.53	0.22	0.42	0.55	0.94	1.07
C_3H_6	0.12	0.04	0.12	0.18	0.34	0.40
C_3H_8	0.02	0.02	0.04	0.03	0.03	0.09
O_2	2.94	2.99	4.50	8.60	8.60	7.01
N_2	94.14	121.88	121.88	183.26	183.26	183.26
CO_2	21.84	25.56	27.56	36.16	40.36	39.24
Total	131.06	167.25	175.31	245.06	257.20	257.22

(c)

Bed depth = 20 cm, Temperature range = 801 – 900 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6
CO	6.19	7.00	8.24	4.34	6.23	6.83
H_2	0.41	0.52	0.91	0.45	0.66	0.74
CH_4	0.77	0.66	1.13	0.55	0.85	1.00
C_2H_2	0.02	0.00	0.01	0.02	0.03	0.03
C_2H_4 / C_2H_6	0.34	0.11	0.21	0.18	0.31	0.35
C_3H_6	0.08	0.02	0.06	0.06	0.11	0.13
C_3H_8	0.01	0.01	0.02	0.01	0.01	0.03
O_2	1.89	1.48	2.23	2.82	2.82	2.30
N_2	60.53	60.33	60.33	60.11	60.11	60.11
CO_2	14.04	12.65	13.64	11.86	13.24	12.87
Total	84.27	82.79	86.78	80.38	84.36	84.37

(d)

Bed depth = 20 cm, Temperature = 801 – 900 K

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6
Prod. gas flow Dry(Nm³/min)	0.0843	0.0828	0.0868	0.0804	0.0844	0.0844
Wet(Nm³/min)	0.0931	0.0896	0.0936	0.0849	0.0888	0.0899
Per Kg fuel Dry (Nm³/min)	1.311	1.673	1.753	2.450	2.572	2.572
Wet(Nm³/min)	1.448	1.810	1.890	2.587	2.709	2.709
Steam content of wet gas (%)	9.5	7.6	7.2	5.3	5.1	5.1
Carry-over (g/min)	0.425	0.500	0.500	0.268	0.268	0.268
Bed-char + ash (g/min)	0.257	0.198	0.198	0.131	0.131	0.131

(e)

Bed depth = 20 cm, Temperature range = 801 – 900 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION	THERMAL
	Carbon	Hydrogen	Oxygen	Nitrogen	EFFICIENCY %	EFFICIENCY %
RUN 1	38.2	7.2	61.9	100.0	11.4	20.1
RUN 2	46.6	6.7	64.6	100.0	13.5	23.3
RUN 3	53.5	11.9	73.2	100.0	18.6	29.8
RUN 4	59.0	10.0	71.9	100.0	15.5	28.9
RUN 5	72.8	16.0	81.8	100.0	23.6	39.0
RUN 6	74.6	18.6	79.3	100.0	26.6	42.6

TABLE 13

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(a)

Bed depth = 20 cm
Temperature = 901 – 1050 K
Air feedrate (15°C) = 80 lit/min

	RUN 1	RUN 2	RUN 3
CO	15.23	13.88	10.38
H_2	2.80	2.69	1.48
CH_4	1.01	1.31	1.58
C_2H_2	0.02	0.02	0.15
C_2H_4 / C_2H_6	0.55	0.58	0.69
C_3H_6	0.11	0.11	0.20
C_3H_8	0.01	0.02	0.01
O_2	2.17	0.96	1.98
N_2	62.01	64.64	68.14
CO_2	16.09	15.79	15.39
CV MJ/Nm³ (dry)			
Gross	3.19	3.15	2.89
Net	3.06	3.01	2.74
Fuel feedrate Kg/min	0.0643 / 4.0× S	0.0643 / 4.0× S	0.0328 / 2.1× S
Temp. K	973	1003	904
Run duration mins	45	45	60

(b)

Bed depth = 20 cm, Temperature = 901–1050 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3
CO	23.12	20.22	27.93
H_2	4.24	3.92	3.99
CH_4	1.54	1.91	4.24
C_2H_2	0.03	0.03	0.40
C_2H_4 / C_2H_6	0.84	0.84	1.86
C_3H_6	0.17	0.16	0.55
C_3H_8	0.02	0.03	0.03
O_2	3.30	1.40	5.34
N_2	94.14	94.14	183.26
CO_2	24.43	23.00	41.40
Total	151.84	145.65	268.93

(c)

Bed depth = 20 cm, Temperature = 901–1050 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3
CO	14.87	13.00	9.16
H_2	2.73	2.52	1.31
CH_4	0.99	1.23	1.39
C_2H_2	0.02	0.02	0.13
C_2H_4 / C_2H_6	0.54	0.54	0.61
C_3H_6	0.11	0.10	0.18
C_3H_8	0.01	0.02	0.01
O_2	2.12	0.90	1.75
N_2	60.53	60.53	60.11
CO_2	15.71	14.79	13.58
Total	97.63	93.65	88.21

(d)

Bed depth = 20 cm, Temperature = 901–1050 K

	RUN 1	RUN 2	RUN 3
Prod. gas flow			
Dry (Nm³/min)	0.0976	0.0936	0.0882
Wet (Nm³/min)	0.1064	0.1024	0.0927
Per Kg fuel			
Dry (Nm³/min)	1.518	1.456	2.689
Wet (Nm³/min)	1.655	1.593	2.826
Steam content of wet gas (%)	8.3	8.6	4.8
Carry-over (g/min)	0.358	0.550	0.268
Bed-char + ash (g/min)	0.257	0.257	0.131

(e)

Bed depth = 20 cm, Temperature range = 901 – 1050 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	57.4	14.9	82.2	100.0	24.3	37.2
RUN 2	53.0	15.6	72.2	100.0	22.9	35.9
RUN 3	89.2	28.6	84.9	100.0	38.6	56.7

TABLE 14

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(a)

Bed depth = 5 cm
Temperature range = 801 − 900 K

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
CO	16·62	9·40	5·82	6·51	5·38
H_2	2·25	1·42	0·85	0·90	0·73
CH_4	1·64	1·22	0·79	1·03	0·24
C_2H_2	0·23	0·19	0·08	0·11	0·03
C_2H_4 / C_2H_6	0·52	0·49	0·25	0·40	0·10
C_3H_6	0·07	0·06	0·04	0·05	0·01
C_3H_8	0·01	0·02	0·01	0·02	0·00
O_2	3·99	4·54	4·98	4·37	1·30
N_2	67·61	72·56	80·29	76·62	87·16
CO_2	7·06	10·10	6·89	6·99	5·05
CV MJ/Nm³(dry)					
Gross	3·62	2·38	1·43	1·75	0·97
Net	3·47	2·27	1·36	1·67	0·94
Fuel feedrate Kg/min	0·0576	0·0224	0·0224	0·0224	0·0288
	3·6 × S	1·4 × S	1·4 × S	1·4 × S	1·8 × S
Temp. K	875	801	832	859	852
Run duration mins	40	30	40	40	30

(b)

Bed depth = 5 cm, Temperature range = 801 − 900 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
CO	25·80	34·69	19·42	21·88	12·88
H_2	3·49	5·22	2·81	3·04	1·74
CH_4	2·55	4·51	2·63	3·48	0·59
C_2H_2	0·36	0·71	0·27	0·36	0·07
C_2H_4 / C_2H_6	0·80	1·78	0·85	1·34	0·24
C_3H_6	0·10	0·22	0·13	0·18	0·03
C_3H_8	0·02	0·09	0·04	0·09	0·00
O_2	6·20	16·74	16·61	14·69	3·12
N_2	104·93	267·68	267·68	267·68	208·51
CO_2	10·95	37·28	22·99	23·48	12·08
Total	155·21	368·93	333·39	336·20	239·24

(c)

Bed depth = 5 cm, Temperature range = 801 − 900 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
CO	14·86	7·77	4·35	4·90	3·71
H_2	2·01	1·17	0·63	0·68	0·50
CH_4	1·47	1·01	0·59	0·78	0·17
C_2H_2	0·21	0·16	0·06	0·08	0·02
C_2H_4 / C_2H_6	0·46	0·40	0·19	0·30	0·07
C_3H_6	0·06	0·05	0·03	0·04	0·01
C_3H_8	0·01	0·02	0·01	0·02	0·00
O_2	3·57	3·75	3·72	3·29	0·90
N_2	60·44	59·96	59·96	59·96	60·05
CO_2	6·31	8·35	5·15	5·26	3·48
Total	89·40	82·64	74·68	75·31	68·90

(d)

Bed depth = 5 cm, Temperature range = 801 − 900 K

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
Prod. gas flow					
Dry (Nm³/min)	0·0894	0·0826	0·0747	0·0753	0·0689
Wet (Nm³/min)	0·0973	0·0857	0·0778	0·0784	0·0728
Per Kg fuel					
Dry (Nm³/Kg)	1·552	3·689	3·334	3·362	2·392
Wet (Nm³/Kg)	1·689	3·826	3·471	3·499	2·529
Steam content of wet gas (%)	8·1	3·6	3·9	3·9	5·4
Carry-over (g/min)	0·300	0·140	0·140	0·290	0·180
Bed-char + ash (g/min)	0·230	0·090	0·090	0·155	0·115

(e)

Bed depth = 5 cm, Temperature range = 801 − 900 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	47·0	16·9	59·3	100·0	28·2	39·9
RUN 2	92·1	30·4	76·0	100·0	43·8	62·9
RUN 3	53·3	16·3	52·5	100·0	23·7	40·4
RUN 4	59·2	21·5	52·3	100·0	29·4	47·4
RUN 5	29·3	5·4	27·7	100·0	11·7	23·8

TABLE 15

(a)

Bed depth = 5 cm
Temperature range = 901 – 1060 K

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
CO	17·06	22·51	6·45	8·45	10·53
H_2	2·52	3·49	1·12	1·49	3·21
CH_4	1·79	2·50	1·06	1·42	2·21
C_2H_2	0·30	0·27	0·12	0·19	0·18
C_2H_4 C_2H_6	0·65	0·77	0·48	0·61	0·94
C_3H_6	0·08	0·11	0·09	0·09	0·19
C_3H_8	0·01	0·02	0·01	0·01	0·02
O_2	2·02	3·48	4·32	3·07	2·99
N_2	66·45	62·03	72·69	70·37	66·27
CO_2	9·12	4·82	13·66	14·30	13·46
CV MJ/Nm³(dry)					
Gross	3·90	5·10	1·87	2·45	3·56
Net	3·74	4·88	1·78	2·32	3·35
Fuel feedrate Kg/min	0·0448	0·0576	0·0288	0·0368	0·0368
	2·8 × S	3·6 × S	1·8 × S	2·3 × S	2·3 × S
Temp. K	1034	1053	906	993	1015
Run duration mins	30	40	30	50	50

(b)

Bed depth = 5 cm, Temperature range = 900 – 1060 K

Nm³/Kg × 10²	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
CO	34·53	38·07	18·51	19·62	25·98
H_2	5·11	5·90	3·23	3·45	7·91
CH_4	3·62	4·24	3·06	3·29	5·46
C_2H_2	0·60	0·45	0·35	0·43	0·43
C_2H_4 C_2H_6	1·32	1·30	1·39	1·41	2·31
C_3H_6	0·16	0·19	0·24	0·22	0·46
C_3H_8	0·02	0·03	0·03	0·03	0·05
O_2	4·08	5·88	12·40	7·12	7·36
N_2	134·53	104·93	208·51	163·48	163·48
CO_2	18·46	8·16	39·17	33·23	33·21
Total	202·46	169·17	286·84	232·31	246·68

(c)

Bed depth = 5 cm, Temperature range = 900 – 1060 K

Nm³/min × 10⁻³	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
CO	15·47	21·93	5·33	7·22	9·56
H_2	2·29	3·40	0·93	1·27	2·91
CH_4	1·62	2·44	0·88	1·21	2·01
C_2H_2	0·27	0·26	0·10	0·16	0·16
C_2H_4 C_2H_6	0·59	0·75	0·40	0·52	0·85
C_3H_6	0·07	0·11	0·07	0·08	0·17
C_3H_8	0·01	0·02	0·01	0·01	0·02
O_2	1·83	3·39	3·57	2·62	2·71
N_2	60·27	60·44	60·05	60·16	60·16
CO_2	8·27	4·70	11·28	12·23	12·22
Total	90·70	97·44	82·61	85·49	90·78

(d)

Bed depth = 5 cm, Temperature range = 901 – 1060 K

	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5
Prod. gas flow					
Dry (Nm³/min)	0·0907	0·0974	0·0826	0·0855	0·0908
Wet (Nm³/min)	0·0968	0·1053	0·0866	0·0905	0·0958
Per Kg fuel					
Dry (Nm³/Kg)	2·024	1·692	2·868	2·323	2·467
Wet (Nm³/Kg)	2·161	1·829	3·005	2·460	2·604
Steam content of wet gas (%)	6·3	7·5	4·6	5·6	5·3
Carry-over (g/min)	0·240	0·300	0·180	0·260	0·260
Bed char + ash (g/min)	0·179	0·230	0·115	0·147	0·147

(e)

Bed depth = 5 cm, Temperature range = 901 – 1060 K

	MASS BALANCE EFFICIENCIES %				GASIFICATION EFFICIENCY %	THERMAL EFFICIENCY %
	Carbon	Hydrogen	Oxygen	Nitrogen		
RUN 1	68·2	25·0	68·0	100·0	39·6	57·0
RUN 2	61·1	27·9	65·3	100·0	43·2	60·0
RUN 3	72·7	20·8	77·8	100·0	26·7	44·3
RUN 4	67·7	22·0	75·8	100·0	28·2	45·6
RUN 5	80·2	39·7	80·9	100·0	43·3	63·9

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