THE DESIGN, EVALUATION AND COSTING OF BIOMASS GASIFIERS

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Doctor of Philosophy

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September 1988

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SUMMARY

The objective of this study has been to enable a greater understanding of the biomass gasification process through the development and use of process and economic models.

A new theoretical equilibrium model of gasification is described, using the operating condition called the adiabatic carbon boundary. This represents an ideal gasifier working at the point where the carbon in the feedstock is completely gasified. The model can be used as a "target" against which the results of real gasifiers can be compared, but it does not simulate the results of real gasifers. A second model has been developed which uses a stagewise approach in order to model fluid bed gasification, and its results have indicated that pyrolysis and the reactions of pyrolysis products play an important part in fluid bed gasifiers. Both models have been used in sensitivity analyses: the biomass moisture content and gasifying agent composition were found to have the largest effects on performance, whilst pressure and heat loss had lesser effects.

Correlations have been produced to estimate the total installed capital cost of gasification systems and have been used in an economic model of gasification. This has been used in a sensitivity analysis to determine the factors which most affect the profitability of gasification. The most important influences on gasifier profitability have been found to be feedstock cost, product selling price and throughput. Given the economic conditions of late 1985, refuse gasification for the production of producer gas was found to be viable at throughputs of about 2.5 tonnes/h dry basis and above, in the metropolitan counties of the United Kingdom. At this throughput and above, the largest element of product gas cost is the feedstock cost, the cost element which is most variable.

Key words: Biomass gasification, equilibrium model, stagewise model, capital costs, production costs.

Acknowledgements

Firstly, I would like to thank my supervisor, Dr Eric Smith, for his invaluable help with this work. I am also grateful to Dr Tony Bridgwater, who gave advice on some aspects of the work, particularly on gasifier economics.

I would like to thank the Heads of the Chemical Engineering Department during my time at Aston, Professor GV Jeffreys and Dr EL Smith, for making available the facilities for carrying out the work.

I am grateful for the financial support of the Science and Engineering Research Council.

Finally, I thank my family and friends for their support, especially my fiancée, Miss Michelle Parish, whose encouragement enabled me to complete this thesis.

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CHAPTER 1
INTRODUCTION

Gasification is a process for converting carbonaceous feedstocks to a gas which may be used either directly as a fuel or as a feedstock for processing into liquid fuels and chemicals. In its widest sense, gasification includes both true gasification processes, carried out with the aid of an oxidising (or gasifying) agent, and pyrolysis processes based on the thermal breakdown of the feedstock without the aid of oxidising agents and optimised to give a gaseous product. Gasification is part of the field known as thermochemical processing, which includes the related topics of carbonisation in which the pyrolysis process is optimised to produce a solid, pyrolysis in which the process is optimised to produce liquid products and direct liquefaction.

The first commercial process for converting a solid fuel into a gaseous fuel was set up in London in 1812, and produced a gas for street lighting from coal by dry distillation - a pyrolysis process. The first true gasifier, using air as a gasifying agent, was constructed by Bischoff of Mägdesprung in 1839 with coal or peat as fuel. [1,2] Gasifiers were developed technically and enjoyed a heyday in the first half of the twentieth century. The gas was used for lighting, heating, and fuelling gas engines (a form of internal combustion engine). Coal was the most important feedstock, but peat and wood were also used. During the Second World War, about 700 000 vehicles had their engines fuelled by gas from small gasifiers, mostly using charcoal feedstock. [3] In the post-war years, cheap oil and natural gas led to a decline in the use of gasification, but interest was re-kindled by the Oil Crisis of 1973. The emergence of the "Green" lobby has encouraged work on renewable sources, and biomass gasification has been one process option of interest.

Since 1973, many research projects have been undertaken with the eventual aim of designing a commercial scale biomass gasifier. A list of commercial or near-commercial scale biomass gasifiers is given in Bridgwater. [4] Although there have been many gasifiers built, there are still gaps in the understanding of the whole process of gasification. The survey by Reed is notable for its coverage of all aspects of gasification. [5]

The author's study has its roots in previous work carried out at Aston University [6,7,8] and was concerned with the production of fuels and chemicals using the gasification of biomass followed by synthesis of products from the gas produced. This study was based on computer modelling to estimate the process performance and economics of systems using fixed bed gasifiers. The emphasis was on the understanding of the parameters which influence the gasification process, and this theme has been continued in the author's work.

The objective of the author's study has been to enable a greater understanding of the gasification process through the development and use of computer models. Greater understanding of the gasification process will aid the design of gasifiers in the future and the development of new processes of which gasification forms only a part. The models which have been developed have been centred on three areas which are important to the understanding of the gasification process - chemical equilibrium in the gasification process, modified equilibrium modelling of gasification to predict performance more accurately where it deviates from ideality, and the economics of the gasification process. Because previous work at Aston was concentrated on fixed bed gasifiers, [6,7,8] the modified equilibrium modelling has concentrated on fluid bed gasifiers, which are more suited to synthesis gas production at high throughputs than fixed bed types. [4]

Chapters 2, 3 and 4 of the thesis give the background to the work. Chapter 2 outlines the complex network of reactions which comprises the gasification process, including the true gasification reactions and the pyrolysis reactions which also take place in a biomass gasifier. Chapter 3 describes the types of gasifiers which have been used for biomass gasification. Chapter 4 is a survey of the various models of biomass gasification which have been developed by other workers in the field.

Chapter 5 describes a new equilibrium model of gasification, which uses the operating condition called the adiabatic carbon boundary - an operating condition which gives the highest thermal efficiency for an idealised gasifier. The main function of this model is to provide a "target" against which the results of real gasifiers can be compared; it is not intended to be a model of the performance of real gasifiers. It has been used in a sensitivity analysis to study the effects of various parameters such as gasifier pressure and feedstock composition, on the performance of an idealised gasifier. A computer program has been written using the adiabatic carbon boundary model and which also incorporates other equilibrium models of gasification.

One reason why gasifiers do not perform exactly as predicted by equilibrium models is that hydrocarbons derived from the products of pyrolysis bypass the gasification reactions to be incorporated in the product gas. Some work has been done by Shand and Bridgwater [6,7] to account for this in the case of fixed bed gasifiers using a stagewise modelling approach. This type of model conceptually divides the gasifier into a number of connected black box reactors, the connections depending on the geometry of the particular reactor being modelled: each of the black boxes can then be used to represent a group of reactions, typically drying, pyrolysis and char gasification. Fluidised bed gasifiers are increasingly being preferred over fixed bed types for a number of reasons, but no stagewise modelling of fluid bed gasification has been performed. Chapter 6 describes a computer program which uses the stagewise approach in order to model this type of gasifier. By considering various types of stagewise model, the processes taking place within a fluid bed gasifier can be more easily understood.

In addition to understanding the physics and chemistry of the gasification, it is important that the economics of the process be evaluated. Information on the capital costs of commercial or near-commercial gasifier systems was collected, and correlations have been produced to estimate the total installed capital cost. These have been used to construct an economic model of gasification, which is presented in Chapter 7. This has been used in a sensitivity analysis to determine the most important factors which influence the profitability of gasification.

The conclusions which can be drawn from the author's work are summarised in Chapter 8, and their implication in the design of fluid bed gasifiers described. This chapter also includes recommendations for future work which will aid the design process further.

CHAPTER 2
THE FEEDSTOCK AND GASIFICATION PROCESS

2.1 THE BIOMASS FEEDSTOCK

Gasification was first used as a process for converting coal to gas for lighting purposes. Throughout the heyday of gasification in the early part of this century, coal remained the predominant feedstock, but in areas where coal was expensive or not readily available other feedstocks, such as peat and wood, were used. [2,9,10]

Any combustible carbonaceous feedstock may be used for gasification, given certain constraints (eg particle size distribution). Wood was commonly used in gasifiers until their fall from favour after the Second World War. During the Second World War, 700 000 vehicles were fuelled by gas produced in vehicle-mounted gasifiers using wood or charcoal feedstocks. [4]

Recently, the possibility of using refuse as a feedstock for gasification processes has been considered. This has a substantial price advantage over other feedstocks, even though pre-processing treatment is required. As the bulk of refuse is ligno-cellulose material, it behaves similarly to wood in gasification processes. [11]

2.1.1 Composition

There are two ways of expressing the composition of gasification feedstocks - the mean composition, expressed as the proximate and ultimate analyses of the fuels and the full analysis of compounds present in the fuel.

Proximate and Ultimate Analyses

The traditional method for considering the suitability of a fuel for gasification is to look at the proximate and ultimate analyses of the fuel. [12] Methods of performing the proximate and ultimate analyses of feedstocks are detailed in Karr. [13]

The proximate analysis considers the fuel to consist of four fractions: moisture, combustible volatile matter, fixed carbon and ash. This is most useful when considering a fuel for combustion, as it indicates the proportions of combustion which will take place on and above the grate, the fixed carbon remaining within the bed of fuel during combustion and the volatile matter being burnt above the bed. In gasification the proximate analysis is also useful because it provides the ash and moisture content of the fuel.

The ultimate analysis of the fuel is much more important in gasification. The ultimate analysis is an elemental analysis of the fuel, usually expressed as mass percent on a dry basis. The elements considered are usually carbon, hydrogen, oxygen, nitrogen and sulphur, and ash content is also recorded. When refuse is being considered as a gasification feedstock, it could be useful to consider the chlorine content. Chlorine is present in negligible quantities in most forms of biomass but it is often not negligible in refuse. Chlorine compounds in the product gas will cause corrosion problems with some materials of construction and may create an air pollution problem. [14]

The ultimate analyses of a number of gasification feedstocks are shown in Table 2.1.

As the ultimate analysis is the elemental analysis of a gasification feedstock, it is needed whenever material balances are to be calculated. Material balances are used in gasifier design, in the technical evaluation of real gasifiers, and in gasifier modelling.

Chemical Composition of Gasification Feedstocks

Most biological materials which may be considered as gasification feedstocks are composed of a diverse range of chemical compounds and are therefore difficult to analyse fully. There are even greater difficulties with analysing municipal refuse and other solid wastes which may be considered as gasification feedstocks. Refuse consists of a range of different materials - paper and cardboard, plastics, waste animal material, vegetable matter and non-combustible materials such as glass, ash and ferrous and non-ferrous metals. Each of these fractions vary in composition and each fraction is present in different proportions in wastes from different areas of the country. The animal and vegetable matter will decompose during storage, changing the chemical composition. Because of these factors, it is impossible to express a full chemical analysis of refuse.

Wood is the material most often considered as a biomass gasification feedstock, and its chemical composition is known. [15] Woods from different trees differ, but all woods contain three components: ash, the inorganic fraction of the wood, which usually forms less than 1% of the wood; extractables, the material making up the living cells of the wood, about 4 to 20% of the wood; and cell wall components, which make up the bulk of the wood.

Table 2.1 Ultimate analyses of gasification feedstocks

Feedstock	Mass %, dry basis						Ref. &
	C	H	O	N	S	Ash	notes
Wood and derived materials							
Douglas Fir	52.3	6.3	40.5	0.1	0.0	0.8	15
Redwood	53.5	5.9	40.3	0.1	0.0	0.2	15
Beech	51.6	6.3	41.5	0.0	0.0	0.6	15
Hickory	49.7	6.5	43.1	0.0	0.0	0.7	15
Maple	50.6	6.0	41.7	0.3	0.0	1.4	15
Douglas Fir bark	56.2	5.9	36.7	0.0	0.0	1.2	15
Redwood waste	53.2	6.0	39.9	0.1	0.1	0.6	15
Alabama Oak waste	49.2	5.7	41.3	0.2	0.0	3.3	15
Charcoal	80.3	3.1	11.3	0.2	0.0	3.4	15
Paper	43.4	5.8	44.3	0.3	0.2	6.0	15
Wastes							
Rice hulls	38.5	5.7	39.8	0.5	0.0	15.5	15
Refuse	30.7	4.2	20.5	0.3	0.5	43.8	16 §
Municipal solid waste	33.5	4.6	22.4	0.7	0.4	38.4	17
Coals							
Ekenberg peat briquette	58.4	5.4	26.0	-	0.5	9.7	18
Malay lignite	68.4	5.8	17.8	-	0.4	7.6	18
Yorkshire bituminous	73.1	5.5	8.7	-	1.2	11.5	18
South Wales anthracite	87.5	3.3	3.5	-	0.7	5.0	18

Note

§ Ash includes metals etc.

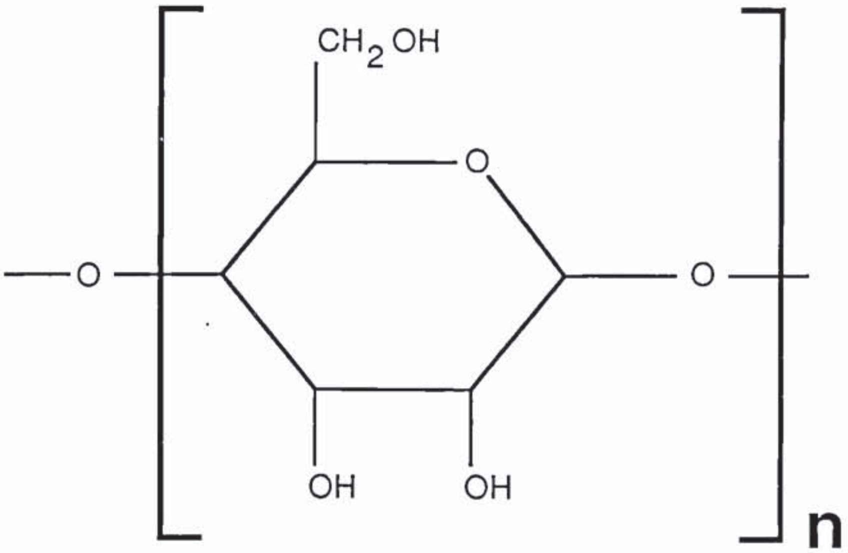


Figure 2.1 Structural Formula of Cellulose

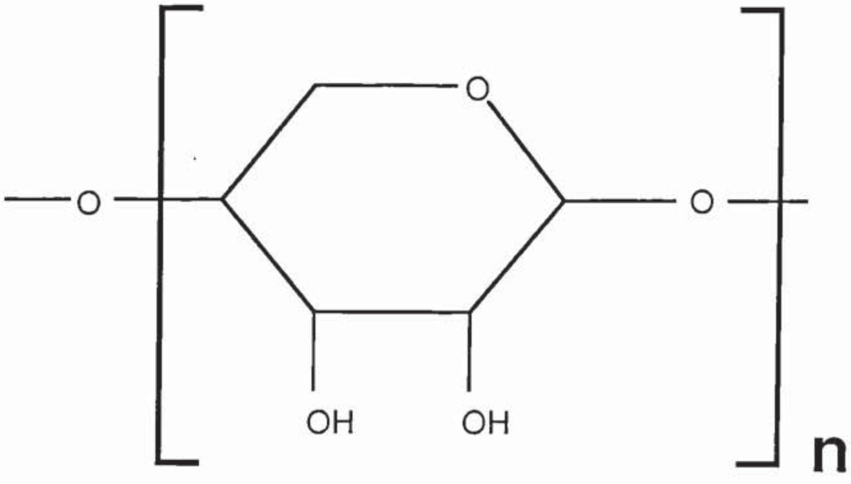
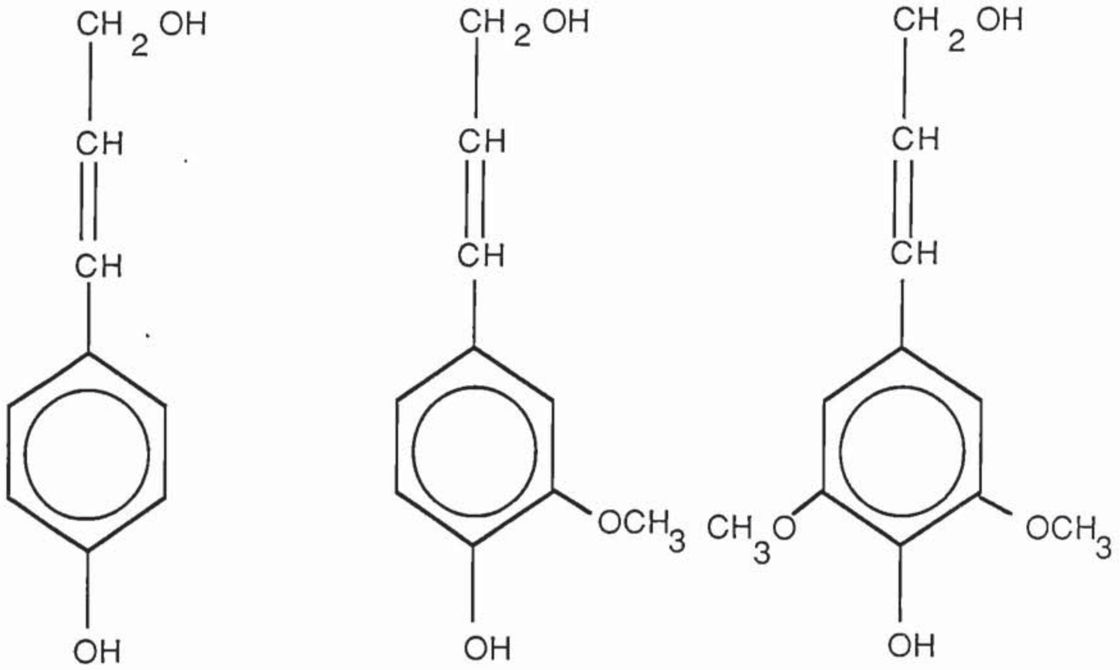


Figure 2.2 Structural Formula of Xylan Hemicellulose



p-coumeryl alcohol coniferyl alcohol sinapyl alcohol

Figure 2.3 Monomers of Coniferous Lignin

The cell wall components may be divided into two fractions, lignin, and the hollocellulose fraction consisting of polysaccarides. Lignin forms about 25 to 40% of the wood and hollocellulose 40 to 60%.

Hollocellulose forms the fibrous part of the cell wall, and may be divided into cellulose and hemicellulose.

Cellulose is a straight-chain polymer of d-glucose, (Figure 2.1) and forms 60 to 70% of the hollocellulose. It has the general formula

$C_6H_{10}O_5$ and in wood its molecular weight is usually over 100 000.

Hemicellulose is intertwined with cellulose in the cell wall fibres. It is a polymer of various sugars, the most common being d-xylose (Figure 2.2) and d-mannose. Unlike cellulose, it is usually in the form of branched chains of 50 to 200 units.

Lignin is an amorphous substance which surrounds the cellulose fibres, cementing them together. It is a three dimensional polymer of various aromatic units, which are mainly based on phenyl propane (Figure 2.3). Its structure is not certain, but a representative structure has been published by Grabowski and Bain. [15]

The chemistry of these wood components, as it affects gasification, is described in Section 2.2.2.

2.1.2 Other Properties of Biomass Influencing Gasification

Heating Value

As gasification is primarily an energy conversion process, the energy balance is of prime importance in assessing the performance of the process. The chemical energy of the feedstock is the dominant, if not the only energy input to the process, and hence has a large influence on the energy balance. The heating value of the gasification feedstock is therefore very important in the process of gasification.

There are two ways of expressing the heating value of fuels: the higher heating value and the lower heating value. The higher heating value is the heat evolved in combustion of the fuel to carbon dioxide gas and liquid water, whereas the lower heating value is the heat evolved when the products of combustion are gaseous carbon dioxide and water vapour. In combustion processes the latent heat of the produced water vapour is not usable, and so the lower heating value is a measure of the usable heat of combustion. However, in thermodynamic calculations, the heat of reaction is given for the reactants and products in their standard state at 25 °C: the higher heating value is thus the thermodynamically correct measure of the heat of combustion. The higher heating value is also the heat which is measured in a bomb calorimeter. In the author's work, the higher heating value has been used for both feedstocks and products.

The calorific value of gasification feedstocks should always be expressed on a dry basis, as this involves less risk of confusion. Dry basis calorific value can be adjusted easily to account for the actual moisture content of the feedstock in energy balance calculations. If the natural basis (ie wet basis) calorific value is quoted then the moisture content must be stated as well. The moisture content of many forms of biomass is not constant (see below); consequently, the quoted wet basis calorific value must be adjusted to the actual moisture content, giving the wet basis no advantage over the dry basis.

Several equations exist for calculating the dry basis calorific value of gasification feedstocks, using the ultimate analysis. Grabowski and Bain [15] tested three equations for calculating higher heating value, and found the IGT equation to be the most accurate for biomass and biomass derived materials:

$$HHV = 146.5C + 568.78H + 29.45 - 6.58A - 51.53\,(O + N) \qquad [2.1]$$

where:

HHV = dry basis higher heating value, BTU/lb
(To convert from BTU/lb to MJ/kg multiply by 0.002326)
A = mass percent ash, dry basis
C = mass percent carbon, dry basis
H = mass percent hydrogen, dry basis
N = mass percent nitrogen, dry basis
O = mass percent oxygen, dry basis

The IGT equation has been used in this work whenever a feedstock ultimate analysis was available but no dry basis heating value given.

Moisture Content

The moisture content of the feedstock affects both the mass balance and the energy balance of the gasification process (see Section 2.2.3). The biomass moisture content is therefore an important parameter in gasification and must be clearly stated when presenting results of modelling or experiments on real gasifiers.

The moisture content of biomass is usually expressed in one of two ways - both are given as mass percentages, but in one case the moisture is expressed as a proportion of the dry mass of the biomass and in the other case the moisture is expressed as a fraction of the total wet mass. Confusing these two bases leads to large errors - 50% wet basis is equivalent to 100% dry basis. As ultimate analyses and heating values are expressed on a dry basis in this work, moisture contents are also expressed on a dry basis.

Moisture is stored within the biomass in two ways - either as liquid within the cells, or bound into the cell wall structure. When biomass is left to dry in air, the free water within the cells is lost, followed by some of the moisture from the cell walls. Some of the cell wall moisture remains, as there is an equilibrium moisture content of biomass, dependent on the relative humidity and temperature of the atmosphere. Figure 2.4 shows the equilibrium moisture content of wood. Because of this equilibrium moisture content, "air dry" wood may in fact contain 10 to 20% water.

Biomass Physical Structure

Most biomass materials are fibrous in structure, and hence are anisotropic. This could give rise to both process and mechanical considerations when designing gasification processes. To illustrate the effect of anisotropic behaviour on the process, consider a wood particle undergoing gasification: diffusion of gasifying agent and product gases takes place along the grain, as diffusion across the cell walls perpendicular to the grain is negligible. In addition, the thermal conductivity across the grain may be different from that along the grain. However, these process considerations are minor when compared with the influence of anisotropy on gasifier mechanical design. Some types of feeders may be easily jammed by biomass, and knife valves to control biomass flow have to be specially designed. Size reduction can cause problems: wood tends to split along the grain, and the resulting particles are shaped like pins. Such particles can mat together in storage and feeding systems causing bridging problems.

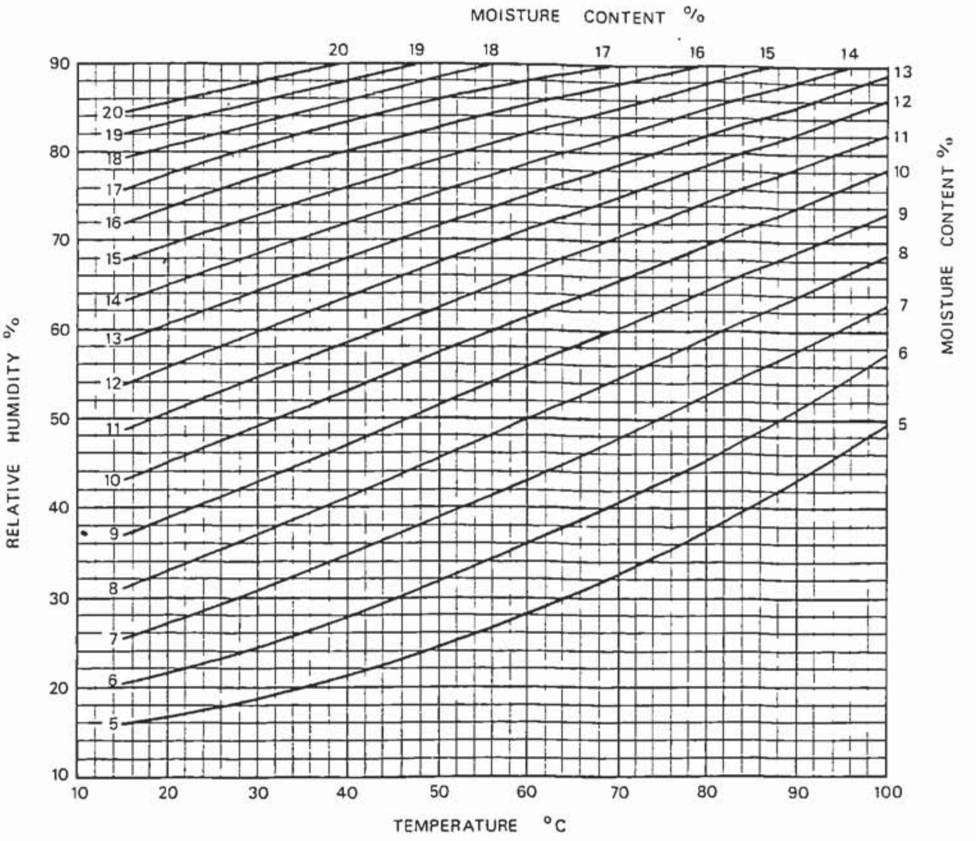


Figure 2.4 Equilibrium Moisture Content of Wood as a Function of Relative Humidity and Temperature

(taken from Timber Research and Development Association Wood Information, Section 4, Sheet 14. Note that moisture contents are on a wet basis)

2.2 PRINCIPLES OF GASIFICATION

A particle of biomass passing through a gasification reactor undergoes a complex series of physical changes and chemical reactions which may be grouped into four steps: drying, pyrolysis, gasification and tar cracking. These steps are not discrete processes and it is difficult to decide where pyrolysis ends and gasification begins. Also, because of the resistance of biomass materials to heat and mass transfer, the centre of a biomass particle may be in the drying stage while the surface is being gasified and an intermediate zone is undergoing pyrolysis.

2.2.1 Drying

Drying is the simplest step of the process in that it involves only a physical removal of moisture from the wood, with no chemical reactions taking place. However, it is of importance because of its effect on the heat balance.

Drying is an energy-consuming process, as the latent heat for the conversion of biomass moisture to water vapour must be supplied. This latent heat is no longer available as useful energy to pass into the outlet gas stream, and hence a higher feedstock moisture content will reduce the thermal efficiency of the gasifier.

2.2.2 Pyrolysis

Pyrolysis is the term given to the thermal decomposition of biomass and may take place in the absence of external reagents. Pyrolysis is not a single reaction but a complex network of series and parallel reactions which can only be understood for the simplest of biomass materials. A number of reviews of pyrolysis are available, including those by Antal and Milne. [19, 20, 21]

The pyrolysis behaviour of biomass materials is different as is that of the three main components of wood - cellulose, hemicellulose and lignin.

To describe the pyrolysis step, it is usual to consider it as a single reaction with three products: char, tar and gases.

$$\text{biomass} \xrightarrow{\text{heat}} \text{char} + \text{tar} + \text{gases} \qquad [2.2]$$

Char

Char is the solid product of pyrolysis. It is usually about 30% of the mass of the original biomass, but it contains about 50% of the chemical energy of the biomass. It is a black solid which retains the structure of the original biomass. Its physical strength, however, is much less than that of the original biomass, and it tends to be friable.

Chemical analysis of the char shows it to be mainly carbon, but it retains some of the hydrogen and oxygen of the original biomass (see entry for charcoal in Table 2.1).

Charcoal is used as a clean burning fuel for cooking in many developing countries and has a long history as both a general fuel and for metallurgical purposes. However, it is not an ideal fuel because of its limited mechanical strength. In gasification, the aim is to reduce char production to a minimum and maximise the production of fuel gas.

Tar

Tar is a term used to describe the condensable products of pyrolysis, excluding water. Tar is a viscous, acidic liquid, usually of a brown or black colour. Many compounds are present in tar and over a hundred have been identified.[22]

Tar has been considered as a substitute for fuel oil but has been rejected because it is carcinogenic, unstable in storage, and generally unpleasant to handle. Tar produced in gasification must be removed from the gas if it is to be piped for even short distances from the gasifier, as it will condense in the pipes. Tar removal is usually performed by scrubbing the product gas with water. Treatment of the used wash water is difficult, as the phenolics and organic acids in the tar are detrimental to the bacteria used in aerobic water treatment processes. Because of these problems associated with tar, it is often useful to optimise gasification processes to avoid the production of tar.

Gases

Gases from pyrolysis include the oxides of carbon, hydrogen and low molecular weight hydrocarbons. In a gasification process the pyrolysis gases will mix with the products of char gasification.

Carbon monoxide, carbon dioxide and hydrogen are present both in pyrolysis gas and in the gas produced by char gasification, and no differentiation is possible between these gases from the two sources. However, very little hydrocarbon is produced by char gasification (see Section 2.2.3), and most of the methane and all of the higher hydrocarbons in gasification product gas come from the pyrolysis step.

Mechanisms

As stated above, pyrolysis involves a complex network of reactions, and the complete chemistry of the process is not understood. However, cellulose pyrolysis has been studied extensively and may be used as an example of the reaction pathways involved in pyrolysis.[19,23]

In the pyrolysis of cellulose, there are two primary reactions - dehydration and depolymerisation. Dehydration involves the evolution of water vapour and the oxides of carbon, leaving a char residue. Competing with it is depolymerisation, the dehydration being favoured at lower temperatures (200 to 280 °C) and depolymerisation taking over at higher temperatures. Depolymerisation takes place by the cleavage of the ether bonds between the glucose units, followed by re-arrangement to give a tar monomer, the most common being levoglucosan (see Figure 2.5).

Levoglucosan is only an intermediate; the final products of pyrolysis arise from secondary reactions in the gas phase. Levoglucosan may react in three ways to form:

i. char and gases, similar to the products of dehydration
ii. polymeric tars
iii. various condensable, volatile products

These secondary pyrolysis reactions are closely allied with the tar cracking described in Section 2.2.4.

The final pyrolysis product mix is largely dependent on these secondary reactions, which are themselves dependent on many factors, especially the atmosphere in which they take place. For example, in a vacuum, the primary pyrolysis products are removed as soon as they are evolved, and no time is available for secondary reactions; consequently, the pyrolysis products under an inert atmosphere are different from those of vacuum pyrolysis.

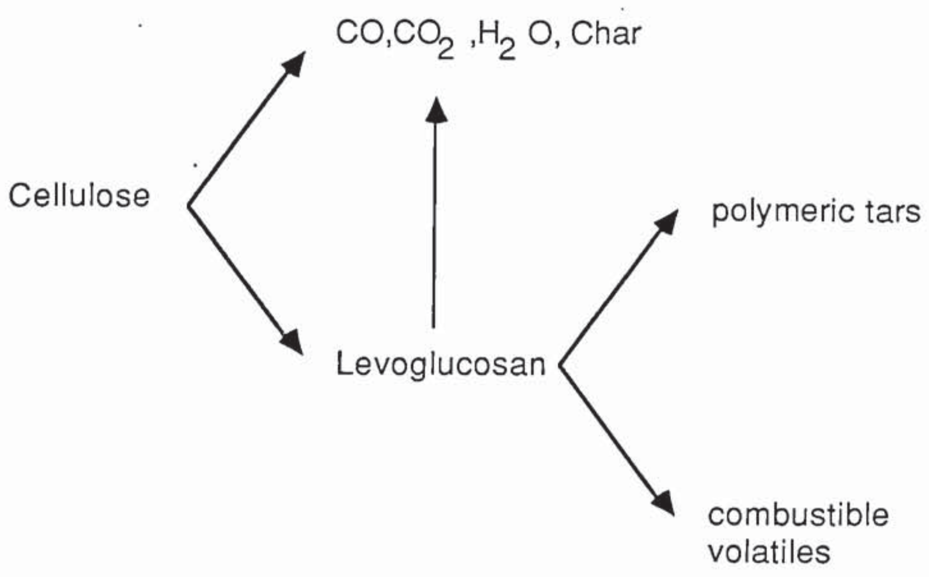


Figure 2.5 Reactions in the Pyrolysis of Cellulose

Hemicellulose pyrolysis is known to be analogous to cellulose pyrolysis in its reaction mechanisms, the intermediate levoglucosan being replaced by a furan derivative. Hemicellulose pyrolysis takes place at lower temperatures than cellulose pyrolysis.

Lignin is the most thermally stable component of woody biomass. Its pyrolysis mechanism is not clearly understood, partly because its structure before pyrolysis is complex and not well defined. Lignin is known to produce more char than hollocellulose in pyrolysis.

Whilst these mechanisms help with the understanding of pyrolysis, it must be remembered that the individual steps within the reaction schemes are actually generalisations and simplifications of the very complex processes actually taking place.

Factors influencing Pyrolysis

The principal chemical and physical factors which influence the pyrolysis reactions and the final distribution of pyrolysis products are shown in Figure 2.6. The factors which affect the product mix resulting from the pyrolysis of a small element of wood within a wood particle are shown. These factors are themselves dependent on the conditions external to the particle, and the size, shape and composition of the particle itself, together with the position of the element within the particle. These relationships are shown in Figure 2.6. As each small element of wood within a particle experiences a different set of conditions, the pyrolysis products from different parts of a wood particle will differ in both composition and proportions of char, tar and gases.

Because of these complex relationships and the complicated nature of the reactions involved, it is difficult to produce a general theoretical model of pyrolysis and models of pyrolysis must be confined to specific idealised cases or must be qualitative models. There is no general model to predict the relative proportions of the products with their compositions.

2.2.3 Char Gasification

Char gasification is the most fully understood of the sub-processes of gasification. Char, a substance which approximates to carbon, reacts with a gasifying agent to form gaseous products at temperatures of 700 to 1000 °C. The gasifying agent may be oxygen, air or steam, or a mixture of these. The reactions between oxygen and carbon are:

Combustion

$$C + O_2 \rightarrow CO_2 \qquad \Delta H = -393.5 \text{kJ/mol C} \qquad [2.3]$$

Boudouard reaction

$$C + CO_2 = 2CO \qquad \Delta H = +172.5\text{kJ/mol C} \qquad [2.4]$$

These may be combined to give:

$$2C + O_2 \rightarrow 2CO \qquad \Delta H = -221.0\text{kJ/mol } O_2 \qquad [2.5]$$

The reaction between oxygen and carbon to give carbon monoxide is exothermic and the energy released is used to heat the reactants to the temperature of reaction.

The reaction between carbon and steam is known as the water-gas reaction:

$$C + H_2O = CO + H_2 \qquad \Delta H = +131.3\text{kJ/mol} \qquad [2.6]$$

This reaction is endothermic and heat must be supplied for the reaction to proceed. There are two strategies which may be adopted to achieve this: one is to mix the steam with oxygen or air, the heat released from reaction [2.5] supplying the energy for the water gas reaction; the other is to transfer heat from a separate combustion process.

Another heterogeneous reaction which may take place forms methane:

$$C + 2H_2 = CH_4 \qquad \Delta H = -74.9\text{kJ/mol C} \qquad [2.7]$$

Studies of the thermodynamics of gasification, for example by Desrosiers [24] have shown that the only species present at equilibrium are H_2, CO, CO_2, CH_4, N_2 and H_2O. Other species, such as higher hydrocarbons and ammonia, are unstable, and only present in very small concentrations under gasification conditions.

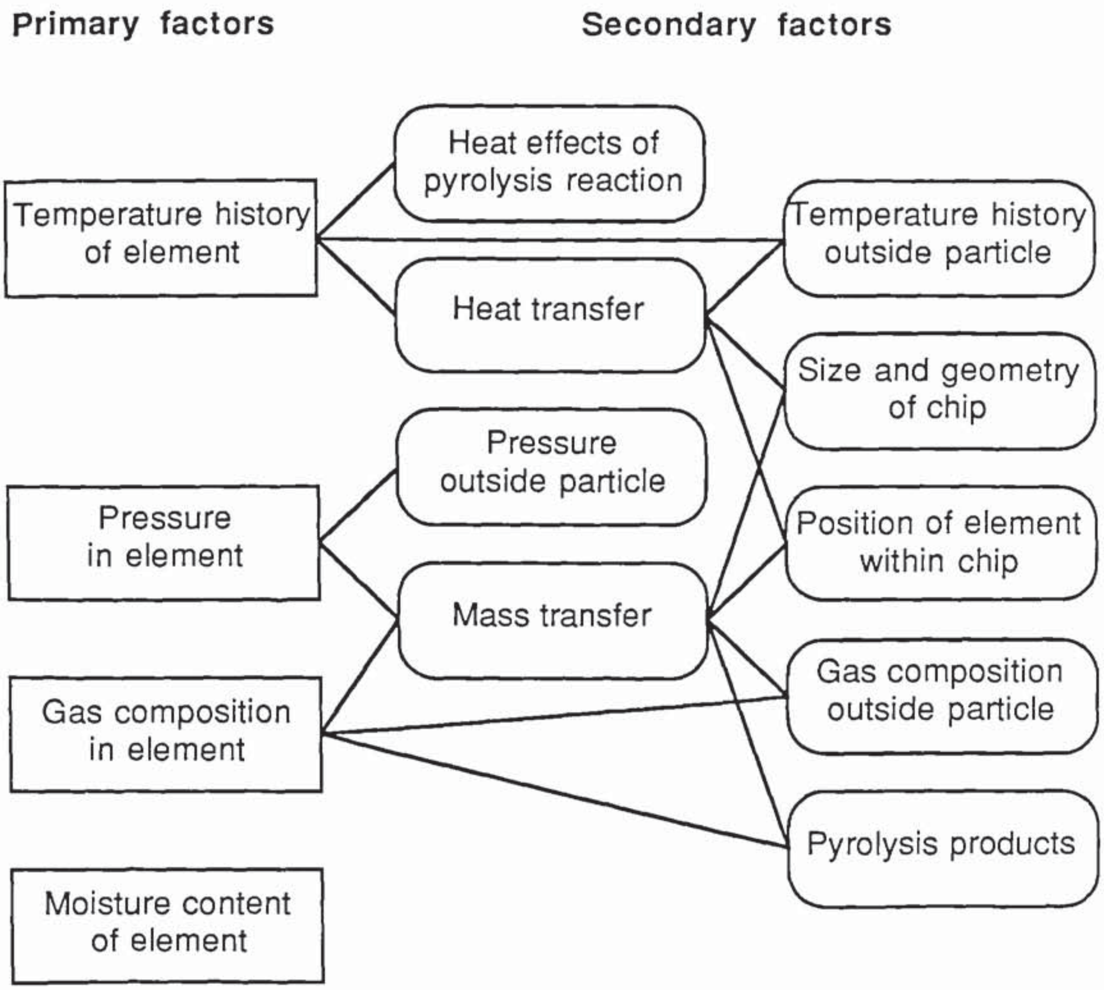


Figure 2.6 Factors Influencing the Pyrolysis of a Small Element Within a Wood Chip Showing Relationships Between Factors

In addition to the heterogeneous reactions, there are gas phase reactions between the species mentioned above, for instance the water-gas shift reaction:

$$H_2 + CO_2 = CO + H_2O \qquad \Delta H = +41.2 kJ/mol \qquad [2.8]$$

This is, in effect, a combination of the heterogeneous reactions [2.4] and [2.6]. Similar combinations of reactions may be constructed to account for the formation of methane in the gas phase.

2.2.4 Tar Cracking and Reactions of Pyrolysis Products

The products of pyrolysis include polymeric tars and volatile organic products including phenolics, acids and hydrocarbons. None of these compounds is thermodynamically stable under gasification conditions. Therefore, in a system where the pyrolysis products pass through the gasification zone, the pyrolysis products will react to produce simpler compounds. If equilibrium is attained, the only products will be the simple gases H_2, CO, CO_2, CH_4, N_2 and H_2O. However, in real gasifier systems, the pyrolysis products are not likely to attain equilibrium, and a range of other products will be formed.

The reactions of the pyrolysis products are in some ways analogous to the oil refinery operations of cracking and steam reforming. The large molecules are broken into smaller molecules either by the action of heat or by reaction with other species in the environment, such as oxygen and steam. However, in oil refining the feedstocks are usually straight chain hydrocarbons, whereas the tars cracked in a gasifier are aromatic and are oxygenated. The position in gasification is further complicated because the second stage of the pyrolysis process, which produces the compounds being cracked, takes place in the gas phase where the cracking and reforming also take place. The cracking reactions are thus strongly associated with secondary pyrolysis, and it is difficult to separate the two groups of reactions.

The cracking processes are not well understood, but it is known that increasing the temperature increases the extent of cracking. [25] Catalytic effects are also important, and dolomite and limestone are known to aid cracking. [26, 27] It is also known that pyrolysis products can be cracked almost completely by passing them through a hot char bed. [9]

2.3 PRODUCTS OF GASIFICATION IN PRACTICAL GASIFIERS

If gasification was taken to equilibrium the product gas would contain only H_2, CO, CO_2, CH_4, N_2 and H_2O. However, in real systems, pyrolysis products or compounds formed by the cracking of pyrolysis products may also be present. These include ethylene and higher hydrocarbons, condensable organics and tars.

The target gas composition depends on the use to which the gas is to be put: fuel gas must contain the maximum available energy for a given cost, whereas a synthesis gas must have hydrogen and carbon monoxide maximised while minimising methane and nitrogen. In most cases, the presence of tars and condensables in the exit gas is not desirable since they may be deposited in pipework or cause other problems: consequently, they must be removed from the product stream. It is, therefore, advantageous to reduce the tars in the reactor exit gas to a minimum. This can be done by using high reactor temperatures and by contacting the gas with either hot char or limestone catalyst, as discussed above.

The choice of gasifying agent has a large effect on the gas produced. The gas from air gasification contains about 50% nitrogen from the gasifying agent, and has a low heating value of about 5 MJ/m^3. The combustible components are mainly H_2, CO, CO_2 and CH_4.

Oxygen gasifiers produce a similar gas, but it is undiluted with nitrogen and has medium heating value of about 10 MJ/m^3. The gas from oxygen gasification is often suitable for use as a synthesis gas.

Gasification where steam is the only gasifying agent is uncommon because of the need to provide heat for the endothermic reactions. Since the temperature in such a gasifier is lower than for air or oxygen gasification the pyrolysis products are not cracked as extensively as in air or oxygen gasifiers. Thus, although the gas has a heating value similar to that of oxygen gasification, it has a higher hydrocarbon content and is less suitable as a synthesis gas.

The products of gasification also depend on the gasifier configuration and this topic is discussed in Chapter 3.

CHAPTER 3
GASIFICATION SYSTEM DESIGN

A gasifier system for converting biomass to usable gas needs four main components - a biomass feeder, a reactor, a gas cooler and a gas cleaner. Feedstock and product pass through the component parts of the system as shown in Figure 3.1; the gas cooling and gas cleaning may be in either order or combined. For some applications a hot, uncleaned gas is acceptable, and so the heat recovery and gas cleaning may be omitted.

In addition to these four essential parts, some pretreatment of the biomass feed may be required. Drying, size reduction, screening or classification, and sorting (for refuse feedstock) may be required to prepare the biomass for gasification. [28,29]

The design of a gasifier system is heavily dependent on the characteristics of the biomass feedstock, and the use to which the product gas is to be put. This is made clear in the following sections in which the four key process components are considered in more detail.

3.1 THE FEEDER

The function of a gasifier feeder is to transport the biomass feedstock from storage into the reactor while preventing the uncontrolled entry of air or the escape of product gas. Biomass materials behave differently from other bulk solids in handling systems, because of their fibrous nature, and non-uniform particle shapes, and the feeder for a biomass gasifier must be designed specifically for biomass. Using a

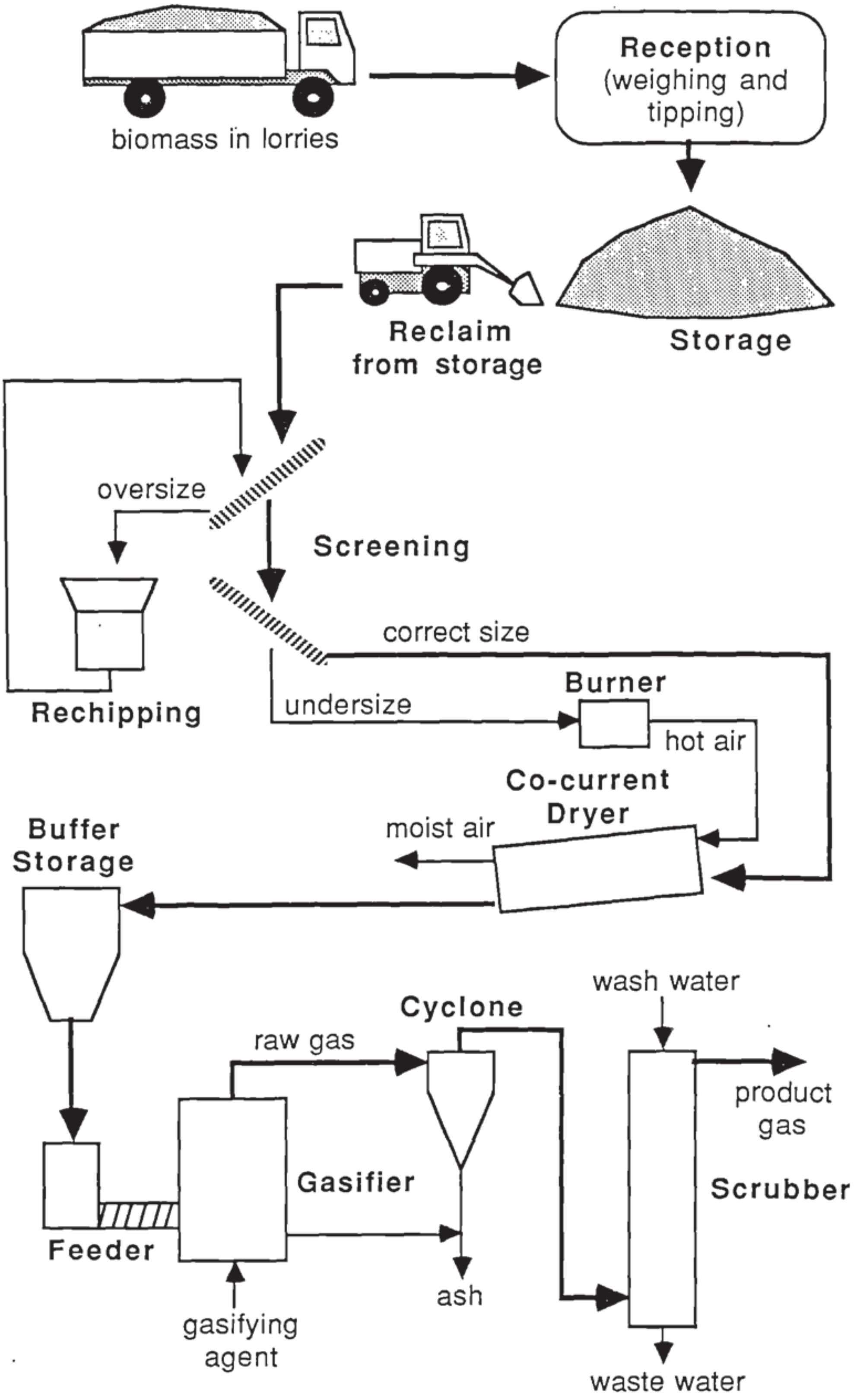


Figure 3.1 A Typical Gasification System

feeder designed to handle coal will cause major problems. The following discussion is based upon the review by Miles. [30]

A feeding system needs three parts: a system of transporting the biomass from the storage area to the gasifier, a lock-hopper or similar device for preventing the transport of gases in to or out of the reactor, and a device for delivering the biomass into the reactor.

Normally, use is made of a small hopper, immediately before the lock-hopper, to provide buffer storage in the feeding system. This can be fed with biomass from the main storage area in one of three ways: manually, using a front-end loader vehicle, or using a conveying system. Manual loading of biomass is most appropriate in the case of small systems, and in locations where labour is plentiful. Front-end loaders are used where storage is in piles on a paved area, whereas conveyor systems require specially designed storage bins or hoppers. This section of the feeding system is not troublesome compared with the lock-hopper and final delivery parts of the feeder.

There are two types of lock hopper-system commonly used in gasifier systems - conventional lock-hoppers equipped with knife valves top and bottom and rotary feeding valves. These components must be designed specifically for use in biomass gasification because of the handling problems associated with fibrous materials. The fibrous structure can cause jamming problems in both lock-hoppers and rotary valves; this is a particular problem with the knife valves of conventional lock-hoppers.

From the lock-hopper or rotary valve, there are three main ways of delivering the biomass into the reactor - by gravity, screw feeder or ram feeder. Gravity feeding is possible where it is acceptable to feed into the top of a reactor; this applies only to fixed bed reactors. Gravity feeding is the simplest method, but the lock-hopper or rotary valve is exposed to the high temperatures of the reactor and this can cause premature pyrolysis of feedstock. In addition, the mechanical parts of the lock hopper must be designed to withstand reactor temperatures.

Screw feeders and ram feeders may be used to feed into any part of a reactor and are thus more versatile. They must be designed to cope with reactor temperatures, and the residence time of the biomass in the feeder must be reduced to a minimum to obviate the possibility of pyrolysis. As with the lock-hopper, there is the possibility of jamming because of the special handling properties of biomass.

Of the three delivery methods, screw feeders are the most common because of their versatility, followed by gravity systems because of their simplicity.

3.2 THE REACTOR

3.2.1 The Reactor Function

The function of the reactor in a gasifier system is to contact the biomass feed with the gasifying agent and so allow reaction to take place. The design of the reactor depends on the feedstock to be used and the application of the product gas. The choice of gasifying agent also has a strong effect on the design of the reactor. For instance, special types of gasifiers are required to overcome the problems associated with steam gasification, see section 3.2.2. Also, the choice of reactor configuration will affect strongly the design of both the feeder and the downstream gas processing plant; therefore, the reactor must be the first component of the gasifier to be designed. [28]

The Influence of the Gas End Use

There are three basic types of product from gasifiers: [4]

i. Raw fuel gas

ii. Clean fuel gas

iii. Synthesis gas

In this context, "raw" gas is a dirty gas containing tar and particulates; it is used as a fuel in situations where the gasifier is close-coupled to the gas burner and where the cleanliness of the combustion products is not important. Raw gas is burned while still hot, thus using the sensible heat of the gas and improving the thermal efficiency of the whole conversion process.

"Clean" gas contains little or no suspended tars or particulates and is used either where clean combustion products are required or in gas turbines or gas engines for power production. Clean gas is usually cold because the methods of gas cleaning employed require that the gas be cooled (see Section 3.4).

Synthesis gas is a mixture of hydrogen and carbon monoxide and may be used in the production of a range of fuels and chemicals, including methanol, synthetic gasoline, and synthetic natural gas. Strict limits are usually placed on impurity levels in the synthesis gas. [31]

Each of the three classes of product requires different optimisation of the reactor system. In the case of both types of fuel gas, gas heating value and gasifier thermal efficiency should be maximised. This applies in the production of both low heating value gas by air gasification and medium heating value gas by oxygen/steam gasification. The difference between raw and clean gas lies in the tars produced: when producing a clean gas, tar production should be minimised to reduce the load on the clean-up system, whereas the presence of tars in a raw gas is not important as they are burned in a close-coupled system.

Synthesis gas should ideally be a mixture of only carbon monoxide and hydrogen. [32,33] Other gaseous components do not take part in the synthesis reactions and can build up to high concentrations in the synthesis recycle loop, requiring a large purge stream, and reducing the efficiency of the synthesis process. Carbon dioxide is not a problem in this context, as it is easily removed; but nitrogen is a problem and hydrocarbons must be converted by expensive reforming steps. Clearly, an oxygen or steam gasifier must be designed and operated so as to minimise the production of hydrocarbons. A clean gas is also required to prevent poisoning of the synthesis catalysts.

The Influence of Feedstock Properties

The properties of biomass feedstocks which are important in the process design of gasification have already been described (see Section 2.1). When designing a reactor the feedstock particle size distribution and mechanical properties are of particular importance. In a gasifier reactor, there is usually a bed of biomass and char, either fixed or fluidised, with a gas flowing though it. The size and shape of the particles and their packing behaviour will affect the structure and movement of a "fixed" bed of solids and will affect the fluidisation regime when the particles are used in a fluid bed. This has a number of consequences for the performance of the reactor. Firstly, the pressure drop across a fixed bed is dependent on the structure of the bed and so one containing many small particles will give an excessive pressure drop. Also, bridging may take place in the bed leading to the formation of voids and to the channelling and maldistribution of the gas. [2,4]

There are two possible solutions to these problems: either to use a fluidised or entrained bed design which is tolerant of small particles or to pretreat the feedstock by size reduction and screening processes. It may be necessary to adopt both of these solutions in combination. As well as adjusting the particle size distribution, it may be necessary or advantageous to reduce the moisture content of the feedstock by drying before feeding it to the reactor. The inclusion of pretreatment steps will increase the efficiency and reliability of the reactor but there is a trade-off between increased reactor efficiency and pretreatment costs.

Information on the feedstock limitations for each type of reactor is included in Section 3.2.2.

Ash Removal

After gasification, the ash or mineral content of the biomass feed is left behind as a residue. This can have important consequences for gasifier design, as the melting temperature of the ash is often in the same region as the gasifier reactor temperature. The ash from a gasifier may thus be removed in either the solid or the liquid state. Molten ash removal is simple, as the ash merely needs to be run off from the bottom of the gasifier and allowed to solidify. [1,10] Gasifiers using molten ash removal are known as "slagging gasifiers" and are not unknown in coal processing technology. [35]

The presence of molten ash is not without problems. In fixed beds, liquid ash formed in a hot zone can re-solidify in a cooler part of the reactor, cementing the bed together and severely impairing the performance of the gasifier. In a fluid bed, partially molten ash tends to agglomerate, forming large lumps which sink to the bottom of the bed, again affecting the performance. [36] Because of these potential problems, it is usual in biomass gasifiers to operate at a temperature below the ash fusion point. One practical method of controlling temperature is to inject steam with the gasifying agent, the endothermic reaction between the steam and char serving to reduce the gasifier temperature.

In fixed bed gasifiers, the ash is usually allowed to accumulate at the bottom of the gasifier, often by allowing it to fall through a grate. It can then be raked out manually or removed by mechanical means, for example, by screw feeder. [37,38,39]

There are two methods of ash removal employed in fluid bed gasifiers: either some of the bed material is continuously taken out of the reactor and the ash removed, for example, by screening, before returning the material to the bed, or the ash is elutriated from the bed and removed from the gas stream by one of the methods described in Section 3.3. [37,40,41]

3.2.2 Gasifier Types

There are a variety of reactor configurations which have been employed for the gasification of biomass, and these can be divided into two classes - fixed beds and fluidised beds. [37,42] There are three types of fixed bed gasifier: updraft or counter-current, downdraft or co-current, and crossdraft or crossflow. Fluid bed designs are not so easy to categorise, but they may be divided into simple fluid bed, fast fluid bed and entrained bed types. Gasification has also been carried out in equipment with mechanical movement or agitation of the biomass bed; examples are the Forest Fuels gasifier and the National Synfuels pyrolyser/gasifier. [43] These attempt to overcome flow problems encountered in packed bed gasifiers by using mechanical means to transport the solid material through the gasifier. The National Synfuels design employs a pyrolyser in the form of a heated screw feeder followed by an updraft gasifier to gasify the char produced, in an attempt to produce a cleaner gas. Because of their complexity, these gasifier types are expensive and are not in common use.

There are a number of less common gasification processes, which require special combinations of reactors. Two of the most promising are the twin fluid bed steam gasifier and the Oxygen Donor Gasifier (ODG). These produce a medium heating value gas without the use of oxygen by using a twin bed system. One bed is a combustor fired by air, which provides heat for the gasification reactions taking place in the other reactor. [44,45]

All of these gasifier reactors are designed for continuous operation.

The Downdraft Gasifier (see Figures 3.2 and 3.3)

The downdraft gasifier is a co-current gasifier; the biomass and gasifying agent pass down through the reactor and the product gas and ash are removed from the bottom of the reactor. [2,43,46] In operation, there is a hot zone in the reactor where drying, pyrolysis and gasification take place in close succession. Below the hot zone, further gasification takes place in a char bed. There are two types of downdraft gasifier - the conventional or throated downdraft and the open-core downdraft.

The conventional downdraft gasifier has a "throat" or constriction at the point where the gasifying agent is injected. This creates an intense hot zone with strong turbulence and mixing, where most of the pyrolysis products are cracked, thus producing a clean gas. [2]

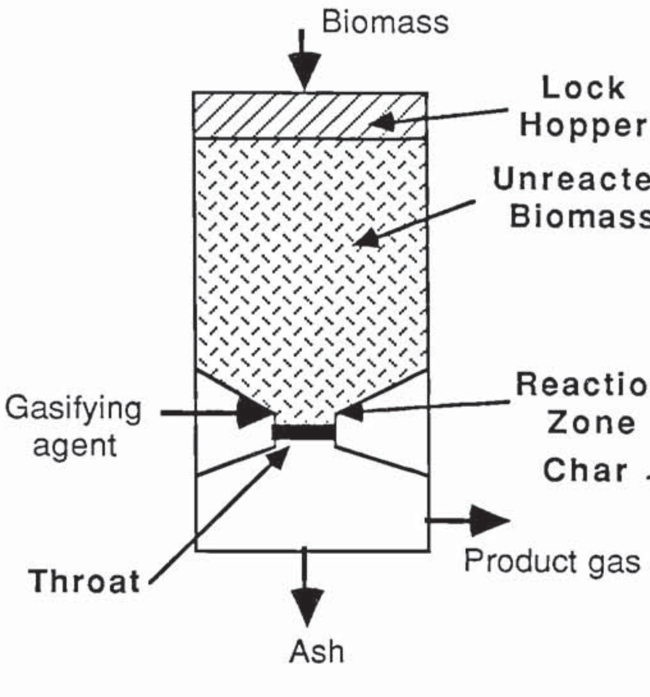


Figure 3.2
Throated Downdraft Gasifier

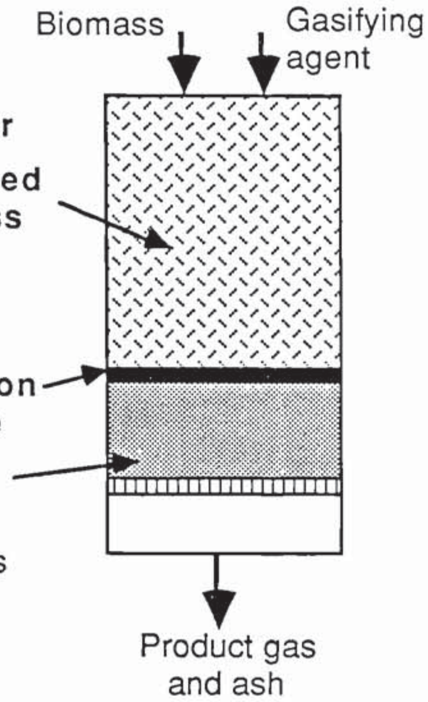


Figure 3.3
Open-Core Downdraft Gasifier

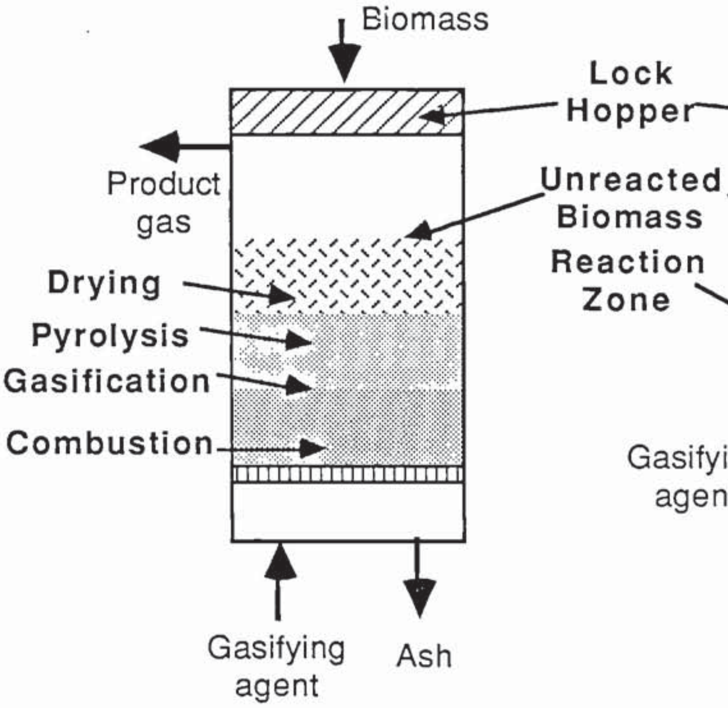


Figure 3.4
Updraft Gasifier

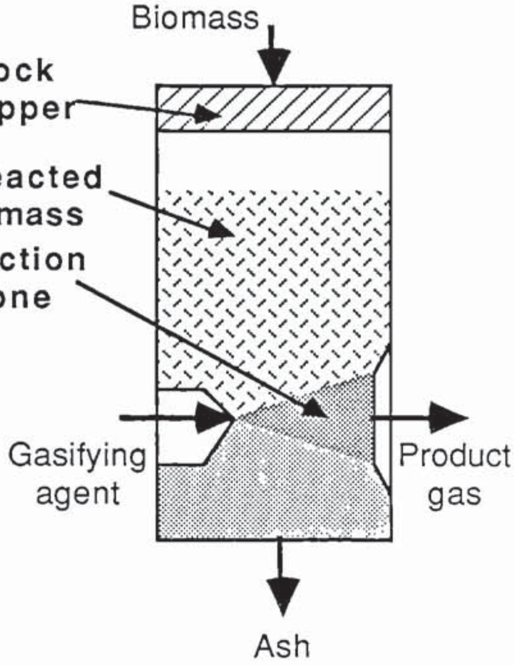


Figure 3.5
Crossflow Gasifier

The product gas is usually at a temperature of 600 to 800°C and so it is often passed through a jacket around the body of the reactor to recover energy by heating the feedstock. [9,38] There are a number of limitations on the conventional downdraft gasifier. Firstly to avoid a high pressure drop across the bed the biomass feed must be in the form of lumps. The biomass must also bridge the throat for correct operation of the reactor, and this constrains the biomass to a size of between 20 mm and 80 mm. Thus, fine particles of biomass must be pelletised before they can be used. The need for the biomass to bridge the throat of the gasifier also introduces a constraint on the diameter and hence the throughput,which is limited to about 0.75 t/h biomass. [4]

Open core downdraft gasifiers consists of a plain tube with a grate at the bottom, through which the gas is extracted. [46] Gasifying agent and biomass are both fed into the top of the tube; a lock-hopper system is not required although one may be used so that the flow of air into the gasifier can be controlled. There is no throat and so the hot zone is larger and cooler and its position is less well defined than in the conventional design. As the temperature is lower, there is less cracking of pyrolysis products. Also, as the zone is more diffuse and less stable, there is more opportunity for pyrolysis products to bypass the hot zone entirely. These two effects mean that the product gas is considerably more dirty than that from a conventional design. However, as the wood does not have to bridge across a throat, the particle size distribution is not as critical and the diameter of the gasifier is only constrained by the usual factors influencing maximum diameter of reactors.

Throated downdraft gasifiers were used extensively during World War Two to provide fuel for motor vehicles. The relatively clean gas requires little treatment before use in internal combustion engines and this made such gasifiers particularly suitable. Open core downdraft gasifiers have not been used on a commercial scale.

The Updraft Gasifier (see Figure 3.4)

The updraft gasifier is a counter-current design: biomass enters at the top and moves downwards while gasifying agent enters at the bottom; the product gas is removed from above the bed. [4,37,43] The gasifying agent first contacts char and combustion and gasification occur. The hot gas from this zone then passes through dry biomass, pyrolysing it to char, and picking up the volatile and gaseous pyrolysis products. In the upper section, the gas dries the wet biomass and some of the pyrolysis products are adsorbed by the biomass. Thus, the product gas contains most of the pyrolysis products and has a heavy load of tar. However, the gasifier may have a higher thermal efficiency than that of downdraft design, as heat is transferred from the product gas to the entering feed; this results in a lower gas exit temperature and smaller losses of energy as product gas sensible heat.

As with open-core downdraft gasifiers, updraft gasifiers are less influenced by particle size distribution than conventional downdraft designs, and there are no special limitations on throughput.

Several designs of updraft gasifier have been developed commercially. They have mainly been used for close-coupled applications where gas cleanliness is not important.

The Crossdraft Gasifier (see Figure 3.5)

The crossdraft gasifier is a much less common fixed bed arrangement. [3,4] The biomass enters at the top and passes down through the reactor, as with other fixed bed designs. The gasifying agent is injected into one side of the bed and the product gas is collected from the opposite side. There is a localised hot zone where the gasifying agent is injected; then, the hot, clean gas from this zone passes through the biomass causing pyrolysis and picking up the pyrolysis products. Therefore, the gas produced is fairly dirty.

There are currently no commercially available crossdraft biomass gasifiers but crossdraft charcoal gasifiers have been used to produce gas for motor vehicle fuel. [3,4] Using charcoal fuel avoids the introduction of pyrolysis products into the gas stream and a clean gas is produced as a result.

The Fluid Bed Gasifier (see Figure 3.6)

The fluid bed gasifier usually consists of a bed of sand or other inert material, fluidised with gasifying agent at a temperature of 700 to 900°C. [4,37,43] Biomass is fed into the bed and rapidly pyrolyses at this temperature, giving char which is then gasified. The bed is usually well-mixed and at an even temperature. The temperature of operation is lower than that of the hot zone of a downdraft reactor and so there is a lower conversion of pyrolysis products. Also, there may be some bypassing by the pyrolysis products in the bubbles within the fluid bed. Thus, the gas produced is dirtier than that from a downdraft gasifier, although it is much cleaner than that from an updraft gas producer. Because the gas velocities are higher than those in a fixed bed gasifier, there is a greater chance of small ash and char particles being entrained in the product gas; consequently, a cyclone separator is required to remove particulates.

The main advantage of fluidised bed gasifiers over fixed bed designs is that they do not require a lump feed; fine particles may be fed into the bed. However, large pieces of feedstock (greater than about 50mm diameter) must be reduced in size before they can be fed into the gasifier. A fluid bed gasifier often may be used with different feedstocks, for example, the Rheinbraun HTW gasifier has been tested with wood, peat, lignite, brown coal and bituminous coal feedstocks with good results. This characteristic makes them versatile in use, and they are especially suitable for a feedstock, such as domestic refuse, which can vary widely in composition from day to day.

Fluid bed gasifiers are fairly simple, both in construction and in operation. They are more complex than fixed-bed producers and require a more sophisticated control system.

The Fast Fluid Bed Gasifier (see Figure 3.7)

The fast fluid bed gasifier, also known as the circulating fluid bed gasifier, is a variant of the fluid bed type in which a higher gas velocity is used. [47] There is more elutriation of particles from the bed and there is no detectable interface between the bed and the freeboard. Elutriated particles are separated from the product gas in a cyclone and returned to the bed. The characteristics of fast fluid beds are similar to those of a simple fluid bed design; Lurgi claim that they have produced a tar-free gas from their circulating fluidised bed design.

The Entrained Bed Gasifier (see Figure 3.8)

In the entrained bed gasifier, the gas velocity is increased to a level where all the particles are entrained in the gas stream; there is no bed of particles present, just a homogeneous mixture of particles and gasifying agent. There is usually no inert carrier material as is the case in most fluid bed designs. [47]

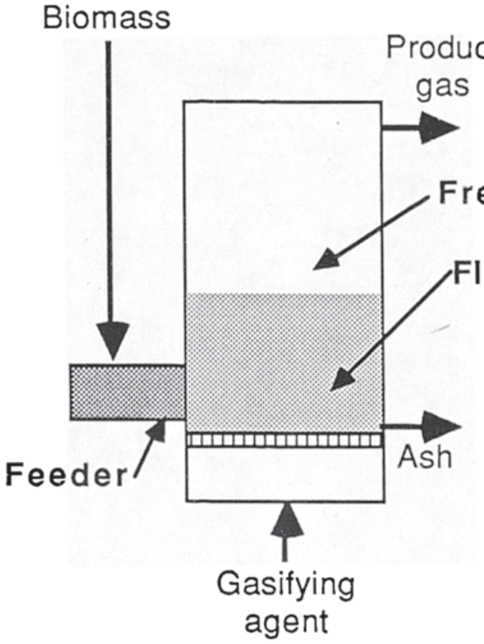


Figure 3.6
Fluid Bed Gasifier

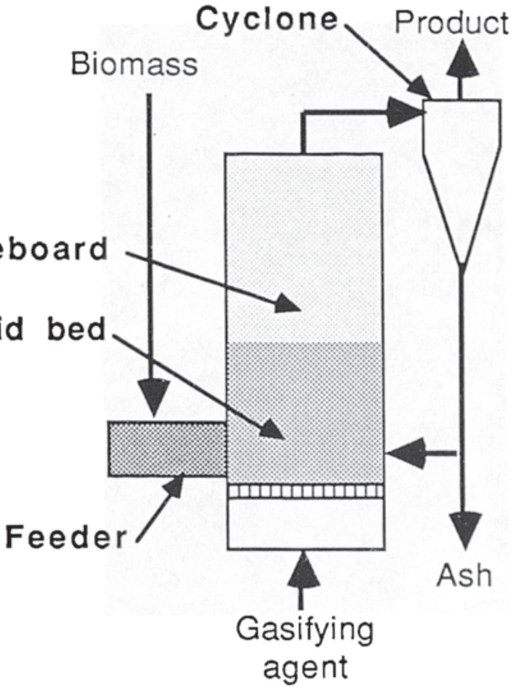


Figure 3.7
Fast Fluid Bed Gasifier

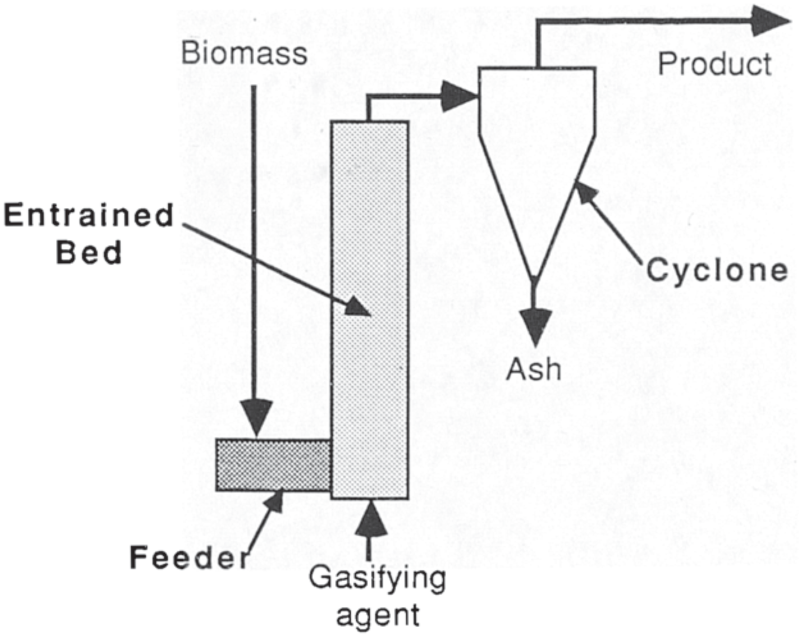


Figure 3.8
Entrained Bed Gasifier

The entrained bed gasifier represents a higher level of technology than fluid bed designs and in addition, the biomass must undergo expensive size reduction to produce particles small enough to be entrained. For these two reasons, entrained bed gasification systems have not passed the research stage. [4]

Steam Gasifiers (see Figure 3.9)

As mentioned in Section 2.2.3, the reaction between steam and char is endothermic. Consequently an external heat source is required for steam gasification. [4,45] There are two methods of supplying this heat to the gasification process - by indirect heat transfer or by direct heat transfer from a heat carrier.

The heat carrier method may be accomplished by using a twin fluid bed arrangement, using sand or other inert particulate material as the heat carrier. The gasification reactor is a fast fluid bed fluidised with steam, into which the biomass is fed. Sand and char are elutriated from the bed and removed from the product gas stream in a cyclone. They then pass to a combustor, where the char is burned in air, raising the temperature of the sand, which is returned to the gasifier. The heat carried in the sand is given up in the gasifier, thus sustaining the steam gasification reaction.

Steam gasification can also be performed using indirect heat transfer through a heat transfer wall. [40] In this case, there is a combustor section or combustion tubes in which biomass or some of the product gas is burned. Biomass is fed into the gasifier bed, which is again fluidised with steam, the heat for the gasification being transferred by conduction from the combustion section.

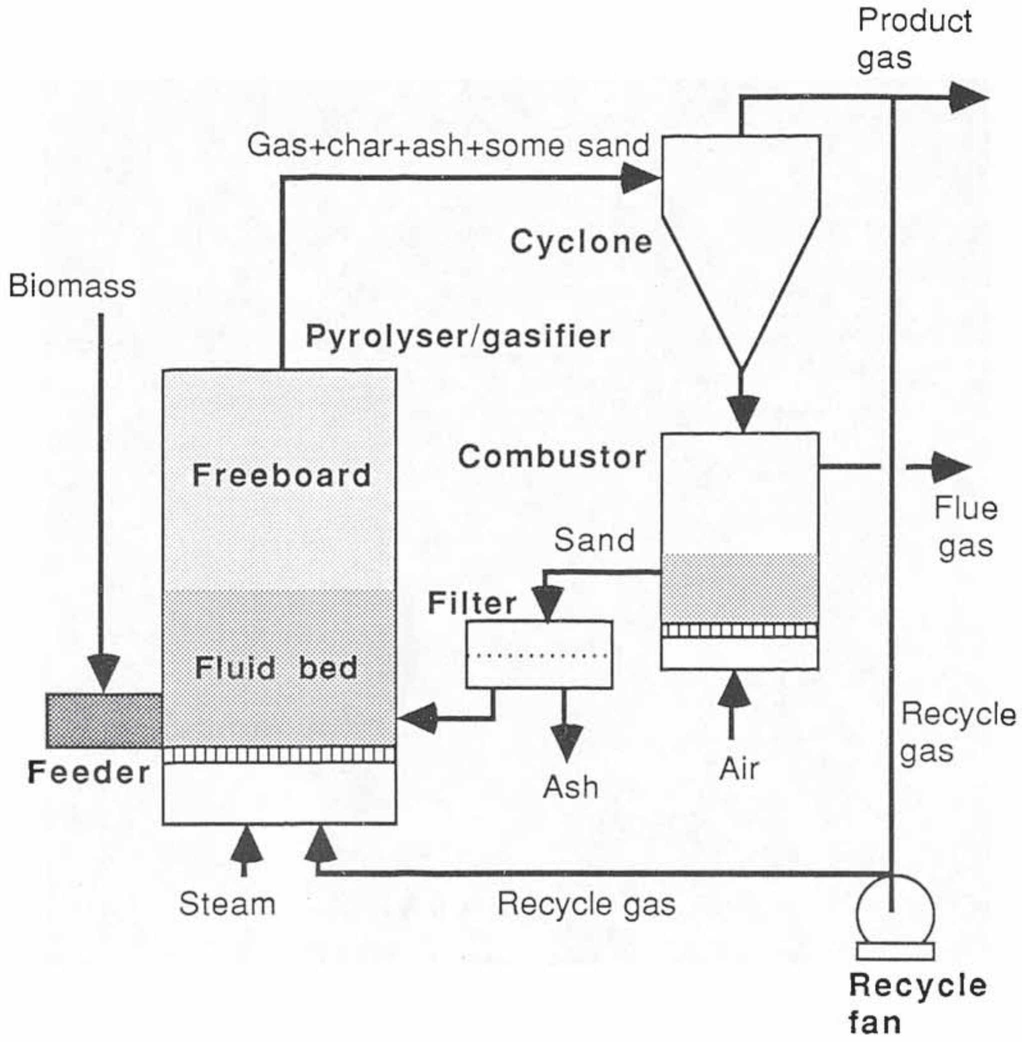


Figure 3.9 Twin Fluid Bed Steam Gasifier

The advantage of steam gasification over conventional processes is that pure oxygen is not required for the production of a medium heating value gas. There are three main disadvantages: the twin-bed systems are more complicated than conventional air or oxygen gasifiers; heat from the combustor flue gas must be recovered to make the system economic; and the low temperatures of the gasifier can cause formation of a dirty product gas.

The Oxygen Donor Process (see Figure 3.10)

There is one other gasification process which has been developed to produce a medium heating value gas without the use of pure oxygen. This process employs the Oxygen Donor Gasifier (ODG). [44] The ODG is a twin fluid bed design with a central gasifier and a surrounding combustor. Bed material is continuously transferred from the gasifier to the combustor and from the combustor to the gasifier through two specially designed "transfer slots". The bed material consists of a mixture of CaS and $CaSO_4$. In the combustor, CaS is oxidised to $CaSO_4$ with air at high temperature. In the gasifier, the $CaSO_4$ gives up its oxygen in the gasification process, reverting to CaS. The biomass is fed into the gasifier bed, where pyrolysis and gasification take place producing product gas. Char from the gasifier passes with the CaS/$CaSO_4$ to the combustor, where it is burned in air, providing heat for the oxidation of the CaS.

While this process has the advantage of needing no oxygen to produce a medium heating value gas, the gasifier system is complex compared with conventional designs and heat must be recovered from the combustor flue gas. This process has been developed as far as the pilot plant stage.

Other Two-Step Processes

Pyrolysis as a method for gas production is inefficient because of the energy wasted in the char byproduct; consequently, other processes have been developed which have pyrolysis as a first step, followed by separate gasification of the char. These systems are similar to steam gasifiers in their characteristics, and it may be that in some types of steam gasifier the gasification reactor is actually behaving as a pyrolyser. [43]

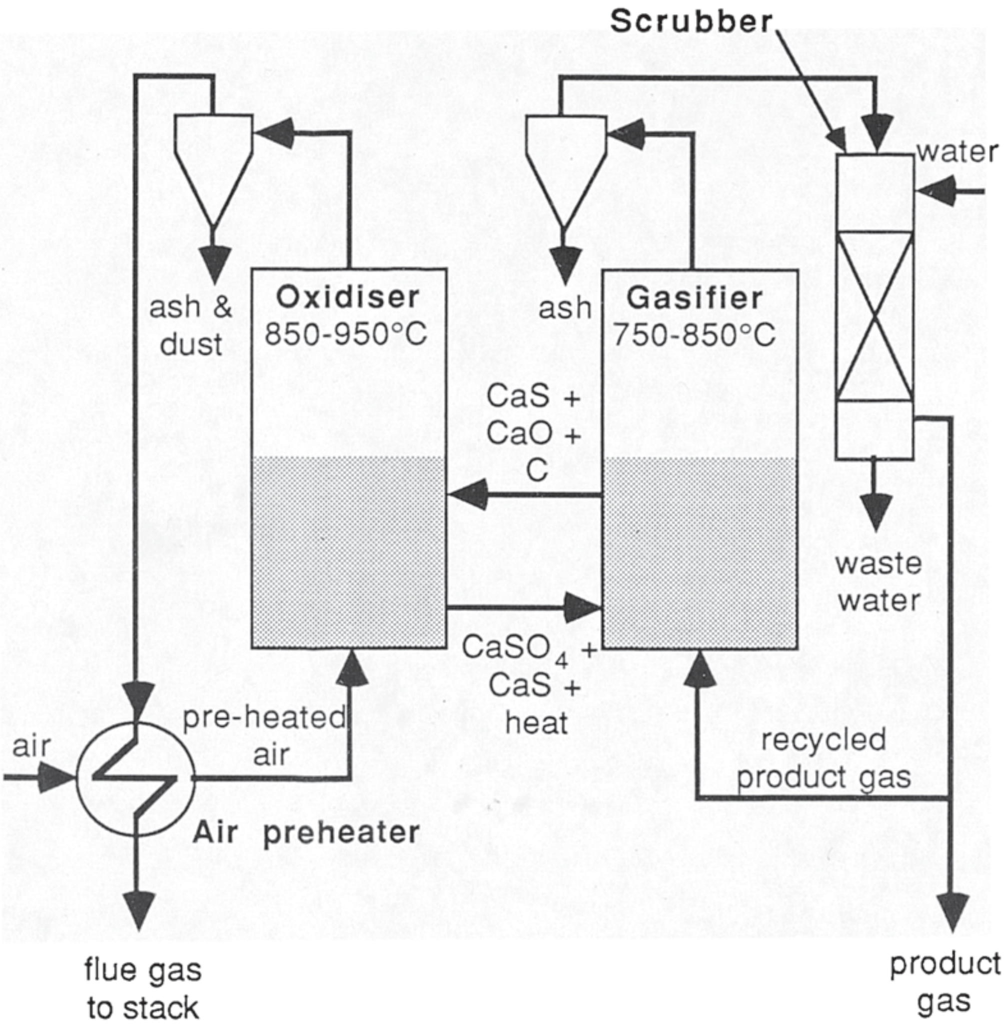


Figure 3.10 The Oxygen Donor Gasification Process

3.3 HEAT RECOVERY

Unless the product from a gasifier is to be burned while still hot, the gas must be cooled and the sensible heat recovered. At first sight, this would seem to be a simple heat exchanger design problem but pollutants in gasifier product gas can cause problems.[48] Even the clean gas from a throated downdraft gasifier contains some tars, and ash is present to a greater or lesser extent in all product gases. Tars will condense as the gas temperature is reduced, coating heat transfer and other surfaces and there polymerising in situ. Ash precipitated from the gas stream will also foul heat exchangers.[32] Heat exchangers and waste heat boilers installed to recover heat from product gas should be designed to minimise this fouling but even well designed systems will need regular cleaning. Unfortunately, gas cleaning technology has not yet reached the point where gases can be cleaned effectively while hot; consequently, gas cleaning cannot be performed before heat recovery to circumvent these problems.[48]

There are a number of uses to which the waste heat can be put, two of which can increase the efficiency of the reactor; the waste heat can be used either for drying and heating the biomass feedstock or for preheating the gasifying agent. The other main use for the waste heat is in the production of steam which can be employed as the gasifying agent in a steam gasifier or added to air or oxygen to moderate the reaction temperature. Alternatively, it may be possible to export steam to other plants in an area local to the gasifier. Low grade heat can also be used to generate hot water which might be exported for space heating purposes.

The design of the heat recovery system must be undertaken with a particular site in mind. For example, there may be no market for steam or space heating locally; additionally, the chosen feedstock may already be at the optimum moisture content. In such cases the options outlined above have to be ruled out.

3.4 GAS CLEANING

The gas produced in a gasifier contains contaminants, primarily tar and particulates. Tar is a generic term for all condensable organic material; it may contain both high molecular weight, aromatic hydrocarbons and lighter, soluble, oxygenated organics, such as phenols and acetic acid. The tars are often present as a fine mist with droplet sizes in the sub-micron range; high temperature gases may contain tars in the vapour state. The particulate fraction consists of char and inorganic ash and, in fluid bed gasifiers using an inert bed, small particles of bed material will also be present. The quantities of tar and particulates present depend on the type of gasifier (see Figure 3.11).[31]

The quantities of contaminants which can be accepted depends on the end use of the product gas. Where the gasifier is close-coupled to a combustion system, it may be possible to omit gas cleaning entirely or to include only a cyclone to remove particulates. [49] However, in many situations, gas cleaning is necessary. In a system where the product gas is to be piped to the end user, it is important to remove materials that could deposit in and eventually block the pipe.

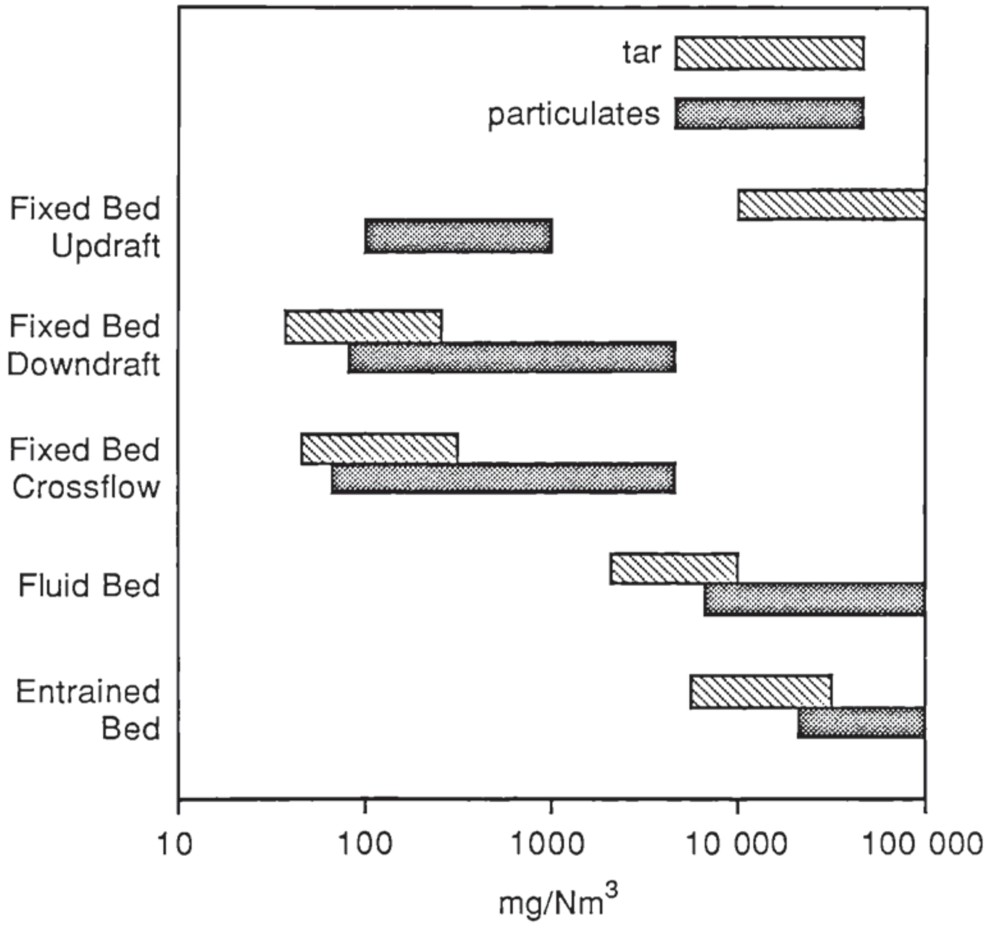


Figure 3.11 Tar and Particulate Loadings in Gasifier Product Gases

In the case of synthesis gas and fuel for gas turbines for power generation, contaminants must be within the range 1 to 20 mg/m^3. Gas engines can accept particulate concentrations in the range 10 to 50 mg/m^3; tars must be kept to a similar level to prevent blocking of the valves and inlet manifold. [50]

The gas cleaning equipment usually considered for gasifier applications includes [50,51]

Cyclones

Liquid scrubbers

Electro-static precipitators

Bag filters.

Cyclones are the cheapest form of gas cleaning device and are used to remove particulates. They have the advantage that they can be operated at high temperatures, but they are inefficient at removing particles of below 5 to 10 µm diameter. [52,53]

High efficiency cyclones have a higher collection efficiency for a given particle size, although they require a higher pressure drop and thus have a higher operating cost.

The next level of gas cleaning is represented by the liquid scrubber. This may use oil or water as the operating liquid, the use of water being much more common. [51] If oil scrubbing is chosen, the pyrolytic oils from the gasification process will probably be unsuitable because of their high viscosity; if this is the case, oil must be purchased and the used oil distilled before recycling. As water is readily available, it is possible to use a once-through scrubbing system, which is in effect a combined direct contact cooling and cleaning process. Effluent treatment would normally be a two stage process - separation of insoluble tars followed by conventional biological water treatment to reduce the biochemical oxygen demand (BOD). The disposal of insoluble tars may present a problem, although they could be incinerated. The main types of liquid scrubber available are plate and packed columns, spray scrubbers and venturi scrubbers. The most efficient of these is the venturi scrubber but it requires a secondary separator to remove entrained water droplets from the gas stream.

When the product gas is to be used either to drive turbines or for chemical synthesis, water scrubbing will not clean the gas to the required specification. In this case, bag filters or an electrostatic precipitator could be used. These devices are physically large compared with other clean-up systems and have a high capital cost. They are, therefore, used only where a very clean gas is required. They may operate at higher temperatures than water scrubbers - about 230°C is the maximum for bag filters and there is an upper limit of about 400°C for electrostatic precipitators. [50,52]

CHAPTER 4
INTRODUCTION TO GASIFIER MODELLING

4.1 INTRODUCTION

Mathematical modelling is used extensively in chemical engineering for problem solving. Models can be used in the following ways:

i. to predict the performance of a proposed system for design purposes; this may be referred to as design modelling.

ii. to predict the performance of an existing system under a set of conditions which have not been previously tested; this may be referred to as predictive modelling.

iii. to predict the performance of a system under a set of conditions which have been tested to obtain a better understanding of the physical and chemical processes taking place within a system; this may be termed investigative modelling.

Models may belong to more than one of these categories: for instance, a predictive model may also give a better understanding of the internal processes of a system.

A mathematical model must meet a number of criteria:

i. it must be as simple as possible, with as few assumptions and experimentally determined parameters as possible, whilst still meeting condition iii. below.

ii. it must be sufficiently simple for a result to be calculated using the methods at the modeller's disposal. Clearly if a solution cannot be calculated, the model is useless.

iii. it must give a reasonably accurate result to the problem being solved.

Many models of gasifier systems have been proposed that fall into all three of the categories above and meet the three criteria to a greater or lesser extent.

The models discussed by the author will all be ones dealing with process aspects of gasification, such as temperatures, flowrate compositions and residence times. Mechanical design aspects, such as flow patterns and fluidisation phenomena, are not discussed except where these aspects have been incorporated into a process model. The author's work has concentrated on models which predict the overall performance of gasifier systems, especially the product gas composition, and this review concentrates on these aspects of gasifier modelling. The author's work has also concentrated on fluid bed gasifier modelling, so the emphasis in this chapter has been on models of this type.

Models in chemical engineering may be divided into two types: steady state models and transient or time-varying models. Steady state modelling is used for many purposes including the construction of mass and energy balances during design. Transient modelling is used for predicting the behaviour of systems at start-up and shut-down and for designing control systems.

Steady state models of gasification may be divided into three main types:

i. Black box models, in which the gasifier is treated as a single reaction stage. This type of model is used for the prediction of the flowrates and compositions of the streams entering and leaving the gasifier. This type of model is described in Section 4.2.

ii. Stagewise models, in which the gasifier is divided into a number of linked black boxes or reaction stages. The results of this type of model might be only the flows and compositions of the outlet streams, or they might also include information about the internal workings of the reactor, depending on how the stages are defined.

iii. Differential models in which the reaction conditions vary throughout the reactor. These models make use of differential equations to describe the variation of the variables of interest with position within the reactor. Most kinetic models fall into this class.

4.2 BLACK BOX MODELLING

A biomass gasifier will usually have two inlet streams - the biomass and the gasifying agent - and two outlet streams - the product gas and a solid waste. These inlets and outlets are shown in Figure 4.1. Usually, the following information will be known:

- biomass composition
- biomass ash content
- biomass moisture content
- biomass inlet flowrate
- biomass inlet temperature
- gasifying agent composition
- gasifying agent temperature.

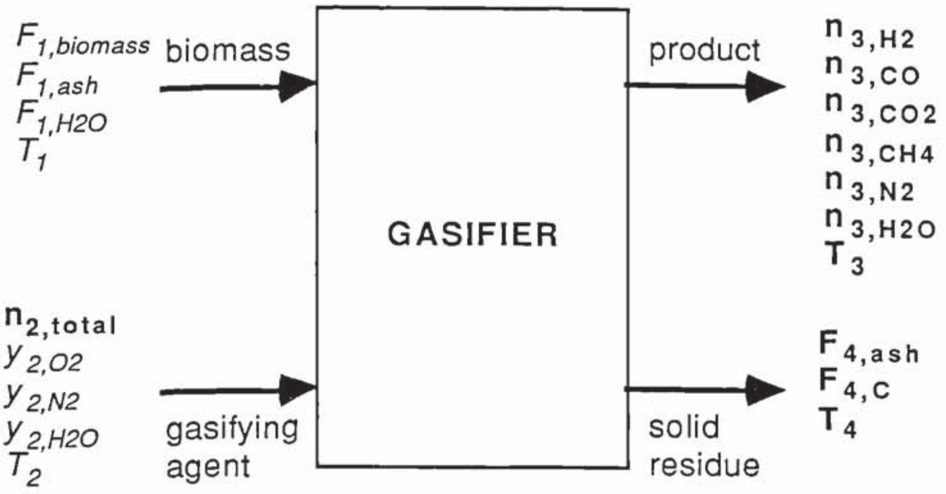


Figure 4.1 Inlet and Outlet Streams of a Biomass Gasifier

Table 4.1 Independent Equations and Unknown Variables in the Overall Mass and Energy Balances of a Biomass Gasifier

Unknown variables	Independent Equations
$n_{2,\ total}$	Carbon balance
$n_{3,\ H2}$	Hydrogen balance
$n_{3,\ CO}$	Oxygen balance
$n_{3,\ CO2}$	Nitrogen balance
$n_{3,\ CH4}$	Ash balance
$n_{3,\ N2}$	Energy balance
$n_{3,\ H2O}$	
T_3	
$x_{4,\ ash}$	
$x_{4,\ c}$	
T_4	
Total = 11	Total = 6

Table 4.1 shows the unknown variables. These are the outlet flowrates of all the gaseous components, the outlet flowrates of the solid components, the inlet gasifying agent flowrate and the two outlet stream temperatures. Applying the material and energy balance equations yields six equations against a total of eleven unknowns. Therefore, there are five degrees of freedom if no more equations are specified.

Some method of estimating the outlet gas composition is needed to reduce the number of parameters required in the calculation. The simplest approach is to use empirically derived correlations, specifying ratios of the gaseous components. However, this method does not have a sound theoretical base and would only apply to a specific system for which the empirical data were available. Empirical models of fluid bed gasification have been reviewed and developed by Maniatis. [54]

Another method for estimating outlet gas composition is to consider the gasifier to be a box into which the input streams are fed and where they are allowed to come to chemical equilibrium. This is a common method of modelling gasification but the calculation is performed in different ways according to the assumptions made; also, the choice of assumptions will affect the results of the model. This class of model - the "equilibrium" or "thermodynamic" model - is discussed in Chapter 5.

Kinetics can only be included in this type of model if it is assumed that the reactor is of the lumped parameter or continuous stirred tank type, with constant reaction conditions throughout the reactor. [55]

4.3 STAGEWISE MODELLING

A more sophisticated approach to gasifier modelling is to split the gasifier into zones; usually drying, pyrolysis and gasification. Models of each of these zones are then linked together to produce a whole gasifier model.

As discussed in Chapter 3, fixed bed gasifiers usually have separate zones for each of the processes taking part within them. In these cases, the black boxes of a stagewise model will correspond to the zones present within the gasifier. This means that in models of fixed bed gasifiers, each stage can be treated as a separate reactor, with its own energy balance as well as material balances. However, in fluidised bed systems, there will be mixing of both the gas and the solid phase of the bed, so that the various reactions take place throughout the bed, and are not spatially confined. In a stagewise model of a fluid bed reactor only the overall energy balance is meaningful: the individual energy balances cannot be used as there will be free exchange of energy between the stages. Thus, only in the case of fixed bed gasifiers will the stagewise model give information about the physical conditions in different parts of the reactor. To give an indication of the spatial variation of conditions within a fluid bed reactor, a more sophisticated model is required, as discussed in Section 4.4 below.

It was stated above that the calculation of chemical equilibrium may be used as the method of estimating gas compositions for a black box model. One method of constructing a stagewise model is therefore to use empirical models for the drying and pyrolysis steps, coupled with an equilibrium model of gasification. This approach is discussed in

Chapter 6.

Stagewise models which do not use equilibrium have been constructed, notably those by Belleville and Capart. [56] In their models, the biomass first undergoes a pyrolysis using empirical data. Part of the residue is reacted with steam to give H_2 and CO, and part of the residue is burned to give a mixture of CO, CO_2 and H_2O. The ratio of the CO : CO_2 is determined by an empirical relationship due to Evans and Emmons. [57] Also, part of the pyrolysis gases may react with oxygen to give CO_2 and H_2O. Belleville and Capart present the results of two models, one in which the heat of reaction is provided by combustion of part of the residue with the pyrolysis products remaining unreacted, and the other in which all of the residue is reacted with steam and the heat of reaction is supplied entirely by the combustion of pyrolysis products. Their comparison of the two models was inconclusive: both gave better predictions of fluid bed gasifier performance than an equilibrium model, but no modification of the equilibrium model to take pyrolysis products into account was attempted.

4.4 DIFFERENTIAL MODELLING

The most complex approach to gasifier modelling is to use differential mass and energy balances in the calculations. This can give a picture of the variations of the variables under study throughout the gasifier. Kinetics are usually included in this type of model to predict the product compositions from the progress of the reactions involved. This is the calculation approach commonly used in chemical engineering reactor design where the reaction mechanisms and kinetics are well established.[55]

Schoeters and Buekens have done much work on applying this type of modelling to biomass gasification and have written a comprehensive review paper on the subject to which the reader is referred.[58] Other work by Schoeters and Buekens is also useful.[59,60]

4.4.1 Modelling the Reactions of a Single Particle

The starting point for a differential model of gasification is a model of a single particle within a reactor. The paper by Buekens and Schoeters[58] includes a review of single particle models. However, the modelling of the behaviour of a single particle in a gasifier is complicated by the sequence of processes which occur in the particle. The particle first undergoes drying, then pyrolysis leaving a char residue, which is then gasified by O_2 and steam. Each of these processes must be modelled separately, although there may be interactions between them.

Pyrolysis modelling is a problematic area. Much work has been done on cellulose pyrolysis, particularly by Shafizadeh.[61] However, the pyrolysis of biomass materials is less certain, particularly at the high temperatures of gasification reactors,[54,58] although models of pyrolysis have been constructed, for example by van den Aarsen.[62] Thus, in a gasifier model, reliance has to be placed on the limited empirical data rather than on a robust model of pyrolysis behaviour.

Char gasification is better understood than wood pyrolysis. The models used for biomass char gasification are influenced by work on coal gasification, particularly the work by Johnson.[63,64]

In predicting the reactions of a particle there are two basic models which are used in chemical engineering: the progressive conversion model, in which reaction takes place uniformly throughout a porous particle, and the shrinking core model, in which a reaction front passes through the particle from the surface to the centre with completely reacted material outside the reaction front and an unreacted core within it. [55] Both types of models have been used to model gasification, but Buekens and Schoeters [58] were of the opinion that the progressive conversion model gives a better representation of the reactions of a char particle in a fluid bed gasifier. This is because of the porous nature of wood char which allows diffusion of reactants and products within the particle. [65] A number of single particle models of char gasification are available, of which that by Groeneveld [2] seems to have been widely adopted. [58,62,65,66]

Char gasification models need values of the rate constants of the gas-solid reactions, in addition to coefficients for mass and heat transfer within the particle and through the stagnant film of gas around the particle. Published results for the rate constants of char-gas reactions vary depending on the source of the biomass and the conditions under which the pyrolysis was carried out. [58,60,62,65,67,68,69,70] This means that one of the important parameters for a kinetic model of char gasification is uncertain and must be determined experimentally for the feedstock of interest.

Results from single particle modelling of pyrolysis and gasification kinetics indicate that the pyrolysis reaction is fast compared with the gasification time. Smith and Shand [66] showed that pyrolysis of a wood chip takes around 1 to 3 minutes, depending on the surrounding temperature, compared with gasification times of between 50 minutes and 500 minutes at the same temperatures. This indicates that gasification is the rate determining step for sizing a biomass gasifier.

4.4.2 Kinetic Models

A kinetic model of a gas-solid reactor has two parts: the model of the reactions taking place on a single particle scale, described above, and a model of the flow patterns in the reactor. In fixed bed gasifiers, the solid phase should pass through the reactor in plug flow for a well designed system. This should also be the case for an entrained bed. However, in a fluid bed system there may be complex flow patterns. [55] The flow of gases within the reactor is also important. Again, there should be plug flow in fixed bed and entrained reactors, but in a fluid bed there is the complexity of bubbles within the bed.

Kinetic models of fluid bed gasifiers have, therefore, introduced models of fluidisation to predict the flow patterns within the bed and their effect on gas-solid contacting within the bed. Again, Buekens and Schoeters [58] have reviewed models of this type, and Schoeters [60] has tested a number of models with different treatments of the fluid bed hydrodynamics. More recently, van den Aarsen [33,62] has developed a model which incorporates empirically derived pyrolysis relationships into this type of model. However, these models do not include secondary reactions of volatile pyrolysis products other than those which take part in the homogeneous water-gas shift reaction. These models were successful in predicting the char hold-up in the reactor, and by implication the gasification rate, as well as predicting the outlet gas composition with reasonable accuracy.

4.4.3 Other Differential Models

It is possible to construct models of gasification which use the differential material and energy balances whilst representing the chemical reactions taking place without the use of kinetics. A notable model of this type is described by Bacon et al. [71] This model was based on the use of assigned degrees of approach to equilibrium for the reactions involved, together with the Modified Bubble Assemblage Model of fluidisation. [72] The results obtained using this model indicated that the gas composition, with the exception of CH_4, was almost entirely dependent on the assigned degree of approach to equilibrium. The values of design variables had a negligible effect on the product gas composition, and the conclusion was that a fluid bed gasifier can be designed solely on the basis of achieving stable fluidisation behaviour.

4.5 TRANSIENT MODELS

Transient models of gasification have also been proposed, in addition to the steady-state models discussed above. They could be used for predicting the response of a gasifier to fluctuating reaction conditions or throughput. They could also be of use in devising control strategies for gasifiers. As the author's work has concentrated on steady-state models, transient models are not considered here. The interested reader is referred to the review by Buekens and Schoeters, [58] and the thesis by Schoeters [60] for a description of this type of model.

4.6 CONCLUSIONS

A wide variety of biomass gasifier models have been constructed at varying levels of complexity and for a number of purposes. Considering models to predict the performance of gasifiers, models range from simple equilibrium models to complex kinetic models incorporating the effects of mass and heat transfer on the gasification reactions. With the exception of the equilibrium models, which are based on simple thermodynamic relationships, experimental data are needed for the calculations. In stagewise models this may be empirical data on pyrolysis products, whilst in kinetic models the values of the rate constants must be determined, in addition to pyrolysis data.

When discussing complex kinetic models of fluid bed gasification, Buekens and Schoeters stated: "The results of computation are sometimes so detailed that their experimental verification is not yet feasible". [58] If what is required is a reasonably accurate prediction of the gasifier heat and material balance, it is clearly sensible to use a simple model rather than a complex kinetic one if both predict the results with a similar degree of accuracy. However, kinetic models are needed for sizing gasifiers in design.

Nomenclature, Chapter 4

$F_{i,j}$	Mass flow of component j in stream i
$n_{i,j}$	Molar flow of component j in stream i
P	Total pressure
p_j	Partial pressure of component j
T	Temperature of reaction
T_i	Temperature of stream i
x_j	Mole fraction of component j
$y_{i,j}$	Mole fraction of component j in stream i

CHAPTER 5
GASIFIER EQUILIBRIUM MODELLING

5.1 CALCULATION OF C-H-O EQUILIBRIUM

As mentioned in Chapter 4, one of the methods of predicting gas compositions for a gasifier is to assume that the outlet gas is at chemical equilibrium. With the elements of carbon, hydrogen, oxygen and nitrogen present there are many chemical species which could possibly exist at equilibrium. However, Desrosiers [24] has shown that under normal gasification conditions the only species present at a concentration of greater than 10^{-4} mol % are H_2, CO, CO_2, CH_4, N_2 and H_2O. The following reactions are all important in the gasification process and include all of the species listed:

$$C + O_2 = CO_2 \qquad \Delta H_{298} = -393.8 \text{ kJ/mol} \qquad [5.1]$$

$$C + CO_2 = 2\,CO \qquad \Delta H_{298} = +172.6 \text{ kJ/mol} \qquad [5.2]$$

$$C + H_2O = CO + H_2 \qquad \Delta H_{298} = +131.4 \text{ kJ/mol} \qquad [5.3]$$

$$CO + H_2O = CO_2 + H_2 \qquad \Delta H_{298} = -41.2 \text{ kJ/mol} \qquad [5.4]$$

$$C + 2\,H_2 = CH_4 \qquad \Delta H_{298} = -74.9 \text{ kJ/mol} \qquad [5.5]$$

However, in the calculation of equilibrium calculations it is not necessary to consider equilibrium of all these reactions. Both Gumz [9] and Desrosiers [24] have shown that there is no O_2 present at equilibrium, and equation [5.1] can be considered to go to completion. Of the other possible reactions only three are required to completely define the equilibrium, provided that all possible species are included in the three equations shown. [73]

Baron, Porter, and Hammond [74] considered equations [5.3], [5.4] and [5.5] in order to calculate equilibrium. In this system of equations solid carbon is assumed to be present, and to calculate the gas composition only the H:O and N:O ratios need to be known. Nitrogen takes no part in the reaction, and is merely an inert diluent. Because reactions [5.3] and [5.5] have an unequal number of gaseous species on either side of the equation, the position of the equilibrium is affected by pressure. The addition of a diluent at a fixed total pressure reduces the partial pressures of the reactive components, thus affecting the final gas composition.

Assuming ideal gas behaviour, i.e. at pressures less than 50 atmospheres approximately, equations [5.3] to [5.5] lead to the following set of equilibrium relationships:

$$K_1 = \frac{p_{CO}\ p_{H2}}{p_{H2O}} = \frac{P\ x_{CO}\ x_{H2}}{x_{H2O}} \qquad [5.6]$$

$$K_2 = \frac{p_{CO2}\ p_{H2}}{p_{CO}\ p_{H2O}} = \frac{x_{CO2}\ x_{H2}}{x_{CO}\ x_{H2O}} \qquad [5.7]$$

$$K_3 = \frac{p_{CH4}}{p_{H2}^{2}} = \frac{x_{CH4}}{P\ x_{H2}^{2}} \qquad [5.8]$$

The equilibrium gas composition is obtained by solving this set of equations together with the element balances. However, the equations cannot be solved analytically, and an iterative solution is necessary. Baron, Porter and Hammond [74] produced an iterative scheme based on estimating the equilibrium $H_2 : H_2O$ molar ratio. The equations are re-arranged in terms of this parameter, to give a quadratic in x_{H2}. Solving the quadratic enables a trial gas composition to be calculated. The $H_2 : H_2O$ molar ratio is compared with the estimated value, and the calculation repeated until the calculated and estimated ratios are equal.

The equilibrium constants for the equations are calculated using the following equation:

$$\log_e (K_i) = \sum \frac{- \nu_j G_{j,T}}{R T} \qquad [5.9]$$

Polynomials for the calculation of $G_{j,T}$ as a function of temperature have been tabulated by Baron, Porter and Hammond.

Desrosiers used a different approach to the calculation of the equilibrium composition, using a general-purpose equilibrium calculating program developed by NASA. 24 However, this requires a large datafile of thermochemical data derived from the JANAF tables, 75 as opposed to the simple polynomials used by Baron, Porter and Hammond. Older methods of equilibrium calculation are reviewed in Gumz. 9 They are mostly graphical, and not readily modified for use in computer programs.

A second type of equilibrium calculation where there is no free carbon present may also be of interest. Here, the total quantities of carbon, hydrogen, oxygen and nitrogen must be known in order to calculate equilibrium. As there is one chemical species fewer (there being no solid carbon), one fewer equation is required for the solution of the problem. The two reactions considered are:

$$CO + H_2O = CO_2 + H_2 \qquad [5.10]$$

$$2 H_2 + 2 CO = CH_4 + CO_2 \qquad [5.11]$$

These chemical reactions give rise to the following equilibrium relationships:

$$K_4 = \frac{p_{CO2}\ p_{H2}}{p_{CO}\ p_{H2O}} = \frac{x_{CO2}\ x_{H2}}{x_{CO}\ x_{H2O}} \qquad [5.12]$$

$$K_5 = \frac{p_{CO2}\ p_{CH4}}{p_{CO}^2\ p_{H2}^2} = \frac{x_{CO2}\ x_{CH4}}{P^2\ x_{CO}^2\ x_{H2}^2} \qquad [5.13]$$

These equations must again be solved iteratively, and the method used in the author's work is presented in Appendix A.

The above case of chemical equilibrium would, at first sight, not seem to be applicable to a gasifier, where solid carbon is present. The usefulness of this type of calculation will be discussed later in this chapter.

5.2 INCLUSION OF EQUILIBRIUM CALCULATIONS IN GASIFIER MODELS

5.2.1 Degrees of Freedom and Assumptions.

The simplest type of equilibrium model treats the gasifier as a black box in which the inlet streams react and reach chemical equilibrium, giving two product streams, a gas and a solid. Figure 5.1 shows the input and outputs of such a gasifier. Some variables will be fixed by the environment and the choice of feedstock: these include the feedstock composition and ash content, and the feedstock temperature. Others will be chosen by the designer: these include the gasifying agent composition and temperature, the biomass moisture content, and the gasifier pressure. A gasifier heat loss may also be chosen, and the gasifier insulation specified accordingly. The variables which are fixed in either of these ways are shown in *italics* in Figure 5.1. The unknown variables (shown in **bold** in Figure 5.1) and the material and energy balance equations are tabulated in Table 5.1.

Depending on the relative quantities of biomass and gasifying agent, there are two possible cases:

i. Some free carbon is present in the output solids

ii. Only ash is present in the output solids, i.e. all the carbon is in the gas phase

In the first case, the equilibrium calculation will be of the first type described in Section 5.1, where three equilibrium relationships are used. In the second case, the lack of solid carbon means that the second type of equilibrium calculation must be used, with only two equilibrium relationships plus the constraint that no solid carbon is produced.

Table 5.1 shows that in both of these cases, there are twelve equations and nine variables, and this implies that either additional assumptions must be made in order to provide additional equations, or that the values of some of the unknown variables must be fixed, or a combination of both of these must be undertaken in order to allow the solution of the equations.

The main area where additional assumptions can be made is in the relationship between the outlet temperatures and the reaction temperature. The assumptions which may be made depend on the type of gasifier being modelled. In fixed bed types of gasifier, the reaction temperature will not be the same as the equilibrium temperature. However, in a fluid bed, it may reasonably be assumed that the gas outlet temperature, the solids outlet temperature and the reaction temperature are all equal, because of the good mixing.

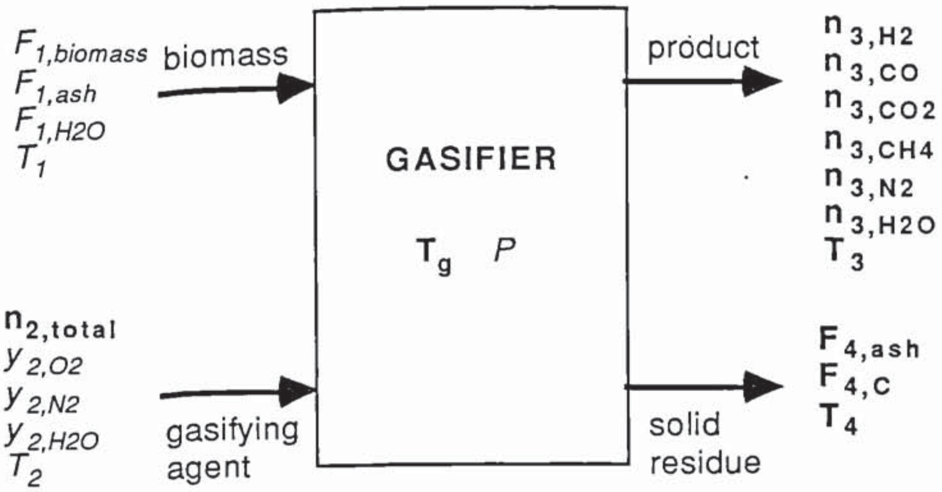


Figure 5.1 Flows and Variables in a Typical Gasifier

Table 5.1 Unknown Variables and Equations in the Black Box Equilibrium Model

Variables	Balance Equations	Additional Equations Solid Carbon Present	Additional Equations No Solid Carbon Present
T_g	Carbon balance	Equilibrium of:	$F_{4,C} = 0$ (no solid C)
T_3	Hydrogen balance	$C + H_2O = CO + H_2$	Equilibrium of:
T_4	Oxygen Balance	$CO + H_2O = CO_2 + H_2$	$CO + H_2O = CO_2 + H_2$
$n_{2,total}$	Nitrogen Balance	$C + 2\,H_2 = CH_4$	$2\,CO + 2\,H_2 = CH_4 + CO_2$
$n_{3,H2}$	Ash Balance		
$n_{3,CO}$	Energy Balance		
$n_{3,CO2}$			
$n_{3,CH4}$			
$n_{3,N2}$			
$n_{3,H2O}$			
$F_{4,ash}$			
$F_{4,C}$			

Total numbers of equations and variables

Variables = 12

Equations (free solid carbon present) = 6 balance + 3 others = 9

Equations (no solid carbon present) = 6 balance + 3 others = 9

This assumption is also consistent with the well mixed black box approach to equilibrium modelling mentioned above, and this assumption has been used in the author's work. This assumption about the temperatures gives two additional equations for the modelling:

$$T_g = T_3 \qquad [5.14]$$

$$T_g = T_4 \qquad [5.15]$$

There remains one degree of freedom in the calculation. The two variables which might be fixed by the designer are the reaction temperature T, or the gasifying agent inlet flowrate $n_{2,total}$. Of these, it is easier to specify a reaction temperature, for a number of reasons. The kinetics of the gasification reactions are dependent on temperature, and if the temperature is too low, then the gas composition will not reach equilibrium within a reasonable time. If the temperature is too high, then there will be problems with the materials of construction of the gasifier.

Instead of specifying either gasifying agent input flowrate or reaction temperature, it is possible to make one additional assumption. Consider a black box gasifier using a gasifying agent containing oxygen or air. First, a small amount of gasifying agent is added to the biomass, and the mixture allowed to come to equilibrium. There will be a gas product, together with some solid carbon residue. If more gasifying agent is added, some of the solid will undergo partial combustion to carbon monoxide and the temperature will increase. If more gasifying agent is added, a point will be reached where the solid carbon will disappear. Beyond this point, the addition of more gasifying agent will boost the temperature of the gases by combustion of some of the hydrogen and carbon monoxide, reducing the total amount of chemical energy in the gas product in favour of sensible heat. Clearly, the point at which the solid carbon disappears is the point at which there is maximum chemical energy in the gas phase. This point is called the "carbon boundary". From the point of view of equilibrium modelling, this point may be found by taking the set of equilibrium equations where solid carbon is present, but adding the additional constraint that the solid carbon residue tends to zero. If this assumption is used with equations [5.14] and [5.15], then there are no degrees of freedom in the calculation. For a set of gasifier operating conditions, (defined by the biomass composition, moisture content and temperature, the gasifying agent composition and temperature, the gasifier pressure, and gasifier heat loss) there is a unique, idealised, optimum operating point which may be called the "equilibrium carbon boundary".

5.2.2 Implementation of the Gasifier Model on the Computer

The usual method of solution of a set of simultaneous equations is to develop an iterative scheme for the simultaneous solution of all of the equations. However, in the case of an equilibrium model of gasification, there are eleven or twelve equations to be solved simultaneously, and some of the equations are non-linear, so that it is difficult to devise an iterative scheme which gives reliable convergence. In addition, fairly simple and straightforward iterative schemes have been developed for the solution of the equilibrium equations, as described above. Therefore, a useful alternative to the simultaneous solution of the model equations is to split the calculation into two parts. The energy balance may be solved simultaneously with the mass balances in a simple iterative calculation, if the equilibrium calculation is performed for each iteration of the balance equations. This approach has the disadvantage of having two levels of iteration, one (the equilibrium calculation) being within the other (the energy balance calculation), adding to the computer time involved. However, the method has the advantage of being easier to understand than a complex one-level iterative scheme, and allows tested methods to be used for the calculation of equilibrium.

In the computer program of the model, the equilibrium calculation is a separate subroutine which can be separately tested, and which can be used in other programs. The mass and energy balances are performed by a separate subroutine which calls the equilibrium subroutine as part of its calculation (see Figure 5.2).

The program developed is called "MODEL" and is written in FORTRAN77. A flow diagram and listing is included as Appendix B. The model can perform three different types of calculation:

i. carbon boundary calculation
ii. fixed temperature calculation where there is no solid carbon present
iii. fixed temperature calculation where solid carbon is present

In each case, the equilibrium carbon boundary calculation is performed first. This calculation gives the equilibrium carbon boundary temperature, and the amount of gasifying agent required to gasify the carbon completely. If a temperature for gasification is specified, as in cases ii. and iii. above, then the program compares the specified temperature with the carbon boundary temperature. If the carbon boundary temperature is below the specified temperature, this implies that no solid carbon will be present, as discussed in Section 5.2.1, and a calculation is performed in which the gas product from the carbon

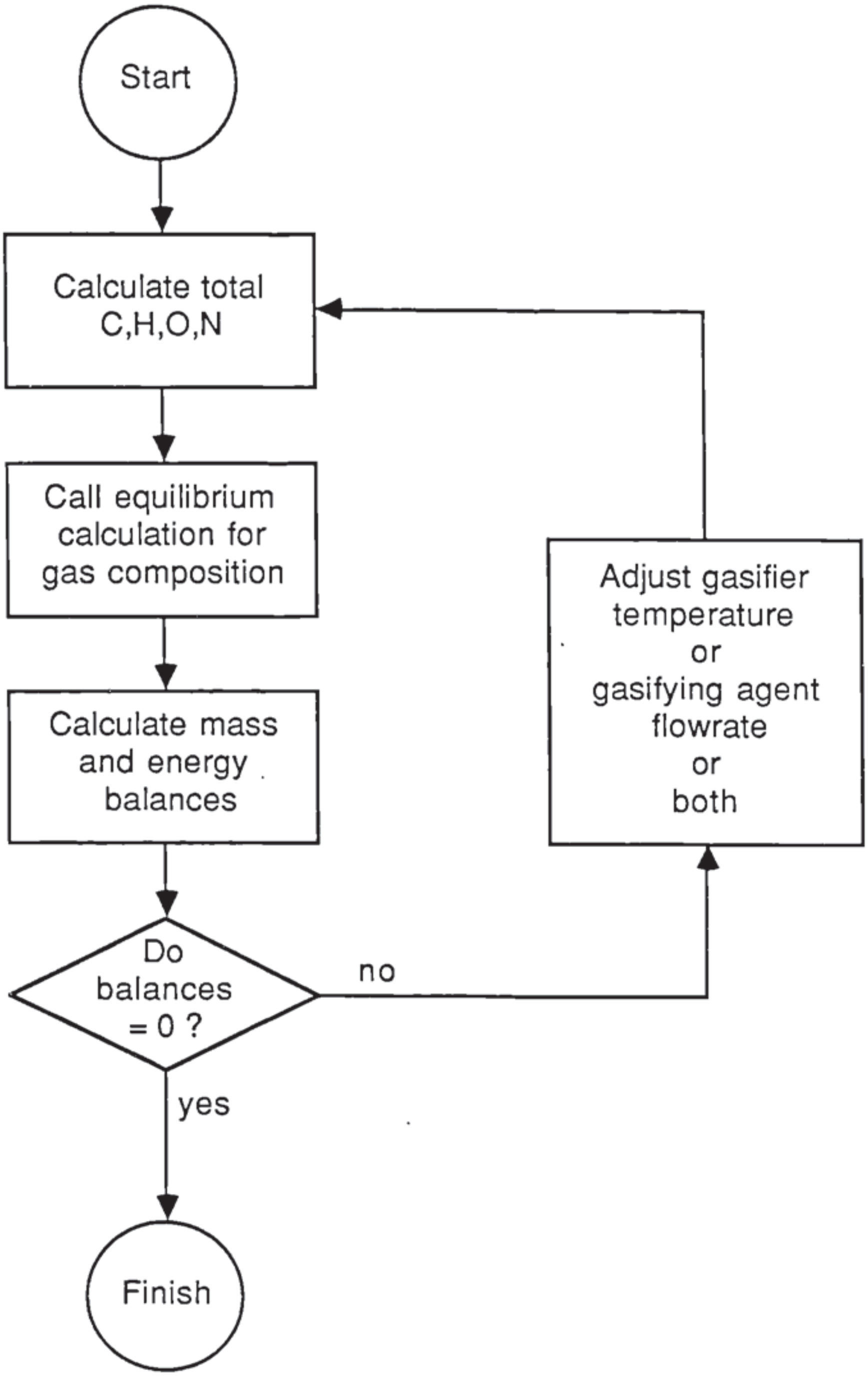


Figure 5.2 **Basic Flow Diagram of an Equilibrium Gasifier Model**

boundary calculation has further gasifying agent added to it, and which comes to equilibrium. The amount of extra gasifying agent added is iterated until the energy balance is satisfied at the specified temperature. In the other case, where the carbon boundary temperature is above the temperature specified, then some free carbon must be present at the end of the gasification. The calculation therefore uses the free carbon equilibrium, and a simple iteration of the gasifying agent flow is employed until the energy balance is satisfied at the required temperature.

The first stage in these calculations, the equilibrium carbon boundary calculation, is more complex than these two simple iterations. In this case there are two variables which must be calculated simultaneously in the heat balance calculation, and a two-dimensional iterative scheme must be used. In the program MODEL, a two dimensional simplex method is used, which searches for the point at which both the error in the energy balance, and the amount of solid carbon in the gasification products are both zero. The variables which are calculated simultaneously are the reaction temperature and the gasifying agent flow rate.

5.3 USE OF THE MODEL IN A SENSITIVITY ANALYSIS

5.3.1 Introduction

There are several variables to be chosen in the carbon boundary calculation which are set by the designer. The values of some of these are set as a result of choosing the gasifying agent and feedstock, and others may be set by the designer, such as the gasifier pressure. Sets of results of thermodynamic gasifier calculations covering a range of conditions have been published before, for example by Desrosiers [24], Gumz [9], Baron, Porter and Hammond [74], and Gibbins [76]. However these calculations have used models of one of three types: models where temperature is specified and input gasifying agent flow iterated to satisfy the heat balance, models where input gasifying agent is specified and temperature is iterated, or models where the energy balance is not used in the calculation, and both gasifying agent flow and temperature are specified. The results from equilibrium carbon boundary calculations have not been published.

However, as was stated above, the equilibrium carbon boundary case is the optimum, idealised case of gasification, and it is useful to look at the effects of varying various parameters on the carbon boundary point. A sensitivity analysis has therefore been performed using the carbon boundary model.

Because of the number of possible variables which may be specified in the calculation (see Table 5.2) it was considered to be impractical to carry out a full factorial analysis. If only two points were taken for each variable, this would have given $2^{16} = 65536$ calculations. It was therefore decided to adopt a sensitivity analysis type of approach, with each variable being varied about a base case. Two base cases were adopted, one representing a typical air gasifier, and the other a typical oxygen gasifier.

Table 5.2 Input Variables in Computer Program "MODEL"

Variable	Oxygen Gasifier Base Case	Air Gasifier Base Case
Gasifier Pressure, atm. absolute	1.0	1.0
Gasifier Heat Loss, % chemical energy of feedstock	5.0	5.0
Carbon content of the biomass, dry ash free molar basis	1.0	1.0
Hydrogen content of the biomass, dry ash free molar basis	1.5	1.5
Oxygen content of the biomass, dry ash free molar basis	0.7	0.7
Moisture content of the biomass, % of dry ash free mass	25.0	25.0
Ash content of the biomass, % of dry ash free mass	1.0	1.0
Temperature of the gasifying agent, K	298.15	298.15
Enthalpy of gasifying agent steam, kJ / kg	2790.0	2790.0
Gasifying agent H_2 mole fraction	0.0	0.0
Gasifying agent CO mole fraction	0.0	0.0
Gasifying agent CO_2 mole fraction	0.0	0.0
Gasifying agent CH_4 mole fraction	0.0	0.0
Gasifying agent N_2 mole fraction	0.0	0.79
Gasifying agent H_2O mole fraction	0.0	0.0
Gasifying agent O_2 mole fraction	1.0	0.21

5.3.2 Results for an Oxygen Gasifier

The results of the sensitivity analysis for an oxygen gasifier are presented in the form of graphs. The more important results are included here as Figures. 5.3 to 5.11. The other results are shown in Appendix C (see Figures C1 to C4).

Figures 5.3 and 5.4 show the effect of varying the elemental composition of the biomass feedstock. In all cases, the feedstock ultimate analysis is quoted as a molar composition on a C_1 basis. Figure 5.3 shows the effect of varying the hydrogen content of the biomass. As might be expected, the main effect of this is to increase the proportion of hydrogen-containing species in the outlet gas stream. The oxygen content of the feedstock has a more marked effect on performance (Figure 5.4). In the program "MODEL" the biomass calorific value is calculated using the IGT equation which was described in Chapter 2. The result of increasing the oxygen content of the biomass is to reduce its calorific value. This affects the energy balance of the gasifier, reducing the temperature and thus favouring CO_2 and H_2O which are stable at lower temperatures. The increased amount of oxygen added to the system also affects the mass balances, but this is a lesser effect.

Figure 5.5 shows the effect of ash content on the performance of the gasifier. Increasing the ash content has the effect of reducing the biomass heating value according to the IGT equation. Increasing ash content therefore has similar effects to increasing oxygen content of the biomass, except that it does not affect the material balances.

The results of increasing the moisture content of the feedstock are shown in Figure 5.6. Both the mass and energy balances are influenced - more hydrogen is introduced into the gasifier, which tends to increase the concentration of hydrogen containing species, and moisture reduces the temperature due to two effects. The latent heat of vaporisation of the moisture must be supplied, abstracting heat from the system. Also, the reaction between H_2O and carbon is endothermic, again abstracting available heat from the system. Thus, an increase in moisture content of the biomass strongly reduces the carbon boundary temperature. As a result of the decreased equilibrium temperature, H_2O, CO_2 and CH_4 are favoured, whilst CO and H_2 are reduced.

The effect of gasifying agent is shown by Figures 5.7, 5.8, and 5.9. Adding nitrogen to the gasifier obviously increases the N_2 in the product gas, at the expense of all other species. The temperature is slightly reduced, due to the need to provide the sensible heat to raise the temperature of the added nitrogen to the gas outlet temperature. Adding steam to the gasifying agent has a similar effect to increasing the moisture content of the feedstock. However, in this case the latent heat does not need to be provided, and the effect on the heat balance is less strong. Gasifying agent temperature is seen to have practically no effect on the results. This is because the sensible heat of the gasifying agent is a much smaller term in the energy balance than the changes of chemical energy.

Figure 5.10 shows the effect of heat loss. The results of this are very similar to the results of biomass ash content, because both of these cases affect the heat balance only.

The final graph, Figure 5.11, shows the effect of pressure. Here, the effects of equilibrium interact strongly with the heat and mass balance effects, and it is difficult to explain the results clearly. Methane is favoured by higher pressures, according to Le Chatelier's principle, but the effects of equilibrium, material balance, and increased temperature with increased pressure cancel each other out in the case of the carbon oxides, the concentrations of which remain approximately constant.

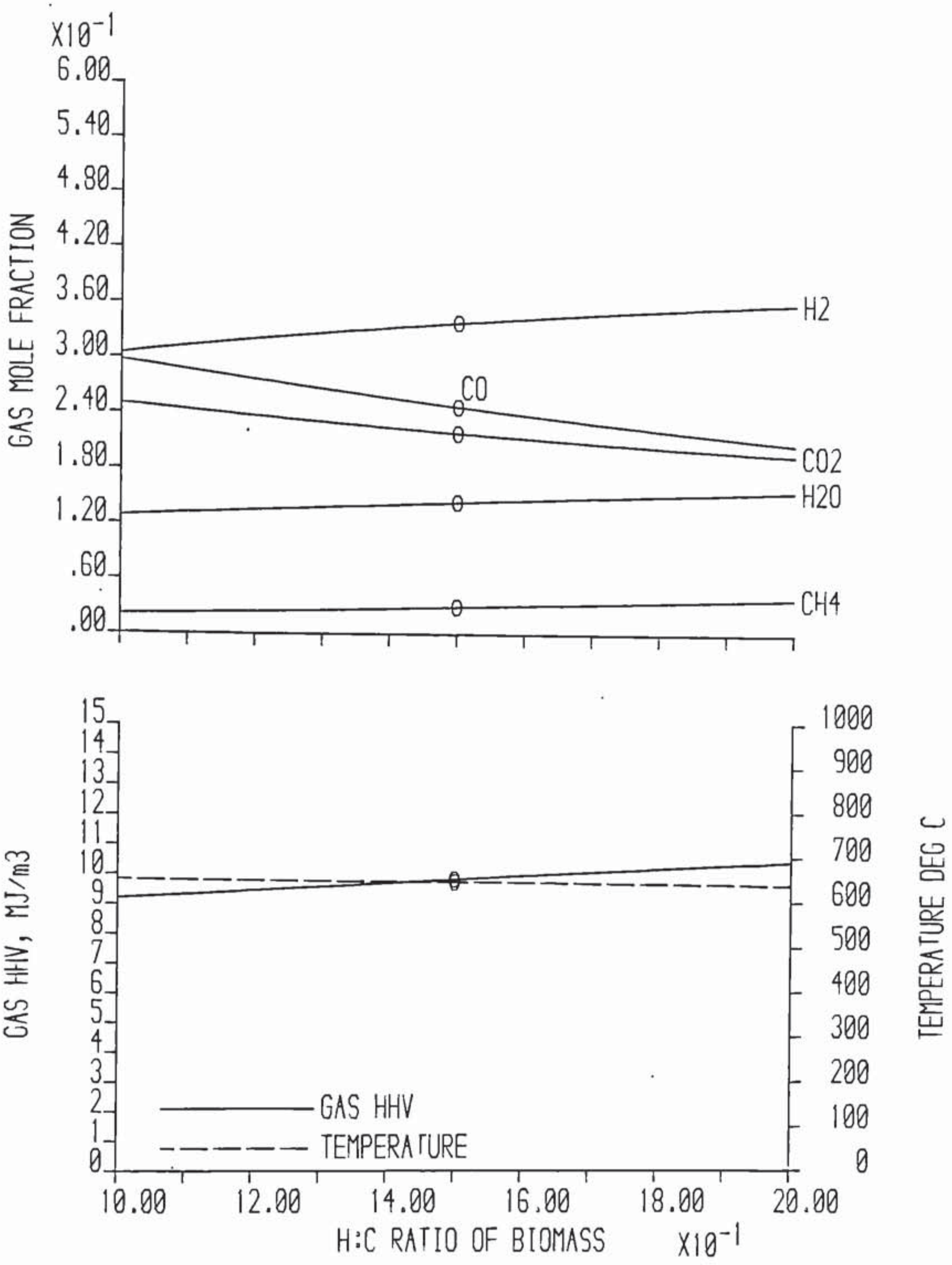


Figure 5.3 THEORETICAL GASIFIER PERFORMANCE AGAINST
HYDROGEN CONTENT OF BIOMASS
OXYGEN GASIFER

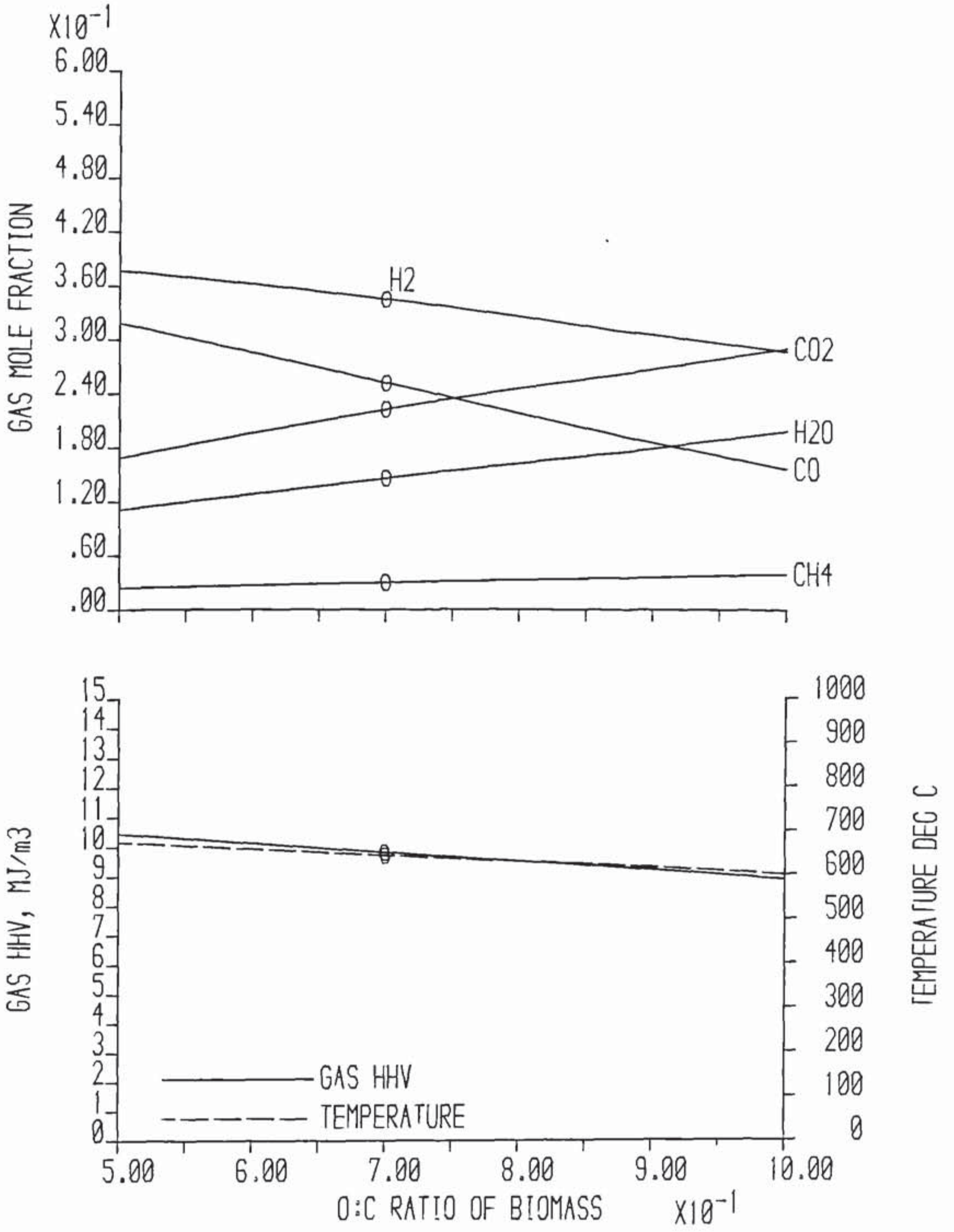


Figure 5.4 THEORETICAL GASIFIER PERFORMANCE AGAINST
OXYGEN CONTENT OF BIOMASS
OXYGEN GASIFER

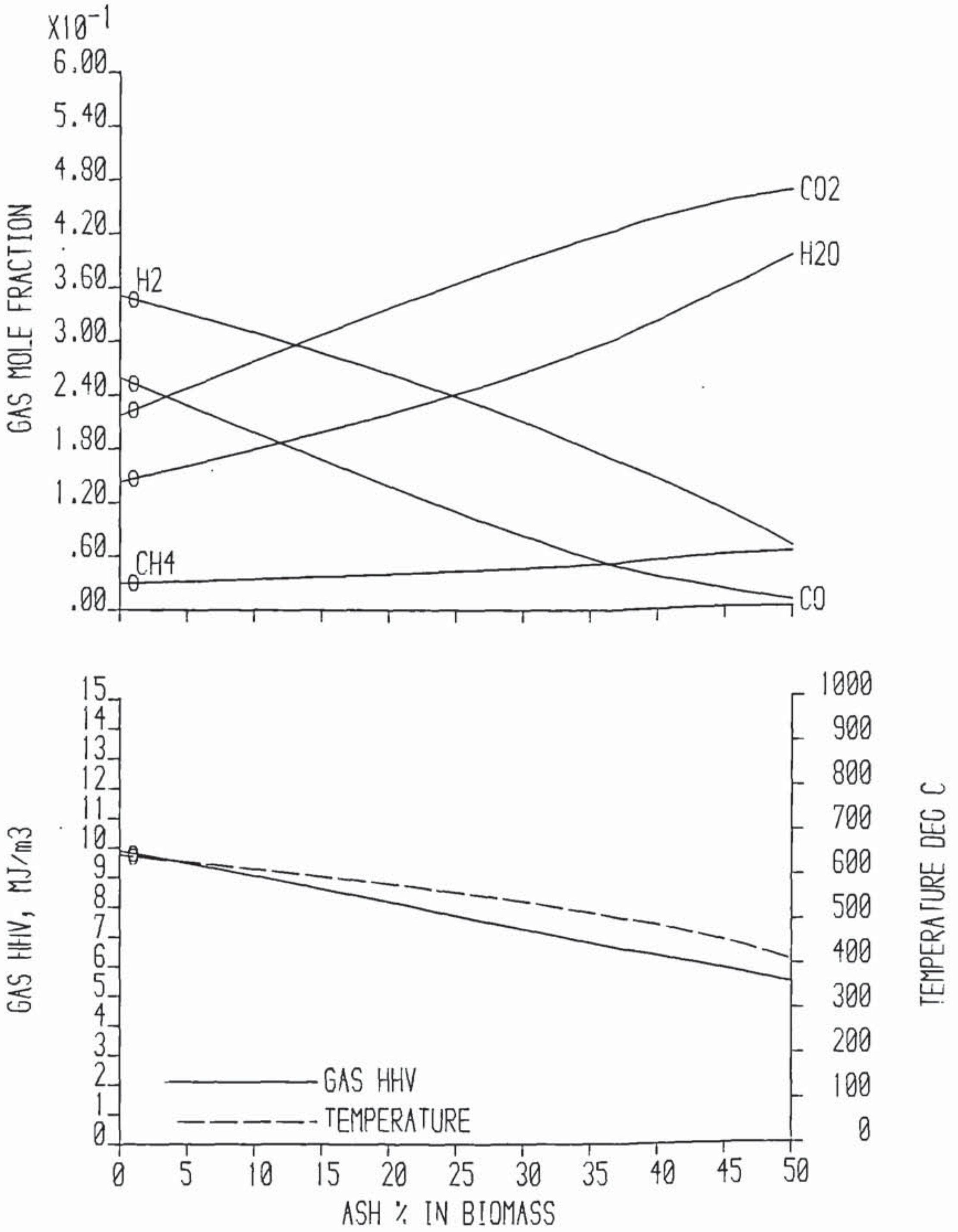


Figure 5.5 THEORETICAL GASIFIER PERFORMANCE AGAINST
BIOMASS ASH CONTENT
OXYGEN GASIFER

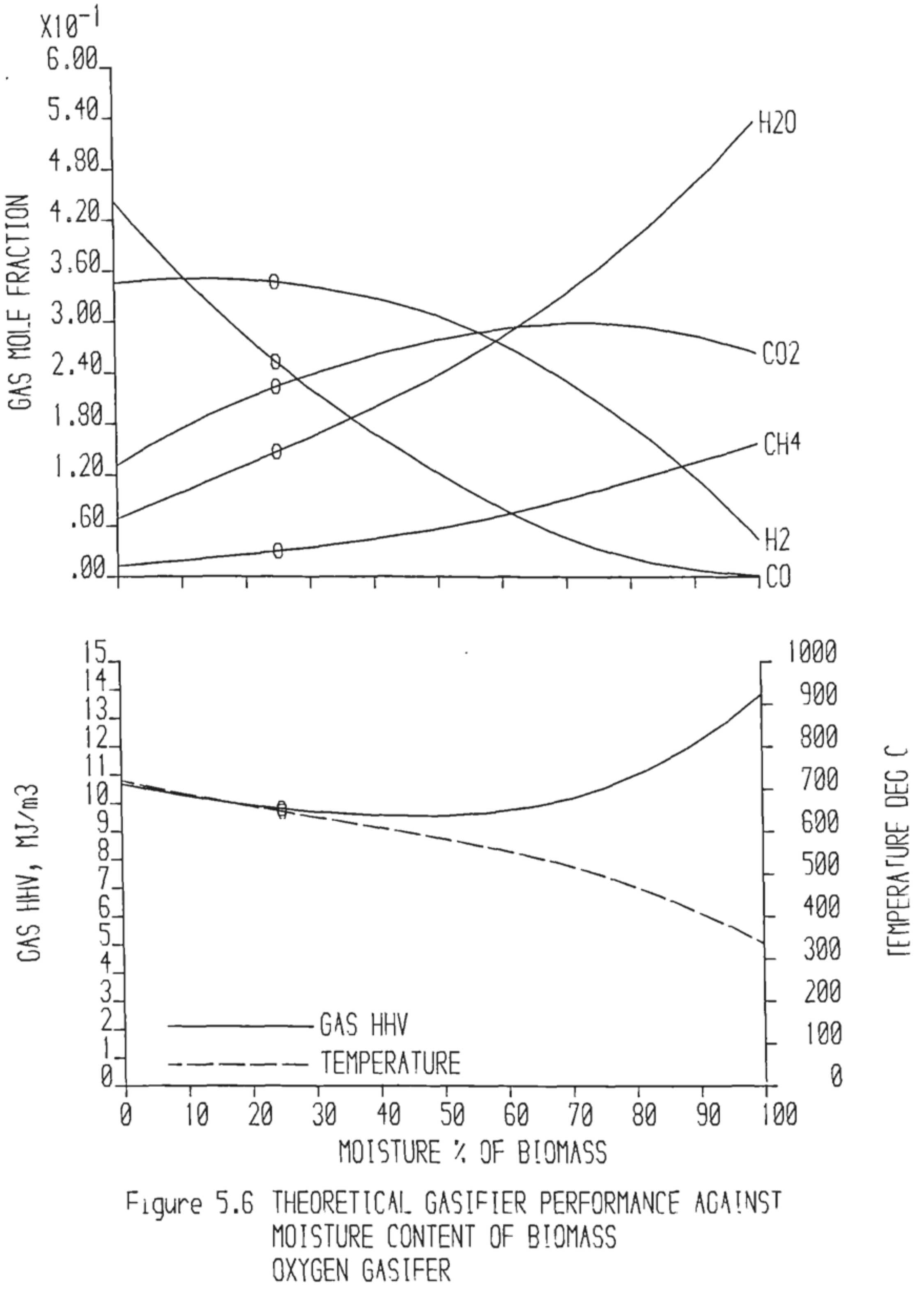


Figure 5.6 THEORETICAL GASIFIER PERFORMANCE AGAINST
MOISTURE CONTENT OF BIOMASS
OXYGEN GASIFER

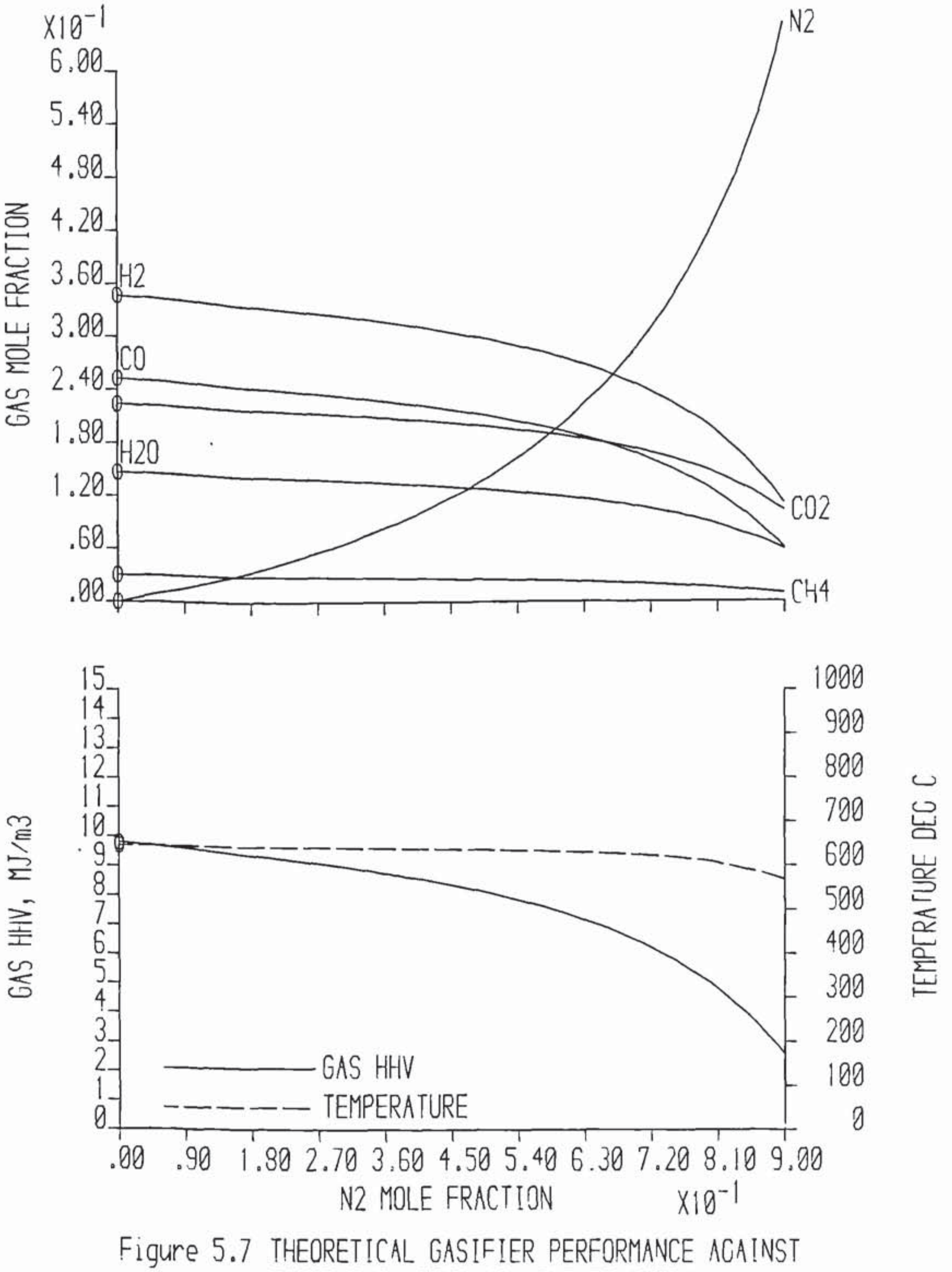


Figure 5.7 THEORETICAL GASIFIER PERFORMANCE AGAINST GASIFYING AGENT N2 MOLE FRACTION OXYGEN GASIFER

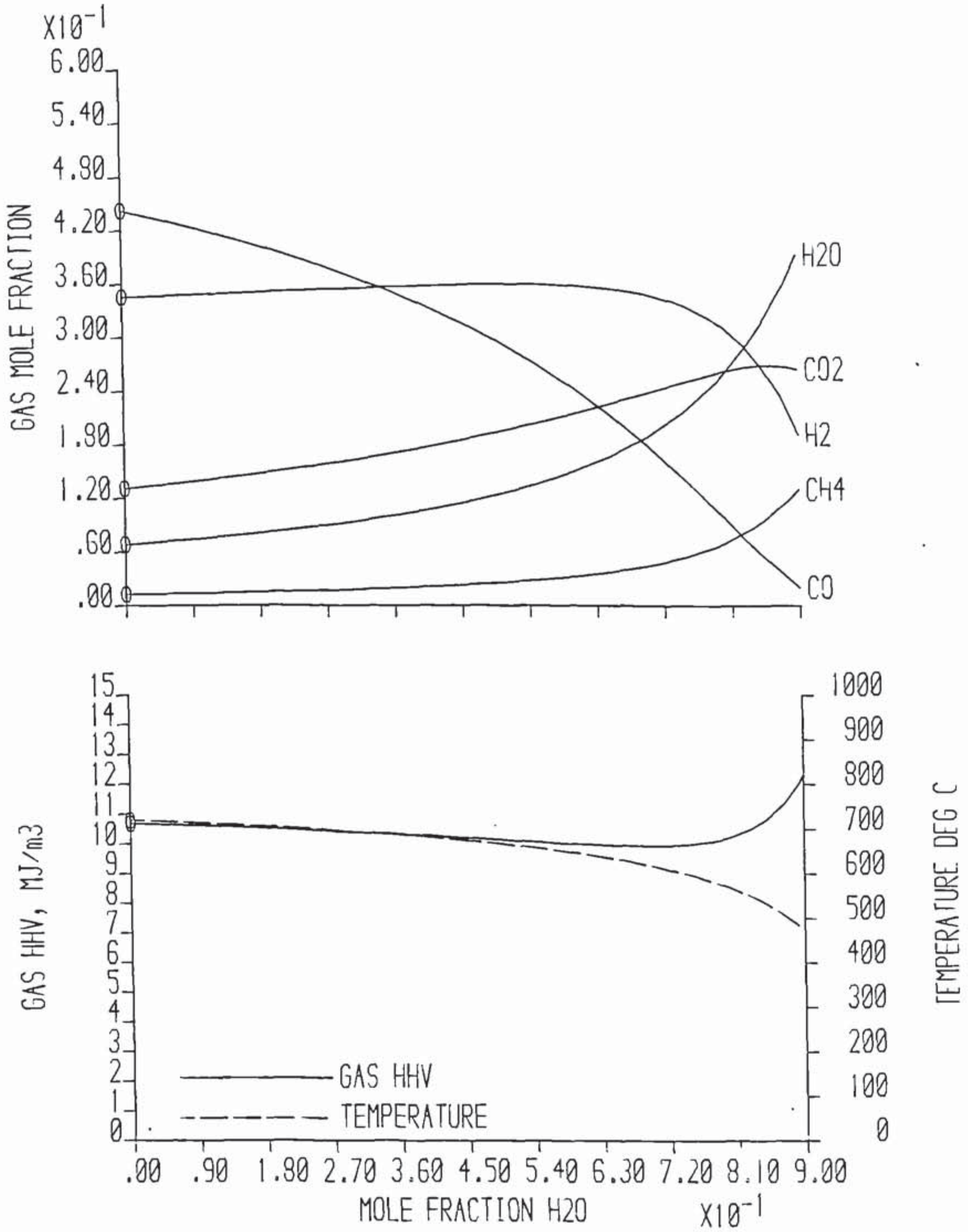


Figure 5.8 THEORETICAL GASIFIER PERFORMANCE AGAINST
INLET GAS STEAM MOLE FRACTION
OXYGEN GASIFER

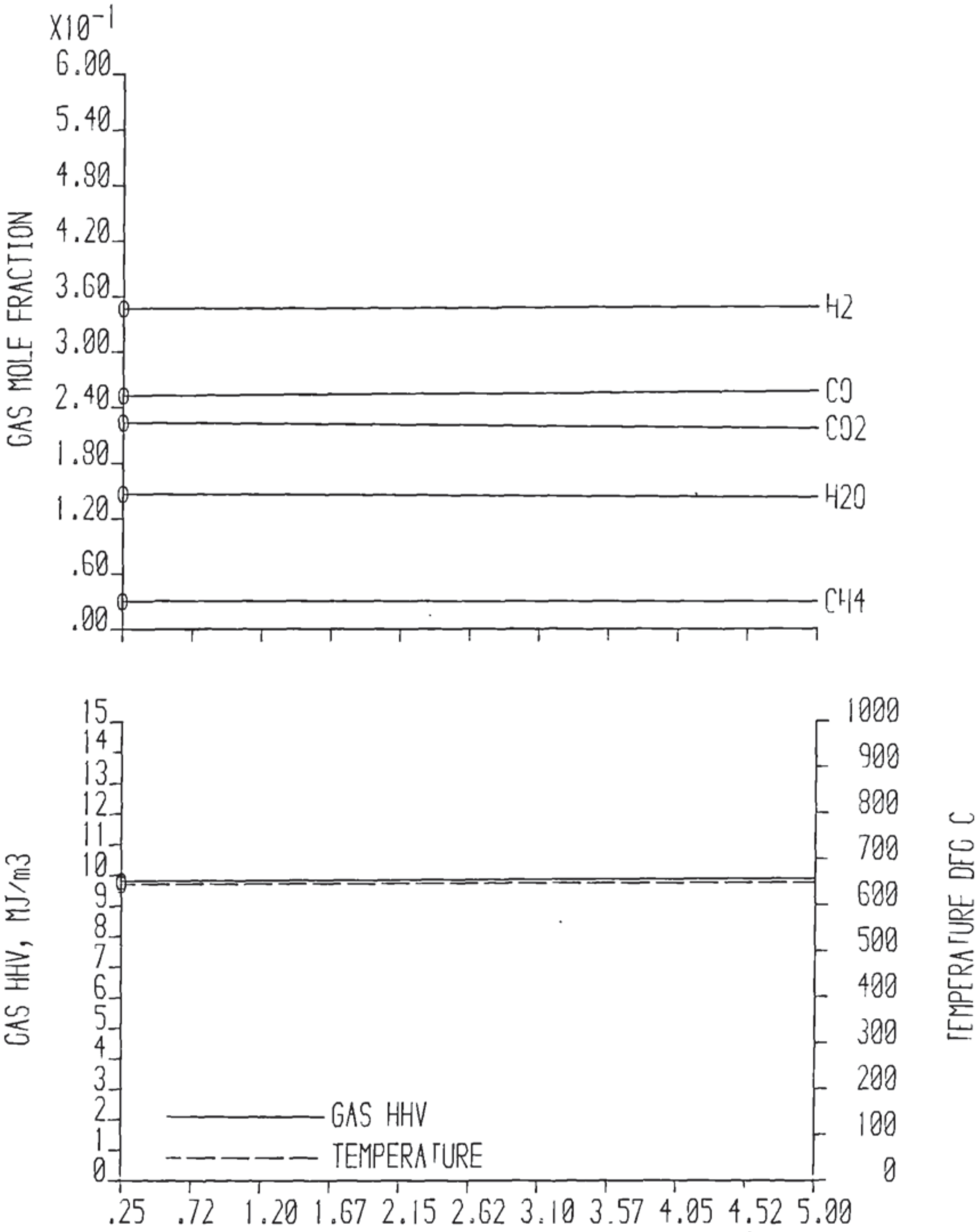


Figure 5.9 THEORETICAL GASIFIER PERFORMANCE AGAINST GASIFYING AGENT TEMPERATURE OXYGEN GASIFER

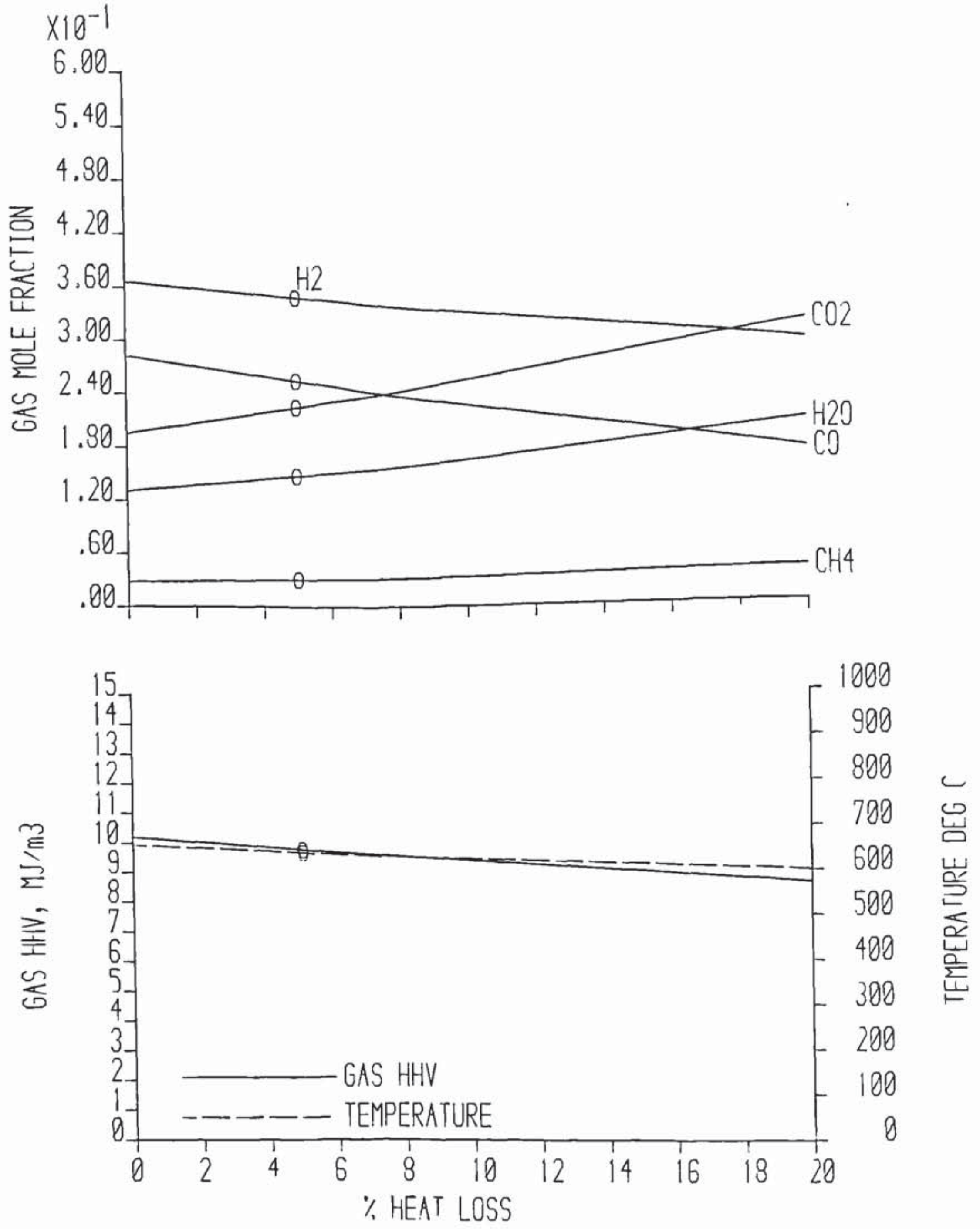


Figure 5.10 THEORETICAL GASIFIER PERFORMANCE AGAINST
GASIFIER HEAT LOSS
OXYGEN GASIFER

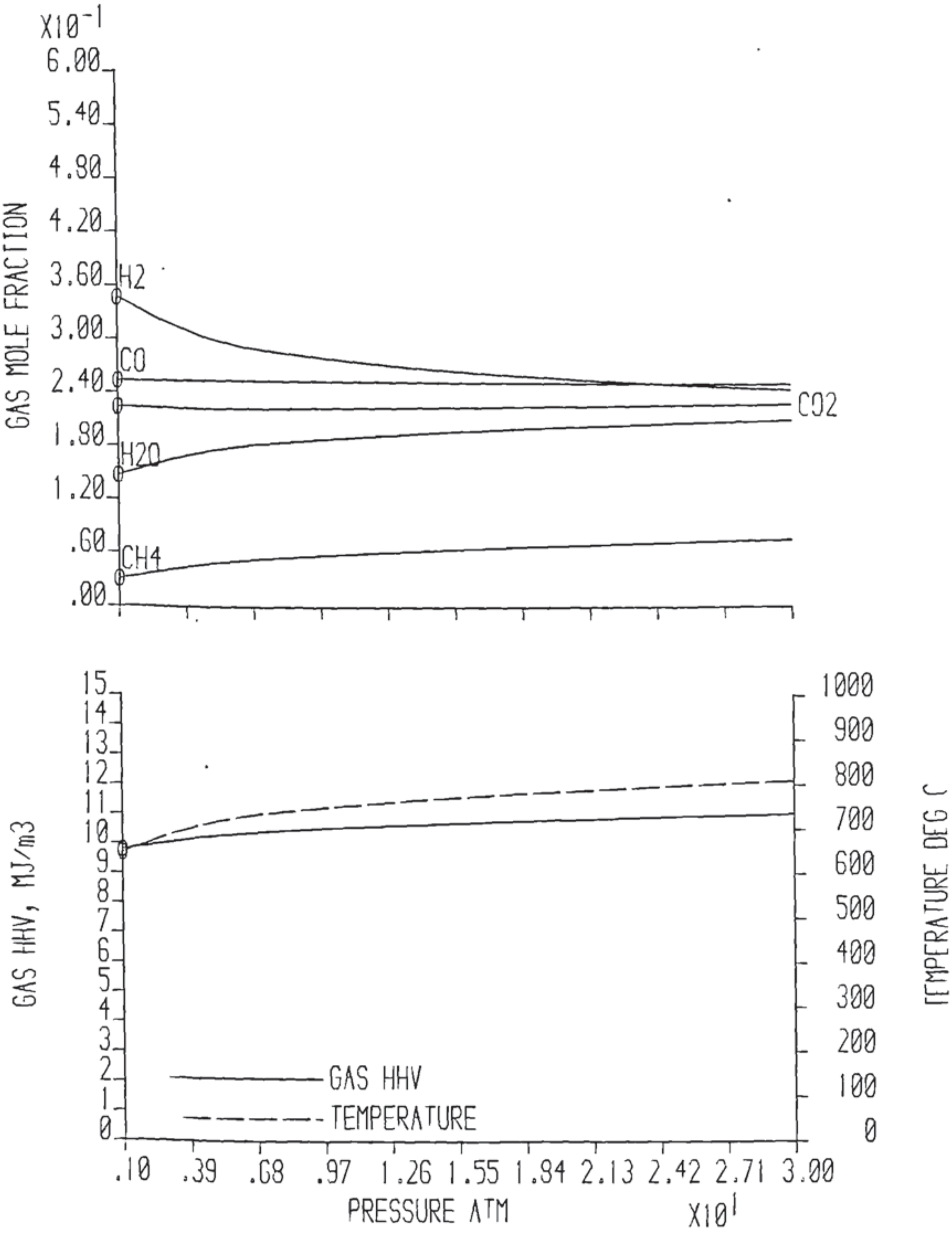


Figure 5.11 THEORETICAL GASIFIER PERFORMANCE AGAINST
GASIFIER PRESSURE
OXYGEN GASIFER

The results of adding gases other than steam and nitrogen to the gasifying agent are shown in Appendix C. It is unlikely that gasifying agents other than oxygen, air, or steam or a combination of these would be used in a practical gasifier, so they are of academic interest only.

5.3.3 Results for an Air Gasifier

The results of the sensitivity analysis using air as the gasifying agent are shown in Appendix C (see Figures C5 to C12). The trends shown in this analysis are similar to those shown in the results for oxygen gasification. The main differences are that the gas products contain about 50% inerts, by volume. The total volume of gas produced is thus about twice that of the oxygen gasifier. Also, the volume of gasifying agent is higher than in the case of oxygen gasification. The higher gas flows in air gasification lead to a greater importance of sensible heat effects.

5.3.4 Discussion of Results of the Sensitivity Analyses

In all of the results of the sensitivity analyses, it may be seen that the carbon boundary gasification temperature is below that of a real gasifier. This implies that when using chemical equilibrium to model biomass gasification, the homogeneous gas phase equilibrium should be used, rather than the calculation which assumes the presence of solid char.

In order to understand the effects of the heat balance more easily, a modified version of "MODEL" was used. In this case, the heat balance was not used in the calculation, but an equilibrium calculation was performed using the carbon boundary at a range of temperatures. The flows of energy in the system and the gas composition were then plotted against temperature. The results are shown in Figures 5.12 and 5.13. It can be seen that the sensible heat of the outlet gas stream is small compared with the chemical energy. As a result, the energy effects of chemical reaction are much more important than the influence of gas sensible heat. The curve of outlet energy against temperature in the carbon boundary case can be seen to be steep around the point where the energy balance is satisfied. Thus, the carbon boundary temperature will not vary much with changing physical conditions, but a change in chemical composition may have a strong influence on carbon boundary temperature, as previously shown.

One interesting effect demonstrated in Figure 5.13 is that below 500K, the equilibrium gas is mainly H_2O and CO_2 with some CH_4, and above 1250K the gas is solely H_2 and CO.

Whilst it is usually possible to explain the results of these calculations, once they have been produced by the computer model, it is often difficult to predict intuitively the effects of changing the various parameters. This illustrates the value of using a computer model to look at the effects of design parameters.

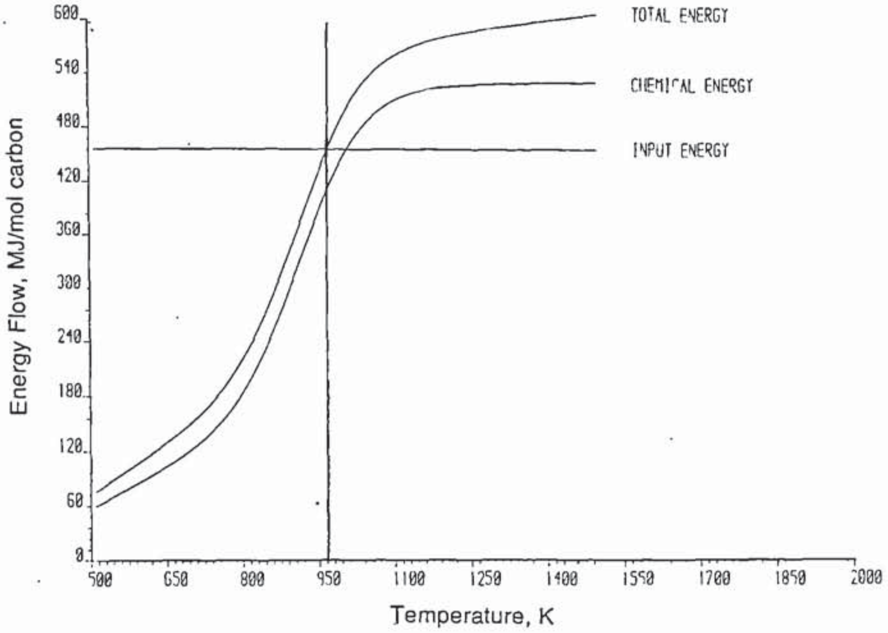


Figure 5.12 Graph to Show Variation of Exit Energies with Temperature, for an Idealised Biomass Gasifier

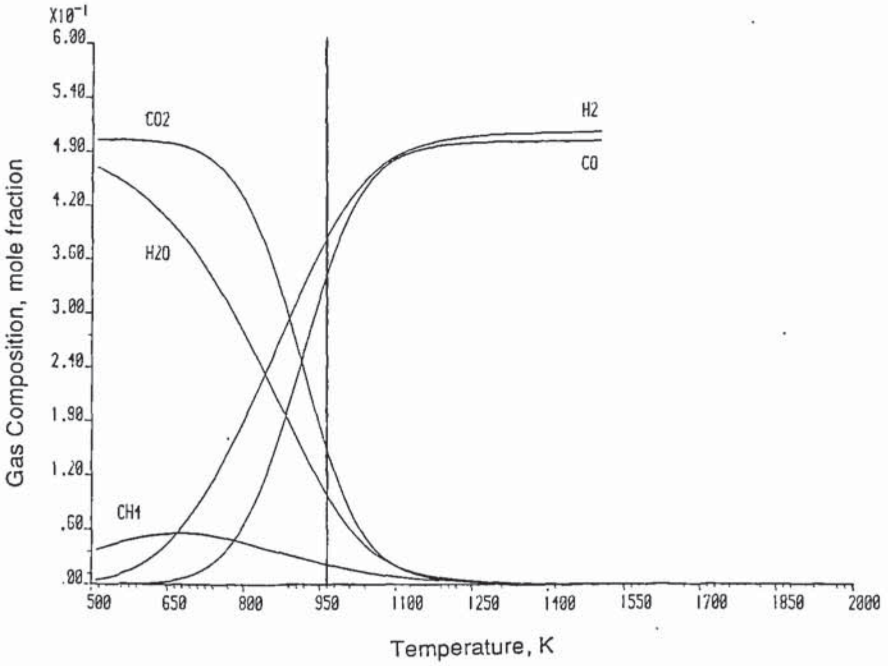


Figure 5.13 Graph to Show Variation of Exit Gas Composition with Temperature, for an Idealised Biomass Gasifier

5.4 COMPARISON OF REAL GASIFIER PERFORMANCE WITH RESULTS FROM THE MODEL

It was decided to compare the results of real gasifiers with results obtained using the equilibrium carbon boundary model: this is a method of comparing real results with ideality. In order to perform the comparison, the gasifying agent, pressure, feedstock composition, and performance of the real gasifier must be known. The feedstock, pressure and gasifying agent data can then be used in the program MODEL in order to calculate the idealised performance of the gasifier, which is then compared with the published gasifier performance.

One problem which was encountered in this work was the lack of comprehensive data on the performance of gasifiers. Although there are many designs of gasifier, much of the data is published in the form of "typical results", which are quoted without the operating conditions to which they apply, and which do not contain the full information required. This is especially true of commercial gasifier systems.

Table 5.3 shows the data used in this comparison between real and ideal gasifier performance. It was decided that the easiest way of making the comparison was to take a single figure measure of performance, which would allow the real performance of a gasifier to be plotted on a graph against the ideal performance. There are several measures of performance which could have been chosen for this purpose. These include various ratios of components of the product gas, such as the H_2:CO ratio or the CO:CO_2 ratio. The disadvantage of using the ratios mentioned is that they are measures of specific components of the gas, and not the product gas as a whole. Where the product gas is to be used as synthesis gas then the total H_2 + CO becomes important, but in the majority of cases, the quality of the gas as a fuel is more likely to be important. The gasifier thermal efficiency would be a useful measure of performance, but there are a number of ways of defining the efficiency, depending on whether the sensible heat of the outlet gas is included, and depending on whether the higher or lower heating values are used. This could easily lead to confusion, and incorrect comparisons of results. Also, if product gas sensible heat is included in the efficiency, it would be possible for a gasifier to produce a large volume of a hot gas, giving a high efficiency, but the gas would have a low calorific value, and thus a poor gas quality. The measure which was chosen was the calorific value of the product gas, because it is a direct measure of the product gas quality. Because of the high calorific value of methane and higher hydrocarbons, which are often present in higher quantities in real gasifiers than is predicted by equilibrium, the heating value can also indicate non-ideal behaviour in a simple manner.

The results of the comparison are shown in Figure 5.14. In addition to the data points from Table 5.3, three lines have been plotted. Two of these lines show the results of the model where the effects of heat loss has been included in the calculations, at levels of 10% and 20% of the chemical energy entering the gasifier in the feedstock. The third line represents the case where some of the pyrolysis products bypass the reaction zone of the gasifier, to appear directly in the product gas. This was calculated by taking 10% of the gas to be hydrocarbons from the pyrolysis process, adding this to the gas composition calculated from the model for gasifying the remainder, and finding the HHV of the resulting gas.

Table 5.3 Comparison of Real against Theoretical Gasifier Performance

	Gasifier	Feedstock	Gasifying agent	Actual HHV	Model HHV	Ref
1	Battelle PNL updraft	wood	O_2 / steam	10.62	10.19	77
2	Foster Wheeler fluid bed	refuse	air	2.67	4.89	78
3	IGT pressurised fluid bed	wood	O_2 / steam	7.07	12.50	79
4	Battelle PNL catalytic fluid bed	bagasse	steam	10.22	17.32	80
5	Univ. Miss.-Rolla fluid bed	wood	air	4.01	4.87	81
6	CPC fluid bed	wood	air	5.47	5.88	82
7	Maniatis fluid bed	wood	air	5.34	5.89	54
8	Goss dowdraft	wood	air	6.28	5.84	83
9	JBEC / Wellman Oxygen Donor	wood	O (from donor)	12.84	10.25	44
10	SERI downdraft	wood	O_2	10.15	10.64	84
11	Koppers-Totzek	coal	O_2 / steam	10.5	5.8	10
12	Open Univ. downdraft	carrot fibres	air	3.10	5.60	85
13	Imbert downdraft	wood	air	5.26	5.77	38
14	Lurgi fluid bed	wood	O_2 / steam	9.91	10.15	47
15	Creusot Loire	wood	O_2 / steam	9.48	10.43	86
16	Aston University downdraft	wood	air	4.42	5.94	87

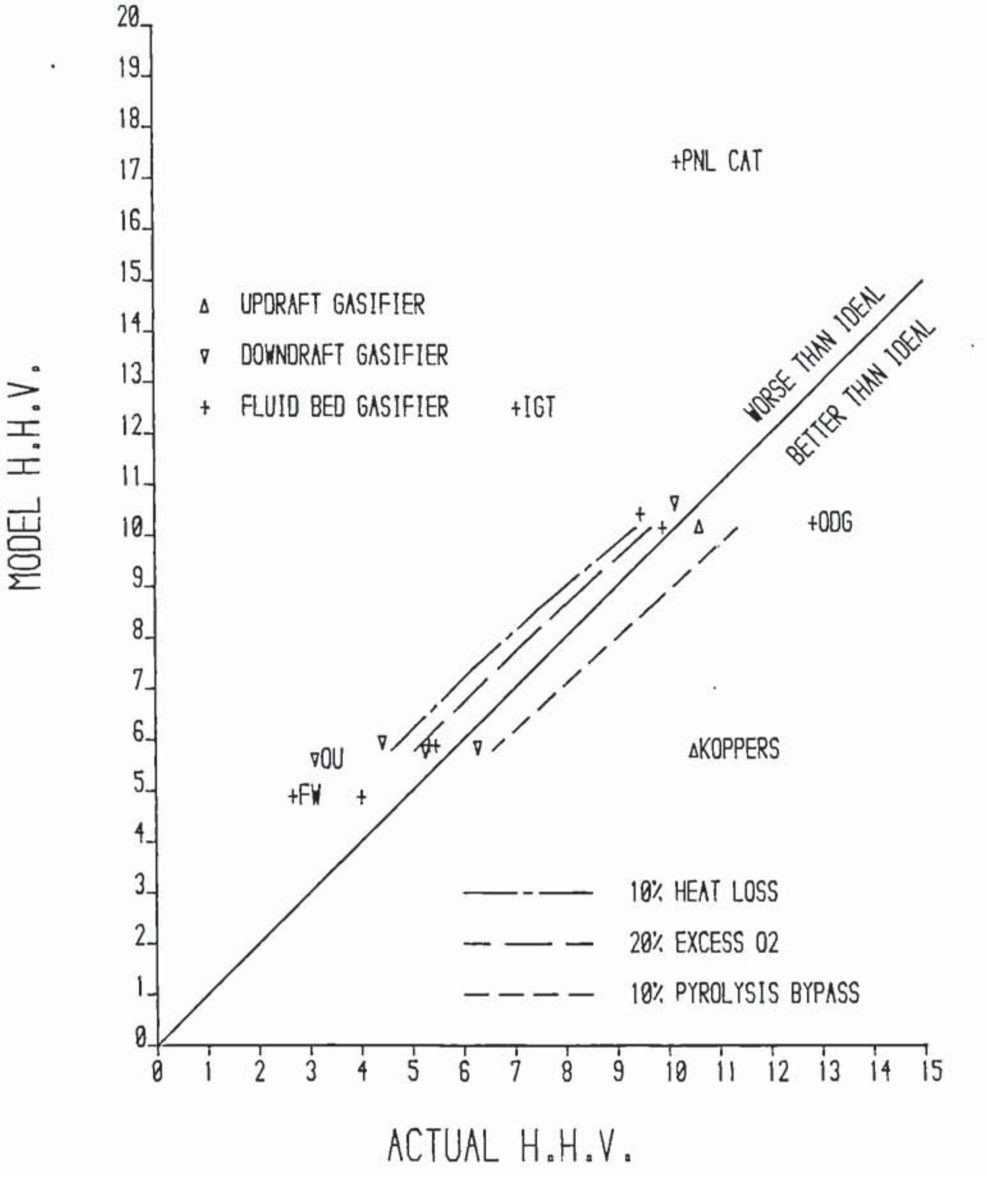


Figure 5.14 THEORETICAL GASIFIER PERFORMANCE AGAINST ACTUAL PERFORMANCE

On examining Figure 5.14, it can be seen that results are grouped in two clusters. This is because most gasifiers use either air, or a mixture of oxygen and steam as gasifying agent. Variations in parameters other than gasifying agent exert a small influence on the product gas calorific value compared with the influence of gasifying agent. Thus, all oxygen/steam gasifiers give fairly similar results, and air gasifiers give a separate group of similar results. There is no sufficiently detailed published data on gasification using oxygen-enriched air, so there are no figures in the region between the two groups of results.

Most of the results lie close to ideality, but show a gas composition that is slightly worse than ideality. (Ideality is shown by a 45° line in Figure 5.14). There are two main reasons for this: heat loss from the gasifier, and the fact that equilibrium carbon boundary gasification temperatures are lower than those experienced in gasification. The higher the temperature of the outlet gases, the more energy leaves the gasifier as sensible heat, and less energy is available as chemical energy in the product gas. Heat loss reduces the gas heating value by affecting the energy balance: the heat lost from the gasifier is no longer available to be included as chemical energy in the outlet gas stream. The higher temperatures of real gasifiers compared with the carbon boundary case (the difference is of the order of 200°C) affect the gas composition because of the greater amount of sensible heat in the outlet gas, reducing the energy which can leave as chemical energy.

The two effects which reduce the calorific value are offset by another effect which tends to give a higher gas heating value than ideality. This is the inclusion of non-equilibrium pyrolysis products in the gas produced by the gasifier. These are mainly methane and other hydrocarbons, which have a much higher calorific value than hydrogen or carbon monoxide, and which have a large influence on gas calorific values at low concentrations.

The gasifiers which stand out as being a long way from ideality are the Oxygen Donor Gasifier, the IGT catalytic steam fluid bed gasifier, the Battelle PNL catalytic steam fluid bed gasifier, the Koppers-Totzek gasifier, the Foster Wheeler fluid bed gasifier and the Open University downdraft gasifier.

The main reason for the Open University and Foster Wheeler gasifiers giving considerably worse quality product gas than predicted by ideality is probably that the design has not been developed sufficiently, and they have not reached the best achievable performance.

The Oxygen Donor Gasifier is of an unusual type, and it has been suggested that its mode of operation is actually as a twin bed pyrolyser / combustor. This mode of operation would tend to favour hydrocarbons over hydrogen and carbon dioxide, giving a product gas of higher calorific value than would be predicted by equilibrium calculations.

The Koppers-Totzek gasifier gives a higher quality gas than that predicted for two reasons: the coal used by this gasifier has a very high ash content, and the IGT equation used by the computer model to predict the feedstock heating value predicts a lower feedstock calorific value than the coal actually has, affecting the heat balance; and the gasifier has a low retention time for the product gases, thus leaving more pyrolysis products in the product.

The adiabatic equilibrium carbon boundary model of gasification is only really applicable to gasifiers where the gasifying agent contains oxygen. Thus, the IGT catalytic fluid bed steam gasifier and the Battelle

PNL steam gasifier give anomalous results. Steam gasification is endothermic, whereas oxygen and air gasification are both exothermic. Therefore an adiabatic model of steam gasification requires the use of a high temperature steam as gasifying agent, most of which passes through the reactor virtually unchanged, apart from giving up its sensible heat to provide energy for the endothermic reaction. The reaction temperatures predicted by an adiabatic model are very low - below the practical range of operation of a gasifier - and the product gas methane content predicted by the equilibrium calculation is thus high. As the model is not applicable in this case, no sensible conclusions can be drawn from this comparison.

It should be stressed that the carbon boundary model of gasification described in this chapter is an idealised model intended to provide a "target" against which the results of real gasifiers may be compared. If an equilibrium model is to be used to estimate the performance of a real gasifier, a fixed temperature equilibrium model of the type described in Section 5.2.1 should be used. Fixed temperature models are included within the program MODEL in addition to the carbon boundary model, as mentioned in Section 5.2.2.

Finally, it should be recognised that the main limitation of the carbon boundary approach is that no account is taken of kinetic and physical rate processes. To estimate spatial temperature and composition distributions within gasifiers, it is necessary to include such processes in a more detailed model of the actual reactor system, as discussed in Section 4.4.

Nomenclature, Chapter 5

$F_{i,j}$	Mass flow of component j in stream i
$G_{j,T}$	Standard Gibbs free energy of formation of component j at temperature T
K_i	Equilibrium constant of chemical reaction i
$n_{i,j}$	Molar flow of component j in stream i
P	Total pressure
p_j	Partial pressure of component j
T	Temperature of reaction
T_i	Temperature of stream i
x_j	Mole fraction of component j
$y_{i,j}$	Mole fraction of component j in stream i
ν_j	Stoichiometric coefficient of component j, positive for products, negative for reactants

CHAPTER 6
MODELLING OF FLUID BED GASIFIERS

6.1 INTRODUCTION

The equilibrium models of gasification described in Chapter 5 are straightforward and present an idealised picture of gasifier behaviour. Whilst the results often approximate to those of real gasifiers, there are also differences between the performance of the real gasifiers and the results of equilibrium models. This chapter describes models of fluid bed gasification which are based on equilibrium modelling, but which incorporate the effects of pyrolysis on gasifier performance. There are several types of fluid bed gasifiers, from single bed gasifiers fluidised by the gasifying agent to complex twin bed designs, as described in Chapter 3. The models described in this chapter are based on simple fluidised bed gasification: a single fluidised bed using air or oxygen mixed with some steam as the gasifying agent.

6.2 PERFORMANCE OF FLUID BED GASIFIERS

The main difference between the gas composition of real fluid bed gasifiers and that predicted by thermodynamic models is that the mole fraction of methane and higher hydrocarbons is much greater in the real gas composition. Table 6.1 compares the results obtained from real gasifiers with results obtained by using the same operating conditions and temperature in the program "MODEL" working in fixed temperature mode, as described in Chapter 5.

Table 6.1 Comparison of actual fluid bed gasifier results with those predicted by an equilibrium model

Gasifier [a]	Gas	Dry gas composition, mol %						Ref.
	Yield [b]	H_2	CO	CO_2	CH_4	HCs[c]	N_2	
Framatome	1.35	30.9	41.1	22.8	0.6	0.0	4.5	86
	1.34	*32.9*	*45.1*	*22.0*	*0.0*	*0.0*	*0.0*	
Lurgi	1.40	32.3	32.4	25.7	4.7	1.3	3.3	47
	1.56	*41.9*	*35.2*	*22.7*	*0.2*	*0.0*	*0.0*	
IGT Renugas	1.70	10.1	7.2	16.2	24.4	1.0	41.1	88
	3.44	*18.0*	*7.8*	*18.4*	*0.0*	*0.0*	*55.8*	
Process Comb.	-	4.2	17.1	14.4	4.3	2.9	57.1	89
	3.03	*22.1*	*13.4*	*16.5*	*0.1*	*0.0*	*48.0*	
Biosyn [d]	-	9.7	13.0	-	6.8	1.9	68.7	90
	2.97	*18.6*	*16.9*	-	*0.0*	*0.0*	*64.5*	

Notes

a For each gasifier, the first line in plain text gives the actual results, and the second line in *italics* are those predicted by an equilibrium model calculating the gas composition at the working temperature of the actual gasifier

b Gas yield is expressed as Nm^3 of dry gas per kg dry ash free biomass feed

c C_2H_4, C_2H_6 and higher hydrocarbons, quoted as C_2H_4

d The CO_2 and N_2 are quoted as total inerts, in the N_2 column of this table

As the formation of methane and hydrocarbons is not favoured by thermodynamic considerations, they must be either products of pyrolysis or derived from the products of pyrolysis by thermal cracking. The model discussed in this chapter takes pyrolysis products into account in the calculations of gasifier performance.

6.3 STAGEWISE MODELLING OF GASIFIERS

6.3.1 Introduction

The reactions experienced by a particle of feed entering a gasifier can be described in stages - drying, pyrolysis, gasification, and gas phase reactions of the pyrolysis tar. One approach to modelling a gasifier is, therefore, to divide the gasifier into these stages, each stage being modelled by a black box model, based on a mass balance and equations or algorithms to calculate the outlet products. Energy balances on each black box are not appropriate in the case of a fluid bed gasifier because, although the fluid bed is conceptually divided into a number of reaction zones, in actual fact each stage is taking place throughout the bed and heat will be freely transferred between the stages. The only energy balance which is applicable is the overall energy balance. The black box models can be linked so that a model of the whole gasification process is obtained, and an accurate prediction of gasifier performance is produced.

The black boxes used in a stagewise model do not necessarily have to represent the actual reactions which take place in the gasifier. A simple form of stagewise model may be envisaged which has two stages: one which produces the non-equilibrium products of gasification such as CH_4 and hydrocarbons, and a second which takes the remaining material from the feedstock and reacts it with the gasifying agent using an equilibrium method to calculate gas composition. In effect, this is an equilibrium model of gasification with minor empirical corrections to account for non-equilibrium products. This type of model was described by Gibbins and Wilson.[76]

Shand and Bridgwater modelled fixed bed gasifiers of both updraft and downdraft types using a stagewise approach.[6,7,8] Their models used empirical correlations for the pyrolysis modelling, with equilibrium modelling of the char gasification. Kosky and Floess[91] modelled updraft gasification of coal in a similar manner. There is a conceptual difference between these stagewise models of fixed-bed gasifiers and those for fluid bed gasifiers of the author: in the fixed bed gasifiers the individual black-box models represent physical zones of the reacting bed, whereas in the fluid bed model the individual steps are separated for modelling purposes only, with each black box representing processes which take place throughout the bed.

6.3.2 Stagewise Modelling of Fluidised Bed Gasifiers

One approach to stagewise modelling is to consider the reaction stages of drying, pyrolysis, tar cracking and gasification, as mentioned in Section 6.3.1. A general model of a fluid bed gasifier which uses these stages is shown in Figure 6.1. Solid biomass enters the gasifier and first undergoes drying. The dry biomass produced passes to the pyrolysis stage, whilst the water vapour produced splits in two, a fraction F1 passing to the gasification stage and the remainder bypassing the gasification reactions and passing directly into the product gas stream. The dry biomass is pyrolysed and the char passes to the gasification stage. The volatile tars, liquids and gases split into three, fraction F2 passing to the gasification stage, fraction F3 passing to the tar cracking stage, and the remainder passing into the product gas stream. The gaseous products from both the gasification stage and the tar cracking stage pass into the product gas, whilst the solid ash and char residue from gasification passes out of the gasifier in a solid residue stream. The final product gas stream is a mixture of products from all four stages of the model. The fractions F1, F2 and F3 will depend on the geometry and mixing within the gasifier.

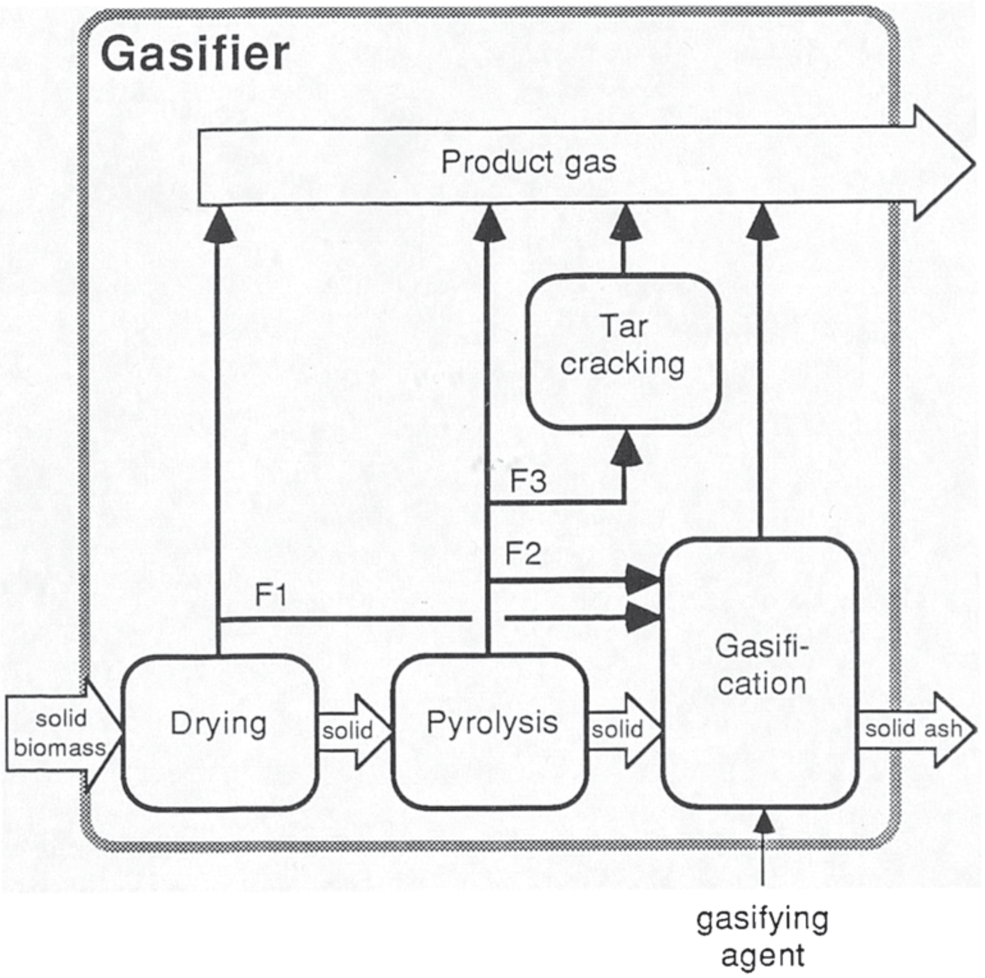


Figure 6.1 A General Stagewise Model of a Fluid Bed Biomass Gasifier

This generalised stagewise model may be simplified in a number of ways. The fractions F1 and F2 may be considered to be equal, as the fraction of water vapour from drying which bypasses the gasification reaction and passes directly to the outlet gas stream should be the same as the fraction of pyrolysis products bypassing gasification. Alternatively, the number of stages may be reduced. One simple reduction would be to combine the drying and pyrolysis. These are unlikely to take place discreetly bearing in mind that the outside of a particle may be pyrolysing while the inside is still drying. Also, there may be interactions between the drying and pyrolysis processes.

6.3.3 Pyrolysis and Cracking in Stagewise Modelling

It was stated in Chapter 2 that pyrolysis is a complex process which is a network of reaction pathways. It was also stated that the products of pyrolysis are highly dependent on reaction conditions and that a good model to predict the products of pyrolysis does not exist. Empirical models of pyrolysis must, therefore, be used in stagewise models of gasification. These empirical models must be based on data from experiments with similar conditions to those which apply in the gasifier. Obviously, an inert atmosphere must be used, rather than gasifying agent, but the temperature, pressure, particle size and residence time of products should be the same.

Tar cracking is a more difficult problem. The boundaries between tar cracking and secondary pyrolysis in gas phase are difficult to draw. In some ways, the easiest approach is to consider tar cracking to be part of the pyrolysis process. The conditions in a well-mixed fluid bed are constant, so pyrolysis data obtained using similar temperature, pressure and residence time should reflect the same tar cracking reactions as those which take place in a fluid bed reactor, providing the gasifying agent plays no part in the tar cracking process. However, the conditions through which the pyrolysis products pass in fixed bed types of reactor are more complex and this approach would be more difficult to use.

6.4 THE STAGEWISE COMPUTER MODEL

The approach adopted in constructing a computer program to perform stagewise modelling of gasification was to divide the program into modules which were as independent as possible. The main units of the program were the initialisation and data input subroutines, a subroutine to perform drying, pyrolysis and tar cracking calculations, a gasification subroutine connected with an iterative heat balance subroutine, and a subroutine to print the results of the calculations. Some data within the program are needed by many or all of the subroutines: these include the gasification conditions and the compositions and flows of the conceptual streams between the stages of the calculation. These data were transferred between subroutines by storing them in common blocks which could be accessed by any subroutine which required them.

It was decided to construct a program which could be used to model both simple and complex stagewise processes. The following approach was adopted for the pyrolysis stage: drying, pyrolysis and tar cracking were all combined into one stage of the model, which had only two outlets - one a bypass stream which mixed directly with the product gas stream, and another the residue stream which passed to the gasification stage. The split and composition of the products from the first stage are read directly from a pyrolysis data file. The first stages of the model are calculated by hand and the results are then loaded into a datafile where they can be used by the gasifier computer model. It would be possible to construct computer programs to generate the pyrolysis data files used by the first stage of the gasifier model, if this was required, provided that a suitable pyrolysis and tar cracking model was available. Different models of pyrolysis and tar cracking may, therefore, be used in the same program, making it very flexible in use. A conceptual diagram of this model is shown as Figure 6.2

Figure 6.3 is a representation of the structure of the computer model. Data are entered interactively and the data entry subroutine stores these in the common blocks. The pyrolysis subroutine then uses the input data and pyrolysis data from a previously created datafile to calculate the bypass products passing into the product gas and the residual material passing to gasification. A suite of subroutines then performs the gasifier calculations. The equilibrium calculations are simplified by the exclusion of methane, the only reaction considered being the homogeneous water gas shift reaction:

$$CO + H_2O = CO_2 + H_2 \qquad [6.1]$$

The gasifier working temperature is specified as part of the input data, so the equilibrium constant can easily be calculated using equation [5.9] from Section 5.1. Note that this equilibrium calculation is **not** based on the carbon boundary equilibrium calculation described in Chapter 5.

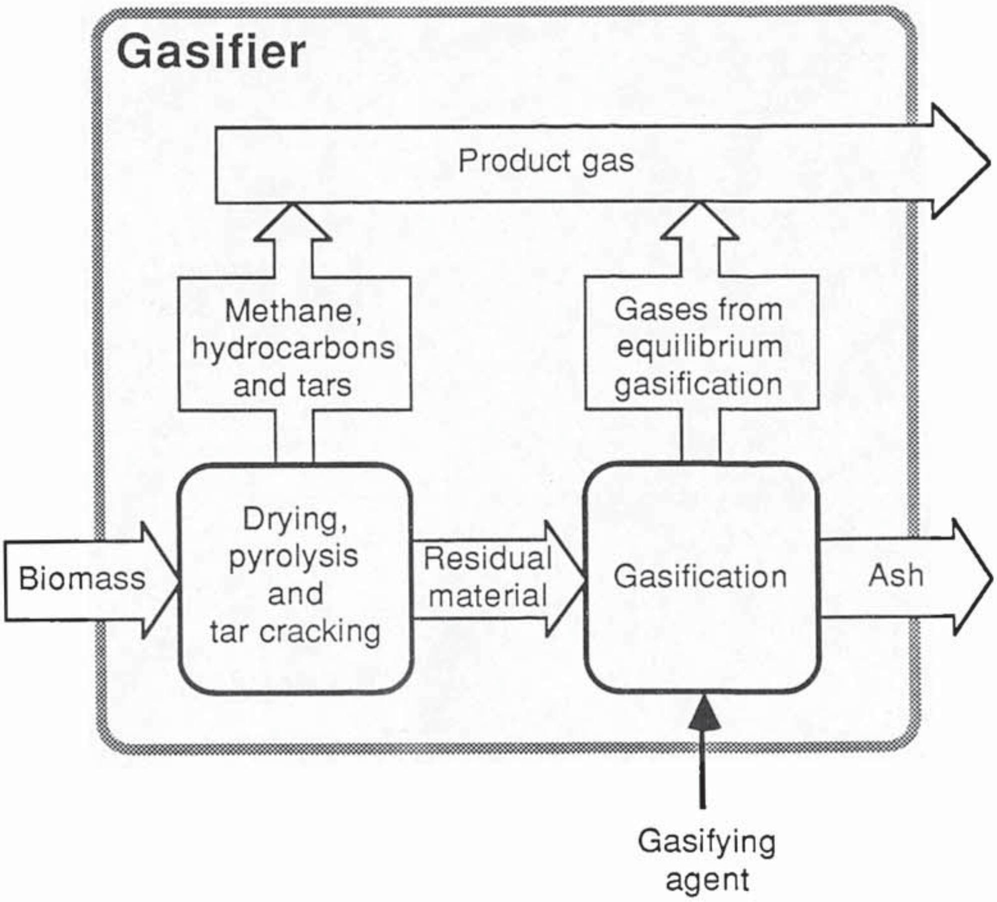


Figure 6.2 Conceptual Diagram of the Stagewise Gasification Model

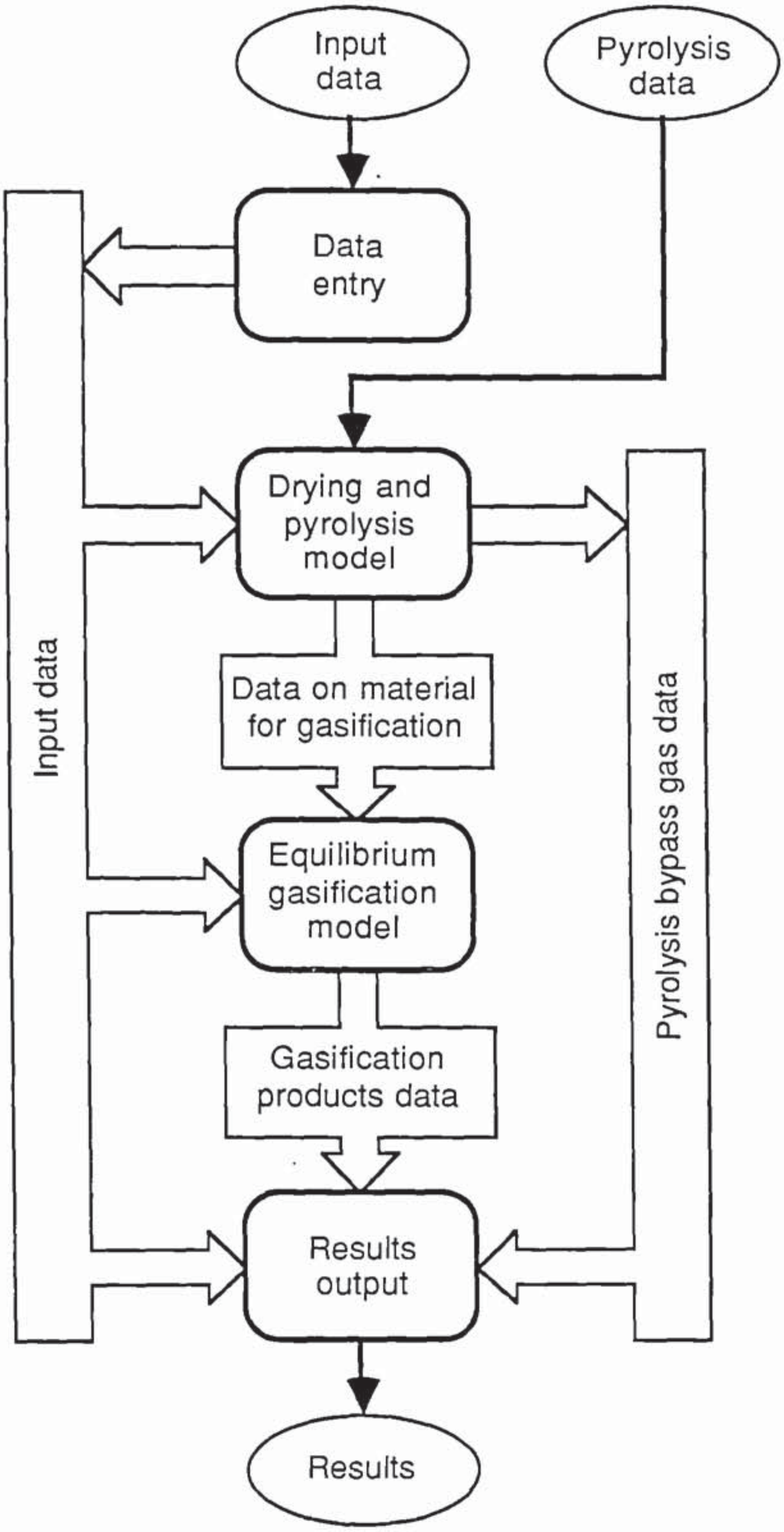


Figure 6.3 Structure and Data Flows in the Stagewise Gasifier Computer Model

6.5 RESULTS OF THE STAGEWISE MODEL

6.5.1 Selection of Pyrolysis Data

Whilst there is an extensive literature about pyrolysis, much of the reported information on product yield and composition is incomplete and the material balances of pyrolysis are not closed. This may be because researchers who generated the data have been interested in one specific product from pyrolysis, for example charcoal, and the other products are often considered as unwanted byproducts which can be disposed of as fuel for the process. However, when incorporating pyrolysis into a gasification model, all products must be accounted for and the elemental balances must be closed.

A second problem in choosing pyrolysis data for the modelling of gasification is to find data from experiments carried out at a similar temperature, pressure and residence time to that of the fluid bed gasifier. Fluid bed gasifiers usually operate at a temperature between 700 and 900 °C, but most work on pyrolysis has been carried out at lower temperatures, often 250 to 600 °C.

The particular data chosen for the author's work were selected because the conditions under which the experiments were performed were deliberately chosen to reproduce the conditions encountered in fluidised bed gasification. The pyrolysis data are shown in Table 6.2. From the raw data, a material balance has been constructed and this is also shown in Table 6.2. It can be seen that the composition of the residue calculated by closing the elemental balances is $C_1H_{0.079}O_{0.011}$. Both the hydrogen and oxygen contents of this residue are much lower than those of pyrolysis tars (which have compositions ranging from about

Table 6.2 Pyrolysis Data Used in the Stagewise Model
(from Belleville and Capart [56])

Reaction Conditions

Wood slabs of 7mm thickness (ie similar dimensions to wood chips)

Moisture, wet basis	22.2 %
Temperature	780 °C
Atmosphere	Nitrogen

Component	**Yield per kg dry ash free wood**
Gases	0.608 Nm^3 = 0.521 kg
Residues (char + tar)	0.282 kg
Water (by difference)	0.197 kg

Gas Composition, mol %	
H_2	34.2
CO	39.0
CO_2	10.5
CH_4	13.4
C_2H_4	2.9

Material Balance

Basis: 1 kg dry ash free wood

Input	**total kg**	**total mol**	**mol C**	**mol H**	**mol O**
Biomass	1.000	-	41.67	60.00	27.50
Outputs					
H_2	0.019	9.28	0.00	18.57	0.00
CO	0.297	10.59	10.59	0.00	10.59
CO_2	0.125	2.85	2.85	0.0	5.70
CH_4	0.058	3.64	3.64	14.55	0.00
C_2H_4	0.022	0.79	1.57	3.15	0.00
H_2O	0.197	10.96	0.00	21.92	10.96
Residue §	0.282	-	23.02	1.81	0.25
Total Outputs	1.000		41.67	60.00	27.50

§ Composition by difference = $C_1H_{0.079}O_{0.011}$

$C_1H_{1.45}O_{0.6}$ [92] to $C_1H_{0.81}O_{0.07}$ [31]) and those often quoted for charcoal ($C_1H_{0.46}O_{0.11}$, see Table 2.1). Yet it has been inferred that the residue is entirely char; the effect of this assumption is tested in Section 6.5.4.

6.5.2 Evaluation of the model

In order to evaluate the model, two scenarios of pyrolysis were used at a standard set of reaction conditions; the results were compared with the results of the same program excluding pyrolysis (Model 1 in Table 6.4). The first pyrolysis scenario is that all of the gases and moisture from the pyrolysis pass to the outlet stream without reacting further and only the char residue from pyrolysis passes to the equilibrium gasification model (Model 2 in Table 6.4). In the second scenario, the components which take part in the water-gas shift reaction pass to the equilibrium model with only the methane and hydrocarbons passing directly to the reactor outlet without undergoing reaction (Model 3 in Table 6.4). The other reaction conditions are shown in Table 6.3.

The results from these models are compared with real gasifier results in Table 6.4. Some of the results from Table 6.1 cannot be used in this comparison, because their operating conditions are unlike those used in the models.

Table 6.3 Base Case Reaction Conditions

Temperature	780 °C (to match pyrolysis data)
Pressure	1 bar absolute
Gasifying agent	Oxygen
Wood composition	$C_1H_{1.44}O_{0.66}$
Moisture content	20 % dry basis, = 16.7 % wet basis
Gasifying agent temperature	25 °C
Heat Loss	5 % biomass calorific value

Table 6.4 Comparison of Actual Fluid Bed Gasifier Results with Results of the Stagewise Model

Case	Gas Yield [a]	H_2	CO	CO_2	CH_4	HCs[b]	N_2	Ref.
Oxygen Gasification								
Model 1 [c]	1.57	40.8	39.4	19.8	0.0	0.0	0.0	
Model 2 [d]	1.14	19.7	58.9	12.7	7.1	1.6	0.0	
Model 3 [e]	1.37	33.0	35.0	24.7	6.0	1.3	0.0	
Lurgi [f]	1.36	33.5	33.6	26.7	4.9	1.3	0.0	47
Mino	-	18.7	35.5	30.3	10.0	1.8	0.0	93
Air Gasification								
Model 1 [c]	3.01	19.0	18.1	12.8	0.0	0.0	50.0	
Model 2 [d]	2.48	8.9	22.3	10.6	3.3	0.7	54.2	
Model 3 [e]	2.59	15.2	15.9	15.6	3.1	0.7	49.6	
Twente Univ.	-	14.4	18.0	16.7	3.6	0.7	46.5	33
Process Comb.	-	4.2	17.1	14.4	4.3	2.9	57.1	89

Dry gas composition, mol %

Notes

a Gas yield is expressed as Nm^3 of dry gas per kg dry ash free biomass feed

b C_2H_4, C_2H_6 and higher hydrocarbons

c Model 1 is the basic stagewise model with no pyrolysis included, ie a simple equilibrium model

d Model 2 is the model where all of the pyrolysis products pass to the outlet gas stream without further reaction

e Model 3 is the model where only the methane and hydrocarbons pass directly to the outlet gas stream, with other pyrolysis products passing to the equilibrium model

f Adjusted to a nitrogen-free basis

It can be seen that Model 3 gives good predictions of the gas compositions of the Lurgi and Twente University gasifiers. However, the results of the Mino and Process Combustion gasifiers give results which are closer to the results of Model 2.

As the results from Model 2 match the results of the Mino and Process Combustion gasifiers with a fair fit, it can be inferred that in these gasifiers the pyrolysis products bypass the char bed and pass directly into the outlet stream. This is likely to occur where the fresh biomass is fed onto the surface of the fluid bed and where there is little mixing of pyrolysing biomass into the bed. It is notable that the Mino gasifier is fed onto the top of the bed, and was designed for a high loading of pyrolysis products in the outlet gas. [94] The methane content of the product gas from these two gasifiers is higher than that predicted by the model. This might be explained in two ways: either less product gas is produced from the gasification reaction than estimated by the model giving less dilution of the pyrolysis products, or the pyrolysis data used in the model is inaccurate for the conditions applying in these two gasifiers.

Another interesting observation is that Model 2 predicts carbon dioxide concentrations below those of the Mino and Process Combustion gasifiers. The explanation may be that the gasifiers are imperfectly mixed beds in which a pyrolysis process is taking place in parallel with both combustion and char gasification, with little reaction between the product gases. In these cases, the non-equilibrium models of Belleville and Capart [56] might be more appropriate than a model incorporating equilibrium.

Model 3 predicts well the results of the gasifiers of Lurgi and Twente University of Technology. In this model, the hydrocarbon products of pyrolysis passed straight to the outlet gas stream, whereas the other components passed to the equilibrium model. The close fit of the model to actual results may be explained by considering the kinetics of the reactions involved. If the reactions of methane and other hydrocarbons were slow compared with the attainment of equilibrium of the water-gas shift reaction, then the short residence times of gases in the fluid bed reactor might allow attainment of water-gas shift equilibrium while giving insufficient time for the methane and hydrocarbons to react. However, it is difficult to substantiate this theory, as the rate of reaction of the water-gas shift reaction is uncertain and is changed by several orders of magnitude depending on whether catalysis is involved. The reaction velocities of the water-gas shift equilibrium and the attainment of equilibrium has been discussed by Schoeters. [60]

It can be seen in Table 6.4 that Model 3 predicts a higher methane content than is found in the Lurgi gasifier. This can be explained by the configuration of the Lurgi gasifier, which is a circulating fluid bed, which allows recycle of some of the product gases. This gives a longer residence time for the gases in the reactor, allowing some reaction of the methane with gasifying agent.

It is interesting to compare the results of the three models. In the equilibrium model (Model 1) and the model where most of the pyrolysis products pass to the equilibrium calculation (Model 3) the H_2 : CO ratio is approximately equal to 1, but Model 2 predicts a ratio of about 0.35. The difference in the ratios of H_2 to CO between Models 2 and 3 is more marked than the difference between the predicted methane concentrations of the two models. This suggests that the two models could be used in conjunction to analyse the processes taking place in a real gasifier. The two models could be run at the same reaction conditions as the real gasifier, and the H_2 : CO ratio compared. If the H_2 : CO ratio of the real gasifier were similar to the ratio predicted by Model 2, then poor mixing of pyrolysis products within the reactor is probable. If Model 3 gives a better prediction, then the reactor is likely to be well mixed.

The low H_2 : CO ratio predicted by Model 2 can be explained by considering the conditions of the equilibrium stage of the model. As the biomass moisture and much of the hydrogen content of the dry biomass material pass straight to the outlet, there is little hydrogen in the equilibrium calculation. As a result, the equilibrium gasification produces only carbon oxides, and much of the hydrogen is lost in the form of steam in the product gas, as shown in Table 6.5. The small amount of hydrogen in the equilibrium stage means that there is little steam, and hence little shifting of CO and steam to CO_2 and H_2, causing a low CO_2 concentration in the product gas.

Model 2 gives the lowest estimated production of gas, but the highest heating value of gas. The total chemical energy of product gas predicted by the three models is similar, so there is a trade-off between gas volume and gas heating value (see Table 6.5).

Table 6.5 Further Comparison of the Three Models

Model	**HHV MJ / Nm³**	**Gas Yield Nm³**	**Chemical Energy MJ**	**H_2O Yield kg**
Model 1	10.19	1.58	16.1	0.22
Model 2	13.77	1.14	15.7	0.40
Model 3	11.83	1.37	16.2	0.22

Note: all yields are based on 1 kg dry ash free biomass feed.

Of the three models, Model 3 is probably the most useful, as fluid bed gasifiers often exhibit H_2:CO ratios of the order of 1, and methane concentrations of about 5 to 10 %, as predicted by Model 3. [4]

6.5.3 Comparison with the Carbon Boundary Model

Figure 6.4 compares the results of the three stagewise models with the adiabatic carbon boundary model described in Chapter 5. The equilibrium model operating at fixed temperature (Model 1) predicts gas heating values lower than the idealised model. This is to be expected as, to raise the temperature of the gas from the carbon boundary temperature to the stagewise model temperature, some of the combustible components of the product gas must be burned, giving a reduction in the gas heating value.

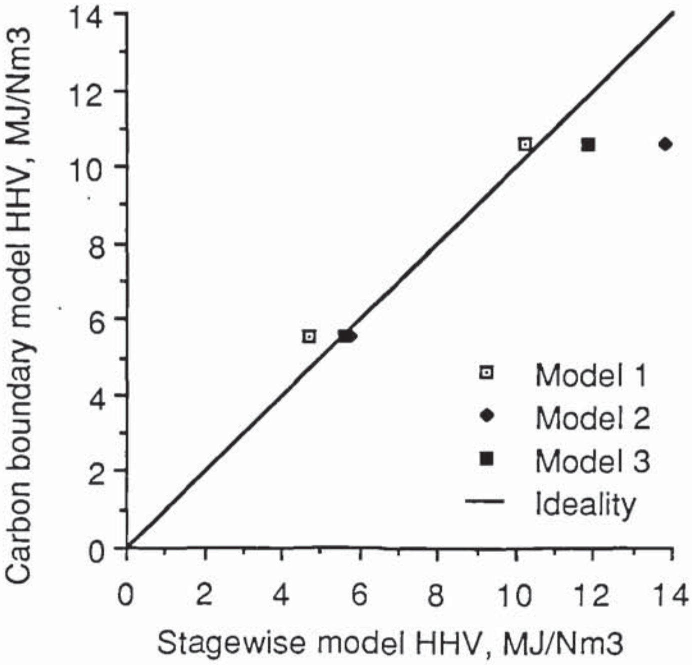


Figure 6.4 Comparison of the Three Stagewise Models with the Idealised Adiabatic Carbon Boundary Model

As would be expected, models 2 and 3 give heating values above those of the idealised case. This is due to the methane and higher hydrocarbons present in the product gas, which more than counteract the effects of the extra gasifying agent required to raise the temperature above the carbon boundary temperature.

6.5.4 The Effect of Tar in Pyrolysis Products

The pyrolysis data used in Model 2 and Model 3 assumed that all of the residue from the pyrolysis process was tar, based on the mass balance information used. However, some tars may be produced in high temperature pyrolysis [33], and it was decided to test the effect of tar production on the results of the model. The pyrolysis bypass data used in Model 3 was modified so that tar passed to the product gas at a mass yield of 5% (Model 4). The results of this modified model are compared with Model 3 in Table 6.6.

The tar yield predicted by this modified model is 5%. However, reported data on the tar content of fluid bed gasifier product gases is in the range 2 000 to 10 000 mg/Nm^3 product gas, as shown in Figure 3.11 of Chapter 3. This corresponds to a tar yield of about 0.25 to 1.3 %, so Model 4 greatly exaggerates the effect of tar loss in the product gas. A further model was therefore constructed, where there was still a 5% production of pyrolysis tars but where most of the tar was cracked to methane giving a 1% tar yield overall (Model 5 in Table 6.6). The idea of tar being cracked under the conditions occurring in a gasifier is not unreasonable. Considering analogous reactions, naphtha may be pyrolysed at 600 to 900 °C at residence times of about 0.3 seconds [95] and crude oil has been cracked in steam at temperatures of 700 to

900 °C at residence times of under 1 second. [96]

Tar has a high heating value, so the production of comparatively little affects the product gas composition considerably, as shown by the results of Model 4: the gas yield is reduced, and the CO_2 content increased, giving poorer gas quality. Model 5, showing the effect of tar cracking, gives a possible explanation of the high methane production of some gasifiers (for example the Mino gasifier shown in Table 6.4). The methane production of these gasifiers is higher than would be expected from the inclusion of methane from the primary pyrolysis reaction, as shown in Section 6.5.2. However, if tar were produced in the primary pyrolysis and cracking of this tar occurred, then the extra methane production would be accounted for.

Table 6.6 Effect of Tar Production on Predicted Gasifier Performance

Case	Gas Yield [a]	Dry gas composition, mol %						Tar Yield [c]
		H_2	CO	CO_2	CH_4	HCs [b]	N_2	
Model 3	1.37	33.0	35.0	24.7	6.0	1.3	0.0	0.00
Model 4	1.22	31.6	31.4	28.9	6.7	1.4	0.0	0.05
Model 5	1.31	31.6	33.0	26.2	7.2	1.3	0.0	0.01

Notes

a Gas yield is expressed as Nm^3 of dry gas per kg dry ash free biomass feed

b C_2H_4, C_2H_6 and higher hydrocarbons

c Tar yield is expressed as kg tar per kg dry ash free biomass feed

6.5.5 Sensitivity Analysis

Model 3 was used in a study of the effects of various operating parameters on fluid bed gasifier performance. Figures 6.5 to 6.9 show results of this sensitivity analysis. The base case used was as shown in Table 6.3 above. The results must be treated with some caution, as the pyrolysis data used was the single set of figures of Table 6.2. In particular, the temperature and biomass moisture content could affect the pyrolysis, which would also affect the results of the gasification model. Ideally, pyrolysis data for each set of operating conditions should have been used but these were unavailable.

The parameters tested were gasifier temperature, gasifying agent air content, gasifying agent steam content, biomass moisture content, and gasifier heat loss. Pressure has no effect on the results of the model. This is because the equilibrium calculation used the homogeneous water-gas shift reaction only, upon which pressure has no effect if gas ideality is assumed. In real gasifiers, pressure has an effect on gasifier performance, affecting the pyrolysis step and the rate of decomposition of pyrolysis products. If suitable pyrolysis data were available, the effect of pressure could be tested using the model.

Temperature is predicted to have a limited effect on the gas composition (Figure 6.5). The increased gasifying agent required to boost the temperature will decrease the H : O ratio in the system, producing CO at the expense of H_2. However, in real gasifiers there may be other effects due to the increasing reaction rate of methane reforming, and changes in the pyrolysis product yields and composition.

The addition of an inert gas to the system by mixing air into the gasifying agent has the effect of diluting the product gas, as shown in

Figure 6.6. The CO_2 concentration is not reduced in proportion with the CO and H_2, because more product gas must be burned to provide the sensible heat of the inert gas passing through the system.

Adding steam to the gasifying agent (Figure 6.7) has the effect of changing the C : H : O ratios, and also removes energy because of the endothermic reactions of steam with char. The net effect is to reduce CO at the expense of H_2 and CO_2, and the gas heating value is reduced. These results indicate that the H_2:CO ratio could be tailored to suit a downstream synthesis process by varying the steam : oxygen ratio of the gasifying agent.

Biomass moisture content (Figure 6.8) has similar effects to gasifying agent steam content, as might be expected. However, in a real gasifier there will also be effects on the pyrolysis step of the process. These would be predicted by the model if pyrolysis data were available at a range of moisture contents.

Figure 6.9 shows the effect of the heat loss from the gasifier, expressed as a percentage of the heating value of the biomass supplied to the gasifier. If heat is lost from the gasifier system, more oxygen must be supplied. This reacts with the product gas, increasing the CO_2 at the expense of combustible gases and reducing the gas heating value.

The results of this sensitivity analysis show the same basic trends exhibited by the results of the adiabatic carbon boundary model reported in Chapter 5, the main differences being due to the stagewise model's operation at a fixed temperature compared with the floating temperature of the carbon boundary model.

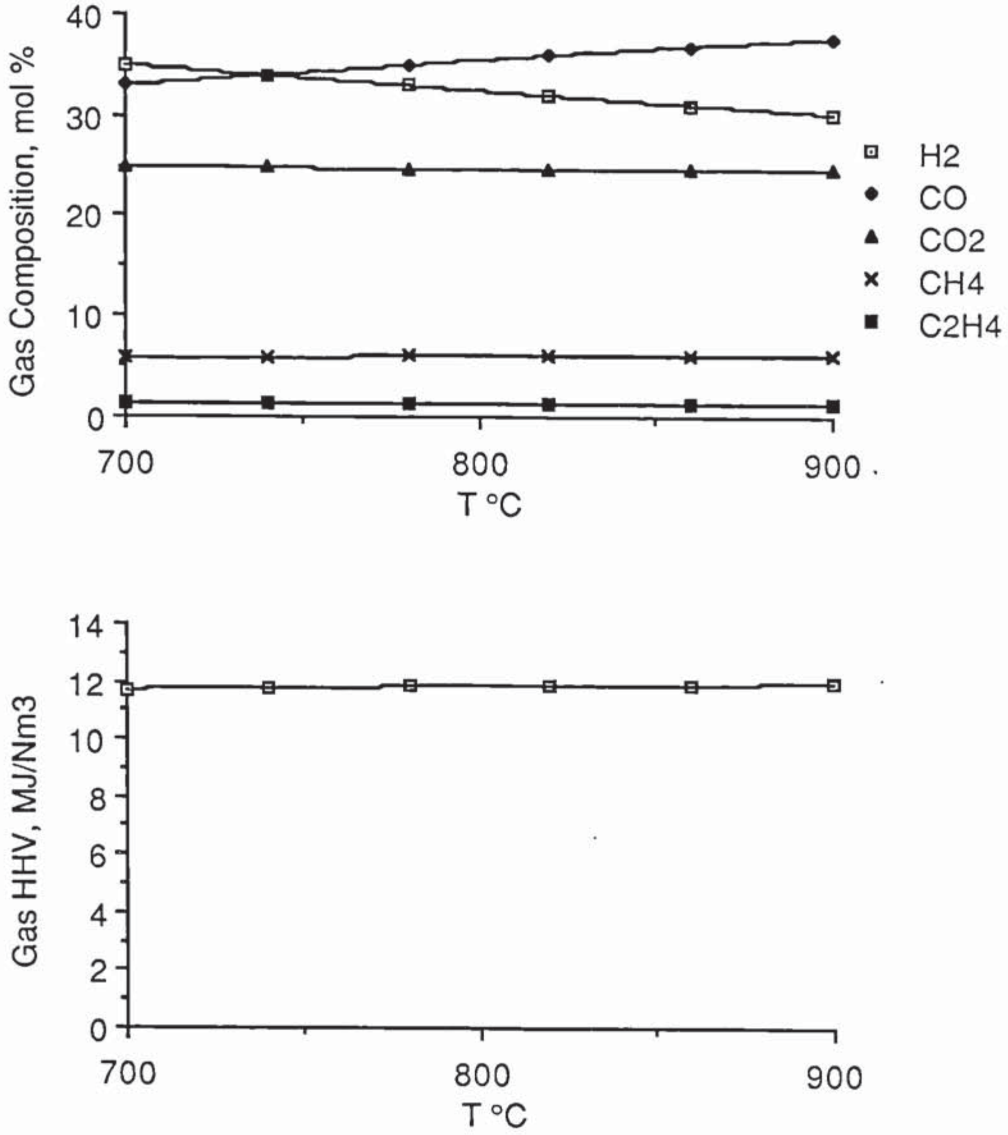


Figure 6.5 Predicted Gasifier Performance as a Function of the Gasifier Temperature

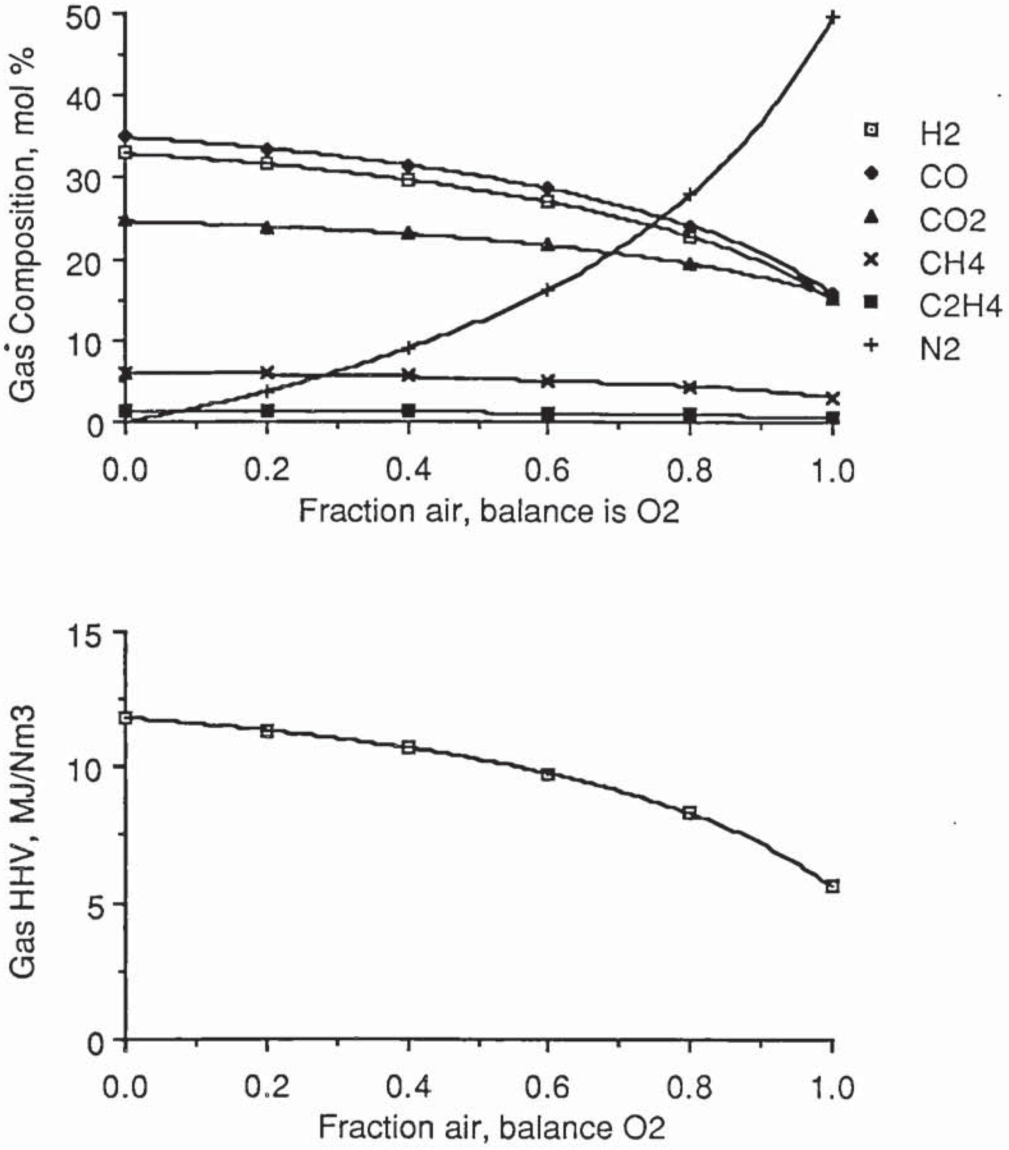
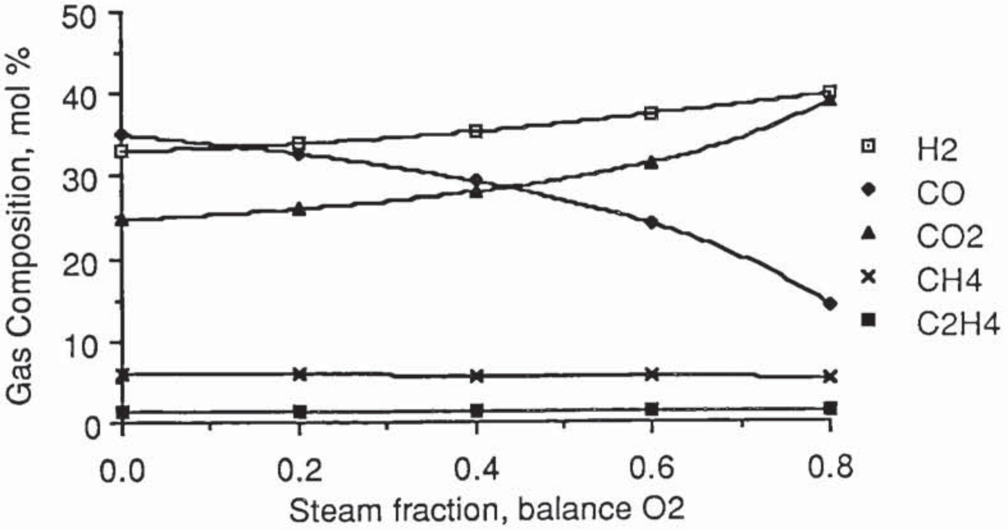


Figure 6.6 Predicted Gasifier Performance as a Function of the Gasifying Agent Air Volume Fraction



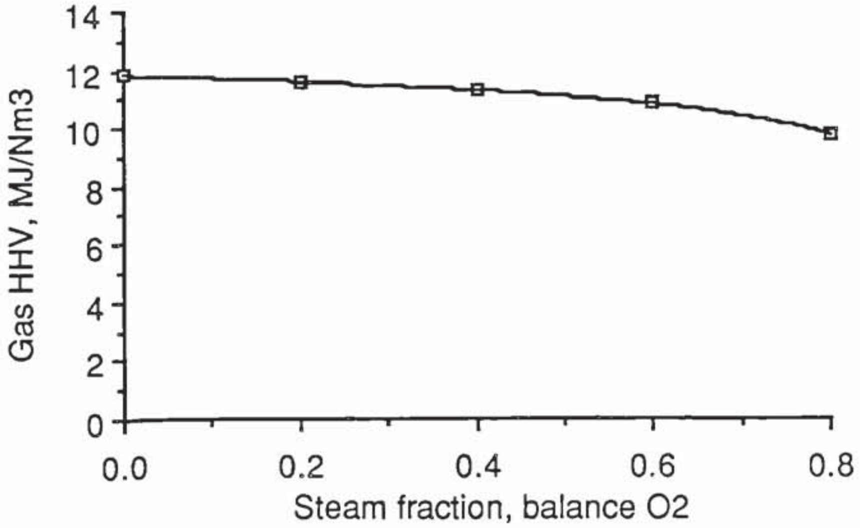


Figure 6.7 Predicted Gasifier Performance as a Function of the Gasifying Agent Steam Volume Fraction

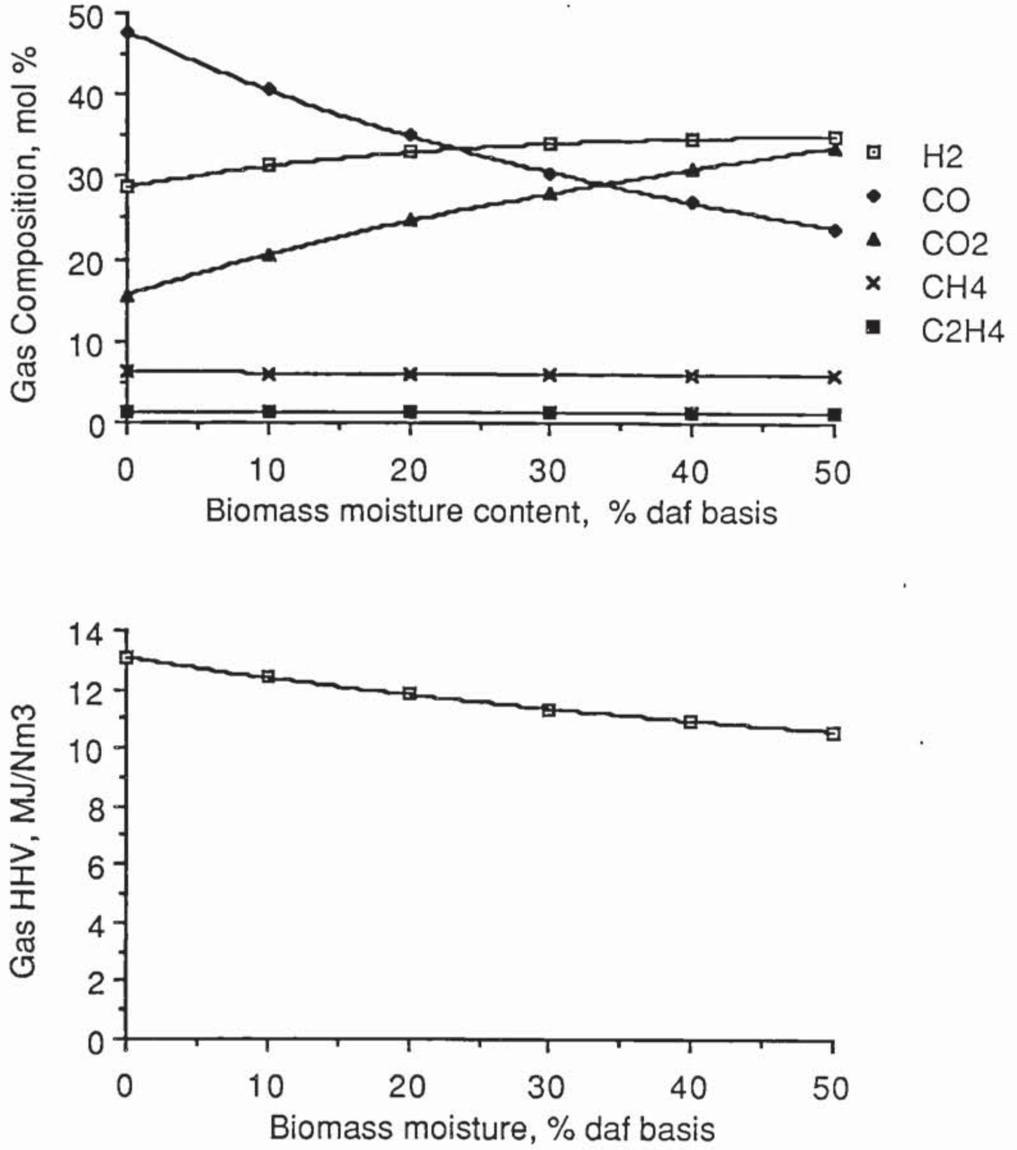


Figure 6.8 Predicted Gasifier Performance as a Function of the Biomass Moisture Content

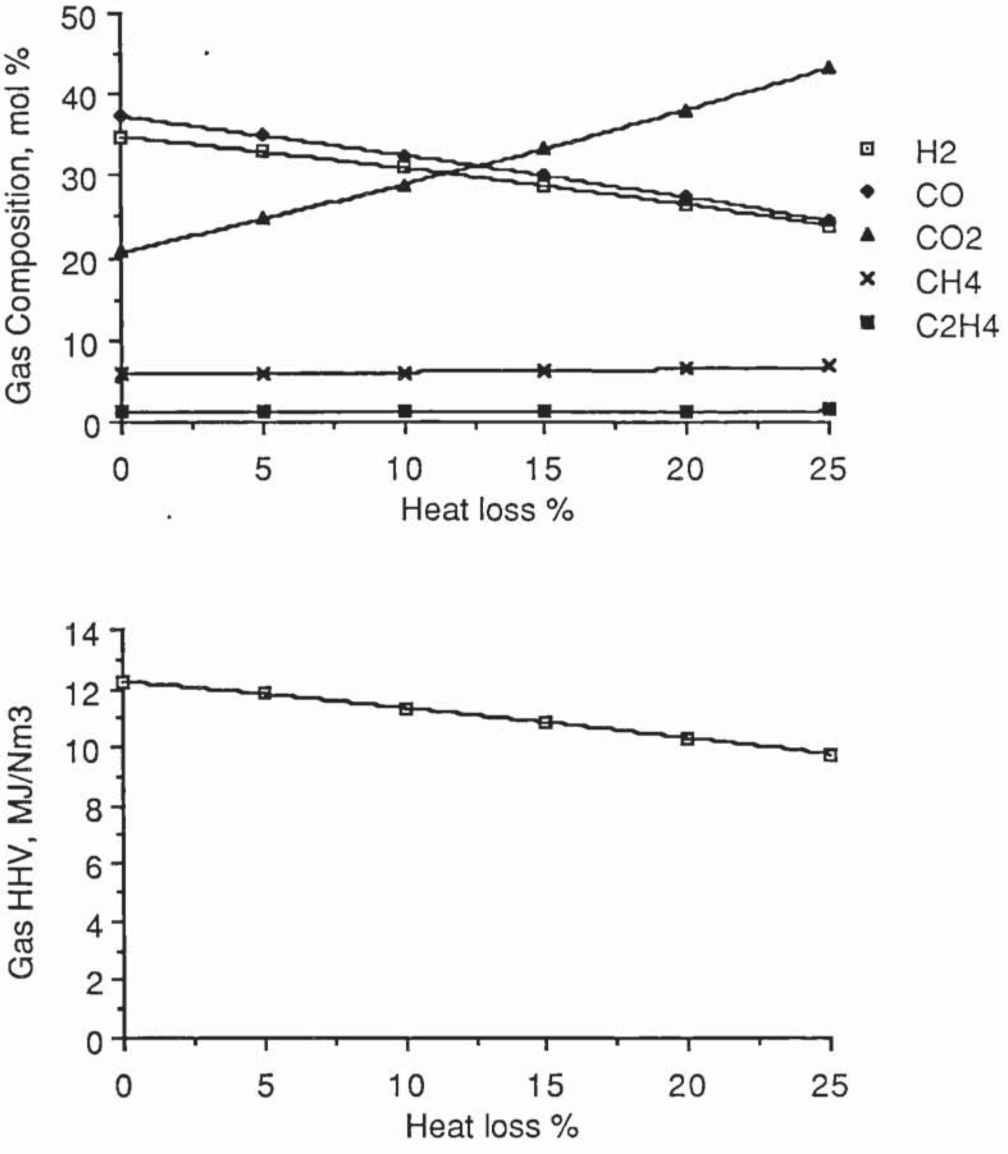


Figure 6.9 Predicted Gasifier Performance as a Function of the Gasifier Heat Loss

6.6 CONCLUSIONS

A computer program has been constructed and tested which allows stagewise modelling of fluid bed gasification with a range of different pyrolysis and tar cracking models.

Both models constructed using this program may be used to estimate the gas compositions from fluid bed gasifiers: Model 2 in the case where there is bypassing of pyrolysis gases, and Model 3 where the feedstock is rapidly mixed into the bed and reaction between pyrolysis products and the fluidising gases can occur. Model 3 gives more accurate predictions of real gasifier performance than Model 2.

The effect of the inclusion of tar in the pyrolysis products has been tested, and it was found that if tar is produced in the pyrolysis, some of this tar must undergo further reaction. Cracking of tar produced in pyrolysis can increase the methane content of the product gas, as demonstrated by Model 5.

The ability of these stagewise models to predict the product distribution of fluid bed gasifiers demonstrates the strong influence of pyrolysis on the product gas composition. This is supported by the results of Maniatis, [54] who found that the gas composition from a fluid bed gasifier was very sensitive to a shutdown of the biomass feed, the hydrocarbons in the product gas falling to zero in about 1 minute, which is about the same period as the time for pyrolysis of the wood chips used. [62,66]

Finally, sensitivity analyses carried out using the model show similar trends to those predicted by the adiabatic carbon boundary model.

CHAPTER 7
GASIFIER ECONOMICS

7.1 INTRODUCTION

For any developing technology to succeed and become a process which is used commercially, the process must be economically viable. However, the economics of a new process cannot be studied in a purely general way but must be based on a specific case where the technology can be employed. In order to study the economic viability of biomass gasification, it was decided to choose a particular application of gasification, and perform economic calculations on this case. The simplest application of biomass gasification technology is to use air gasification for the production of producer gas, which may be used in a number of applications, especially for process heat where the combustion of a solid fuel cannot easily be used. As simple technologies are likely to be adopted before more complicated ones, and as there may be a ready market for the product as a replacement for natural gas in some industrial applications, this case was thought to be the most likely initial use of biomass gasification in the UK and was thus chosen for the economic study.

In order to study the economics of a new process, both the capital cost and the operating costs of the plant must be estimated. The traditional methods of economic evaluation then require a product selling price to be estimated. From these data a cash flow table can be constructed, and measures of profitability such as payback time, return on investment (ROI), net present value (NPV) or discounted cash flow rate of return (DCFRR) are calculated. These methods of profitability assessment are described in many texts, for example by Allen [97] or Peters and Timmerhaus. [98] An alternative approach, which is especially applicable when it is difficult to estimate the product value, is to estimate a production cost which includes an element to cover capital costs, which may then be directly compared with the range of product values to test the viability of a process. In this work, an economic model of biomass gasification for the production of fuel gas in the UK has been developed, and implemented on a computer. This has enabled a series of cases to be evaluated and the influential variables in gasifier economics to be identified.

The calculations were performed at the start of 1986, so the costs used are all for the second half of 1985. Changes in the values of these costs between 1985 and the time of writing may have changed the results slightly in numerical terms, but the conclusions drawn from the results will remain the same.

This work has been published in a report to the UK Department of Energy, [4] and the economic model of gasification has been used in performing economic evaluations, the results of which have been published elsewhere. [99,100]

7.2 CAPITAL COSTS

7.2.1 Definition

A system for the production of a fuel gas from biomass involves a number of processing operations, some of which are always necessary, and others of which are only sometimes required. The processing operations which may be required are listed below: those which may not be essential are shown in parentheses.

Storage
Handling
(Size reduction)
(Screening)
(Drying)
Gasification
Gas cleaning
(Wastewater treatment)
(Gas compression)
(Pipeline to user)

The total plant cost of a gasification system is thus the equipment costs for all required steps listed above, together with costs of design, installation and commissioning.

For the purposes of this survey, capital costs of gasifiers have been collected and correlated, covering the process steps from the gasifier feeding system (feed on the ground) to clean cold product gas. This represents the case where a gasifier is built at an existing site which already has reception and storage arrangements. This would be typical of a gasifier constructed at a municipal refuse reception plant, or a factory wishing to gasify waste generated on site.

The capital cost considered is a total plant cost including equipment, installation, design, project management and commissioning, but excluding land costs. Some data have been collected on a delivered equipment cost basis and converted by suitable factors.

7.2.2 Capital Cost Data

Gasifier capital cost data are available from a variety of sources. [43,101,102] Data are usually expressed as single figures representing the equipment cost of a gasifier systems from the gasifier feeding equipment to clean gas i.e. it includes equipment for the feeding, gasification and gas clean-up steps of the process. In some cases total plant costs are available in addition to or instead of the delivered equipment cost. There are very little data on the breakdown of the capital cost into either costs of equipment for the individual process steps, or elements of capital cost.

The raw data on gasifier equipment cost are tabulated as Table D1 of Appendix D. In order to compare the capital costs on a consistent basis, the data have all been converted to $£_{1985}$ sterling, using the cost indices and exchange rate information from Process Economics International. The raw capital cost data are plotted against gasifier throughput in Figure 7.1. All throughputs are expressed as the flowrate of dry, ash free (daf) feedstock entering the gasifier.

Data on total installed gasifier costs are shown in Table D2 of Appendix D and presented graphically as Figure 7.2. In many cases, data on total plant costs are not published, and these data have been derived in part by multiplying the equipment cost by a suitable factor, as described in Section 7.2.3.

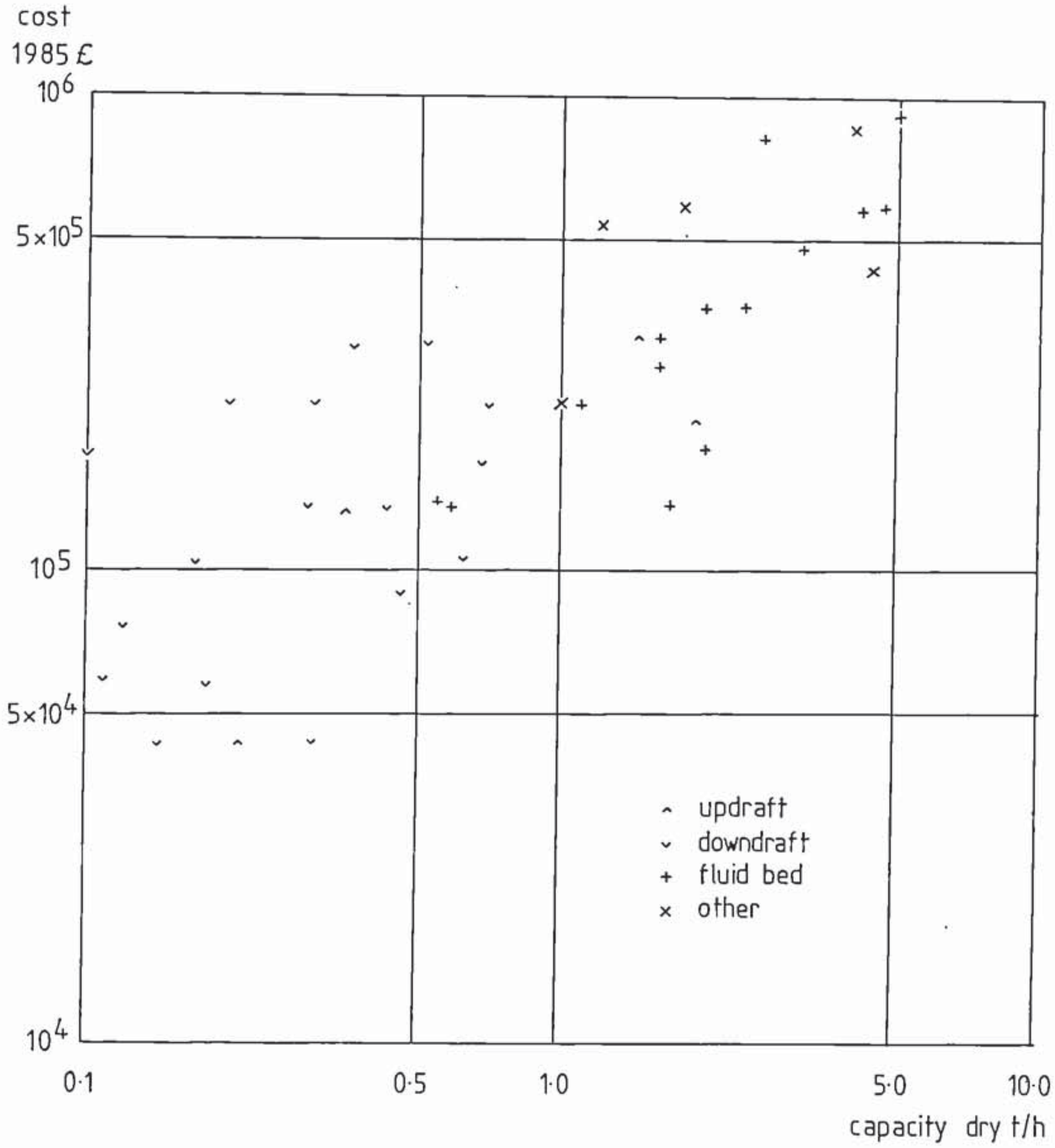


Figure 7.1 Gasifier delivered equipment cost v. gasifier capacity

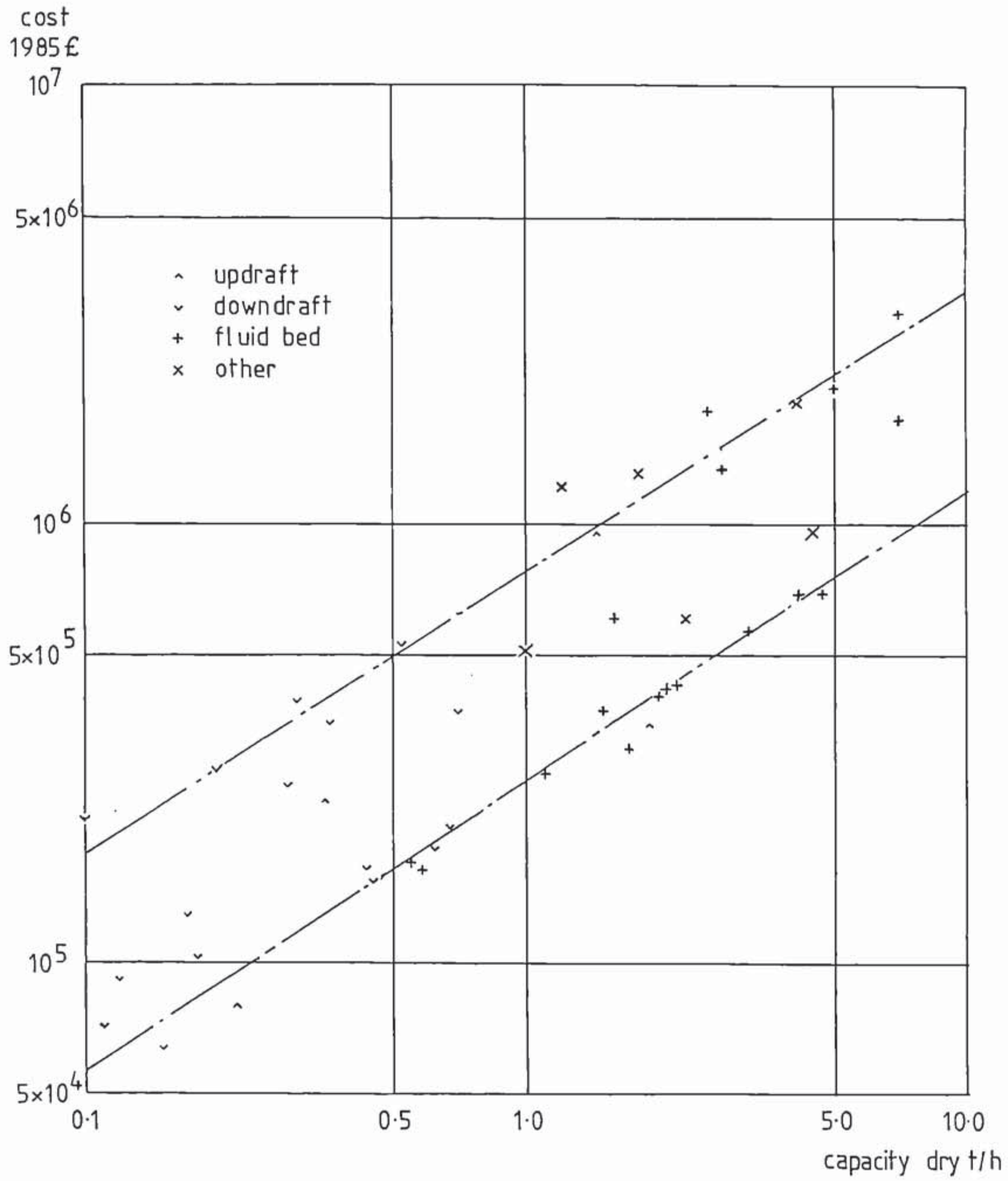


Figure 7.2 Total installed gasifier capital cost v. gasifier capacity

The important figure for calculations of process economics is the total installed capital cost of the plant. Looking at Figure 7.2, it can be seen that the data points generally lie between two parallel lines. The equations of the two lines are as follows:

Lower: $£_{1985}$ Capital Cost = 260 000 (capacity, daf t/h)$^{0.65}$

Higher: $£_{1985}$ Capital Cost = 775 000 (capacity, daf t/h)$^{0.65}$

In each case, the capital cost is expressed in $£_{1985}$ sterling, and the capacity is in daf t/h. Both lines show a scale factor of 0.65 - close to the widely accepted scale factor for processing plant of 2/3. At a given throughput, the upper correlation predicts a capital cost about three times that of the lower line. There are several reasons for this range of capital costs. The level of sophistication of the technology and the quality of engineering differs between gasifiers. Some gasifiers are built and marketed by small companies with low overheads and low levels of research and development backup, whereas others are built by large companies whose high overheads and large research and development budgets are reflected in their higher gasifier selling prices.

The lower line thus represents the lower limit of gasifier capital costs, typically achieved by relatively simple gasifier systems and systems supplied by small companies. This correlation should be used where gas quality and gasifier reliability are not paramount, for example where that product gas is being used as a fuel gas supplementing natural gas.

The upper line represents the cost of high technology systems, for example twin fluid bed systems, systems with extensive use of mechanised feeding and automatic control, and systems supplied by major companies. It would be appropriate to use the upper correlation when a gas of higher quality than normal was required, for example where the gasifier was to produce synthesis gas.

This range of capital costs agrees with data proposed by Earthscan [103] who also proposed two levels of capital cost. Their lower figure agrees well with the lower figure here, but their upper figure is four times higher than their lower figure rather than the factor of three found here. They do not, however, report any economy of scale. Bridgwater [99] proposed a third, lower level of gasifier capital costs for gasifiers built in developing countries, using very simple technology, and designed for manual feeding and control. This would give capital costs about half those of the lower correlation shown above. There is no published evidence to support this suggestion, but it would be expected that simple designs of gasifiers could be built at very low cost.

7.2.3 Capital Cost Relationships

In most cases, capital cost data are not available broken down into capital costs for each process step. As a rule of thumb, the cost of the gasifier reactor and associated feeder is about 60% of the equipment cost. [4] However, in view of the limited data available, further conclusions cannot be drawn, nor can the effect of different feed materials be assessed. An analysis of the capital costs of two systems are presented as Table 7.1.

There are more data available on the relationship between equipment cost and total installed cost of gasifier systems. The ratio of total plant cost to equipment cost varies depending on the type of gasifier construction, but can be estimated with a fair degree of confidence.

For packaged and skid-mounted units, the gasification plant is complete but for a concrete area to stand on, and piping up to utilities. A light shed or cover may also be required. Figures quoted by various manufacturers indicate that the total installed cost for such a system would be about 1.2 times the equipment cost. [104]

For large field erected units, over about 1 t/h capacity, much civil and fabrication work is usually required on site, and available data indicate that the total plant cost is about 2.25 times the equipment cost. [101] In the paper industry (with similar types of processing equipment), the accepted ratio of total plant cost to equipment cost is about 2.25, [4] confirming this conclusion.

For smaller units which are not skid-mounted, less on-site fabrication would be required than for the large units, and the ratio of total plant cost to equipment cost will be smaller than that for large units. Century Research quote a figure of 1.66, although there are indications that for other makes of gasifier the ratio could be higher than this. [43] A figure of 1.75 has been used for calculating total plant costs from equipment costs for this type of gasifier where no installed cost data are available.

Table 7.1 Analysis of the Capital Costs of Two Gasifier Systems

Item	Battelle Columbus 20.8 daf tonnes/h Cost US$	 % *	JBEC / Wellman ODG 5.0 daf tonnes/h Cost £	 % *
Store/handle	713 800	27.5	34 500	3.9
Gasifier & feed	1 321 300	51.0	553 300	63.3
Gas clean-up	0	0.0	111 500	12.8
Heat recovery	377 000	14.5	174 500	20.0
Water treatment	182 400	7.0	0	0.0
Total equipment	2 594 500	100.0	873 800	100.0
Total plant	4 874 400	188.0	1 962 000	225.0

* Percentage of total equipment cost

7.2.4 Comparison of the Capital Costs of Gasification and Related Technologies

Figure 7.3 shows the capital costs of plants for both coal gasification and waste combustion, compared with the capital cost correlations for biomass gasification. The data for this graph are shown as Table 7.2. The capital costs are adjusted to the same basis as the gasifier capital costs in Figure 7.2, in £$_{1985}$ sterling.

Comparison with coal gasification. Two figures for the capital costs of coal gasification plants are included in Figure 7.3. These are

Table 7.2 Total Plant Capital Costs of Coal Gasification and Waste Combustion Technologies

Technology	Throughput daf tonnes/h	Capital cost £$_{1985}$	Ref
Coal gasification	2.7	2 300 000	4
Coal gasification	5.0	4 600 000	104
Refuse incineration	2.5	1 375 000	4
Refuse incineration	4.0	2 200 000	4
Refuse incineration, fluid bed	4.0	1 800 000	4
Refuse incineration, Straw combustion	10.0	4 250 000	4
Waste combustion	0.4	140 000	105
Waste combustion	0.2	118 000	106
Waste combustion	0.3	227 000	107
Waste combustion	2.2	2 061 000	108
Wood combustion	0.1	55 000	109
Waste combustion	0.4	282 000	110
Waste combustion	0.2	85 000	111
Waste combustion	0.2	139 000	112
Straw combustion	2.0	198 000	113

considerably above the line representing the capital cost of high technology biomass gasifiers. However, when the figures are adjusted for the increased calorific value of coal compared with biomass, they move much nearer the line, with the lower size of coal gasifier being almost on the line. Conversely there is evidence that the lower reactivity of coal will require larger reactors which would negate this effect. It can be concluded that on a thermal capacity basis, the costs of coal gasifiers are comparable to high technology biomass gasifiers, both of which tend to be supplied by major companies with high overheads.

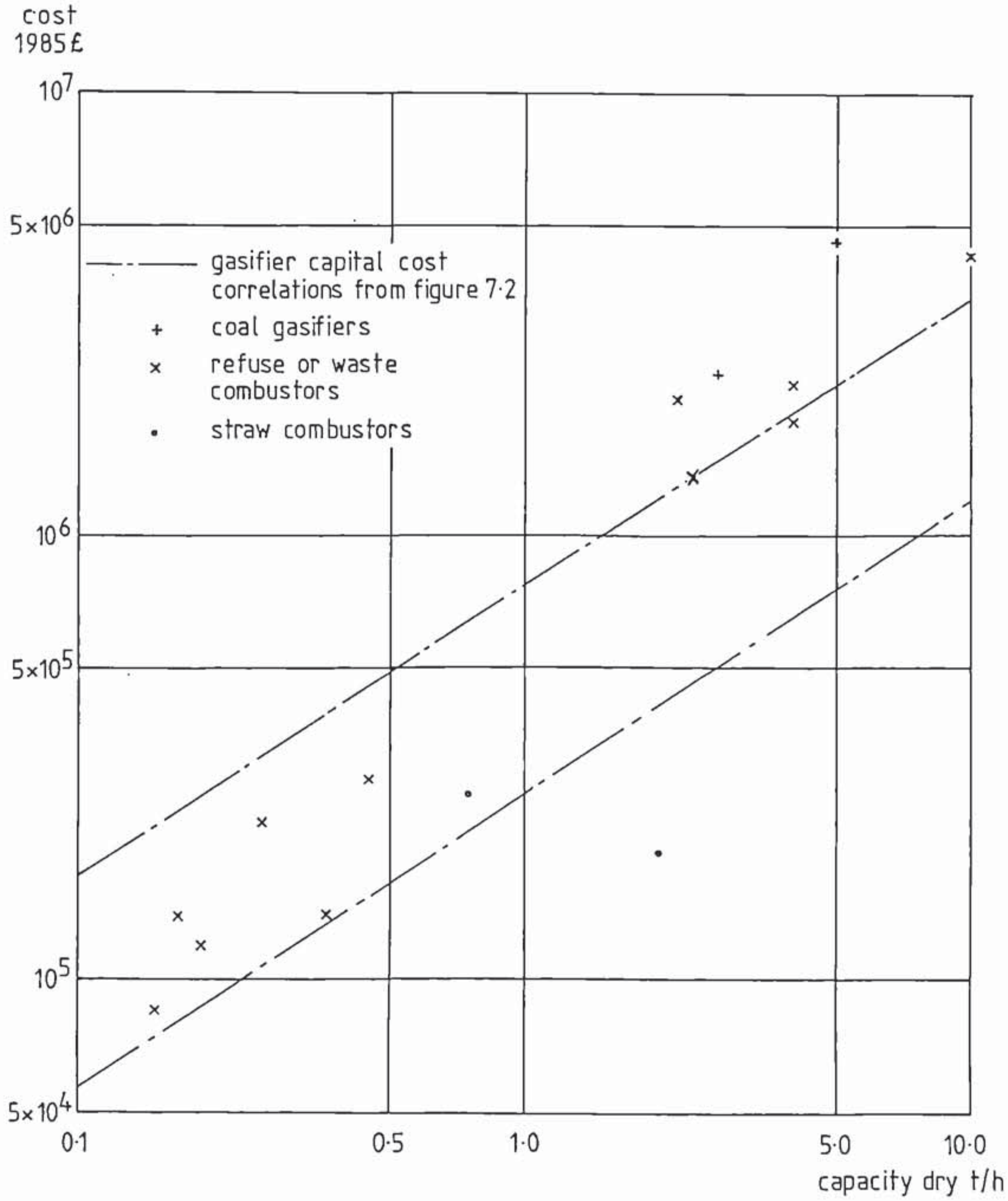


Figure 7·3 Comparison between the total installed capital costs of gasifiers and related technologies

Comparison with waste combustion. Some data are available for the combustion or incineration of wastes, refuse and straw. The data shown in Figure 7.3 are within the range of gasifier capital costs, but there is an apparently lower economy of scale, making combustion appear to be less attractive at higher capacities.

The only exception to this conclusion is the Needham Chalks straw combustor, which has a very low capital cost.[113] It is understood that another quotation for the same combustor was about twice the figure of the combustor purchased, which would bring the cost into the range of low technology gasifiers.

It can be concluded that the capital costs of combustion and gasification are similar for wastes and refuse.

7.3 OPERATING COST

Gasifier operating costs for a system for converting biomass as prepared feed on the ground to cold clean gas are considered in this section. The operating costs considered here are feedstock, utilities, maintenance, overheads and labour.

7.3.1 Feedstock Cost

Bridgwater[4] has discussed the availability and price of renewable feedstocks for gasification in the UK. The feedstocks which are likely to be considered for use are wood and wood waste, straw, and refuse.

Wood and wood waste. These occur in several forms: forestry residues, wood processing residues, wastes from industries which use wood as a raw material, and fuel wood. Of these, only forest residues and sawmill residues are likely to be available in sufficient quantities.

The cost of wood wastes is estimated to be £13/tonne on the site where the waste is generated, or £17/tonne if delivery within a radius of 25 miles is considered. These figures are applicable to wood with a moisture content of 50% wet basis, giving costs of £26/tonne on site or £34/tonne delivered on a dry wood basis. Fuel wood, that is wood grown as an energy crop, is likely to be considerably more expensive than this.

Straw. The largest quantity of agricultural waste in the UK is straw - there is about 7 x 10^6 tonnes/year of unused waste straw generated in the UK. [4] This was estimated to cost £17/tonne on farm, or £22/tonne if delivered to a 25 mile radius. Straw has a moisture content of 17% wet basis, giving dry basis costs of £20/tonne on farm or £26/tonne delivered. However, it is unlikely that sufficient waste straw would be produced by a single farm to supply a biomass gasifier of an economic throughput.

Refuse. As a feedstock for gasification, refuse has the advantage that it must be collected and brought to a single point for disposal. As well as this, there is a cost associated with disposing of refuse, which would be saved if the refuse was gasified. However, the cost of separating and disposing of the non-combustible fraction of refuse must be included when assessing the cost of refuse as a feedstock.

Bridgwater [4] analysed the disposal costs of refuse, and found that refuse disposal costs between £2/tonne and £15/tonne in the UK, with a mean of £6/tonne for all counties, or £10/tonne for the metropolitan counties. He also estimated the costs of preparing the refuse for conversion, which included shredding, and screening and air classification to remove metals and other non-combustible materials. Thus, a refuse which cost £10/tonne to dispose of, i.e. which had a value of £-10/tonne, would cost £5/tonne as a prepared feed, or £15/tonne if drying were included as well.

The feedstock costs used in the economic analysis are summarised in Table 7.5 (p 157).

7.3.2 Utilities

Utilities include:

electricity, for driving pumps, blowers and feeding systems
wash water for product gas scrubbers
boiler feedwater for steam raising.

Steam and oxygen have been excluded. Steam, if required, would be raised using a waste heat boiler to recover heat from the hot product gases. As the current evaluation is of air gasification, oxygen is not required. However, if an evaluation of oxygen gasification were to be performed, the cost of oxygen would probably be significant. In some systems, a fuel such as propane gas is required to preheat the gasifier during start-up. In the current evaluation, only steady state operation has been examined, so this cost has not been included.

Table 7.3 shows data on electricity costs of gasifiers, expressed as £/GJ product gas. The mean of this data is £0.114/GJ. However, this does not include the other utilities costs, where very little information is available. A utilities cost of £0.24 /GJ was used for the base case in the production cost calculations. This has been estimated as the mean of the electricity cost multiplied by 2 to cover the cost of water, etc, and is approximately the figure estimated for the ODG process.

Table 7.3 Utilities Costs of Gasifier Systems

Gasifier	Electricity Cost* £/GJ	Total Utilities Cost £/GJ	Ref
ODG	0.15	0.27	101
Applied Engineering Co	0.105	-	43
Battelle PNL	0.042	-	43
EZ Manufacturing	0.155	-	43
Biomass Corporation	0.146 to 0.195	-	43
Duvant Model GA 20	0.205	-	43
Energy Equipment Engineering	0.043	-	43
Fritz Werner	0.109	-	43
Pyrenco	0.093	-	43
Vyncke Warmetechnick	0.130	-	43
National Synfuels	0.062 to 0.099	-	43
Omni	0.115 to 0.127	-	43
Sur-lite	0.043	-	43
MEAN	0.114		

* Electricity cost is calculated from published figures of electricity consumption, kWh/GJ, using a cost of 3.1 p/kWh [114]

7.3.3 Maintenance

Yearly maintenance cost is usually estimated as a proportion of the gasifier capital cost. Raw data are often expressed in terms of the equipment cost, but for the purposes of the computer model of gasifier economics, the maintenance cost is expressed as a fraction of the total plant cost. Table 7.4 shows the raw data on gasifier maintenance costs, and the data adjusted to a total plant cost basis. The mean maintenance cost is 2.5% on a total plant cost basis, and this figure was used in the product cost calculations.

Table 7.4 Maintenance Costs of Gasifier Systems

Gasifier	Annual Maintenance Cost Fraction of Equipment Cost	 Fraction of Total Plant Cost	Ref
ODG	-	0.002	101
Applied Engineering Co	0.031	0.018	43
EZ Manufacturing	0.03	0.018	43
Biomass Corporation	0.05	0.042	43
Duvant model GA 10	0.04	0.022	43
Duvant model GA 20	0.04	0.022	43
Fritz Werner	0.05	0.042	43
Imbert	0.03 to 0.06	0.024 to 0.047	43
Pillard	0.05	0.029	43
National Synfuels	0.05	0.022	43
Alberta Industrial Developments	0.04	0.018	43
Energy Products of Idaho	0.035	0.016	43
Omni	0.054	0.024	43
Sur-lite	0.04	0.018	43
MEAN		0.025	

7.3.4 Overheads

Annual overheads cost is usually expressed as a proportion of the capital cost, similar to maintenance cost. The economic model of gasification uses a fraction of the total plant cost to calculate annual overheads which include rates (local tax), insurance and all head office expenses. Payroll overheads have been included in the labour costs.

No relevant published information has been found which can be used to estimate overheads for a UK location. Data from overseas will not necessarily be applicable. The only information available suggests the overheads for a commercial Oxygen Donor Gasifier system would be 8% of the total plant cost. [101] The local rates and insurance for a gasifier would typically be 4% of the capital cost. [101] This figure was multiplied by 2 to cover head office expenses, giving a figure of 8%, which matches the Oxygen Donor Gasifier estimate. This figure of 8% of total plant cost was therefore used in the economic analysis.

7.3.5 Labour Cost

The cost per shift used for the base case was £15 000/year. This includes the costs of one operator per shift, plus the costs of supervision and payroll overheads. It has been assumed that the costs of supervision are shared with other plants on a site. Four shifts are required for continuous operation, the base case - three shifts per day, plus an extra shift to cover rest days. If operation is restricted to weekdays only, then three shifts will be sufficient.

7.3.6 Product Gas Value

The calorific value of producer gas for biomass gasification is much lower than that of natural gas. The quality of the gas is likely to be more uncertain than natural gas, and the continuity of supply is also likely to be uncertain. The selling price of the gas, expressed as £/GJ, will therefore have to be less than that of natural gas. It has been estimated that in order to make a producer gas attractive to users, its cost will have to be of the order of 75% to 80% of the price of natural gas to industry. It was decided to use a product value of £2.40/GJ in the calculations, which is 75% of the price of natural gas to industry, £3.20/GJ. [4]

7.4 PRODUCTION COST AND SENSITIVITY ANALYSIS

7.4.1 Methods of Economic Evaluation

There are two approaches to the evaluation of a proposed project - calculation of a product cost, or the conventional profitability measures of payback time, ROI, NPV and DCFRR.

Product cost is calculated by totalling the feedstock cost and operating costs (as described in Section 7.3) and adding to them a capital amortisation charge, which covers recovery of the capital and interest payments on the capital employed. The product cost may then be compared with the product selling price to asses profitability.

In contrast, conventional profitability measures include all cash flows, both income and expenditure, but do not include charges for interest or capital recovery. This is because they are designed as a method of comparing investment in a project with the investment of a similar sum in a bank. DCFRR and ROI are a percentage rate of return, which can be compared with the bank interest rate. NPV is expressed as the value of a project in current pounds, at a fixed rate of return. These measures are usually expressed in real terms, i.e. allowing for the change in the value of money with inflation. Payback time is the time taken for the capital to be recovered, ignoring interest payments, and is usually expressed in nominal terms.

7.4.2 Fuel Gas Production Costs and Economic Evaluation

A computer program has been used which calculates product cost, DCFRR, and payback time. In order to study the economics, two sets of calculations have been performed, and the results are presented in Tables 7.7 and 7.8. The lower capital cost correlation (see Section 7.2.2) was used as this represents a target price for a commercial system. In all cases the production of a cold, clean gas has been considered. All costs and interest rates are those which applied in the second half of 1985.

The first set of calculations considered gasification of wood, straw or refuse at a range of throughputs, in order to identify applications where gasification is economically attractive. The range of feedstocks is shown in Table 7.5. In all of these cases, the values of the other parameters were as discussed above.

The second set of calculations was a sensitivity analysis, using the base case shown in Table 7.6. The parameters for the base case were as in the first set of calculations, but a specific feedstock was chosen - prepared but undried refuse - and a throughput of 2.5 tonnes/h was chosen as this was thought to be the size of gasifier which would be most widely adopted. [4]

Table 7.5 Feedstock Characteristics

Feedstock	Moisture content	Resulting gasifier efficiency	Cost raw basis £/tonne	Prepared cost dry basis £/tonne
Straw, on farm	17%	78%	17	20
Straw, delivered	17%	78%	22	26
Refuse, undried	35%	71%	0	25
Refuse, undried	35%	71%	-5	15
Refuse, undried	35%	71%	-10	5
Refuse, undried	35%	71%	-15	-5
Refuse, dried	17%	78%	-10	15
Wood, on site	50%	62%	13	26
Wood, delivered	50%	62%	17	34

Table 7.6 Data and Results of Economic Calculations for the Base Case

DATA (All tonnes are daf basis)

Gasifier throughput	2.50	t/h
Feedstock heating value	20.00	GJ/t
Gasifier efficiency	71.00	%
Number of shifts	4	
Project life	10	y
Operating hours per year	6000	
Capital cost (including working capital)	£472 000.00	
Feedstock cost for prepared refuse (daf basis, after processing)	£5.00	/t
Product selling price	£2.400	/GJ
Utilities cost	£0.240	/GJ
Yearly maintenance, fraction of capital cost	0.025	
Yearly overheads, fraction of capital cost	0.080	
Total cost of labour per shift	£15 000.00	/y
Target real discount rate	10.00	%
Nominal cost of capital	12.00	%
Inflation rate	5.00	%
Feedstock cost real escalation rate	0.00	%
Product price real escalation rate	0.00	%
Utilities cost real escalation rate	0.00	%
Maintenance cost real escalation rate	0.00	%
Overheads real escalation rate	0.00	%
Labour cost real escalation rate	0.00	%

RESULTS

NPV at 10.00% real discount rate is £1 220 951
DCF rate of return is 57.8 % real, 65.7 % nominal
Payback time is 1.60 years from start-up, nominal basis.
Mean real product cost over project life is £1.41/GJ

7.4.3 Sensitivity Analysis

A sensitivity analysis to test the effect of selected parameters affecting gasifier economics has been carried out, and a summary of the results is shown in Table 7.8. Feedstock cost will have a strong influence on production cost, the effect on the product cost being linear. It will also strongly influence the conventional profitability measures. Product selling price has no effect on production cost, but has a large influence on profitability. As the influence of these two factors on profitability is very important it was decided to study their effect separately from the sensitivity analysis, as discussed in Section 7.4.4.

Gasifier throughput The effect of gasifier throughput and level of technology have been investigated by performing the calculations at throughputs of 1, 2.5, 5 and 10 daf t/h, using both high and low capital cost correlations from Section 7.2. The results are shown graphically in Figure 7.4.

Gasifier efficiency Gasifier efficiencies of 45% to 80% have been used, covering the range of efficiencies for both pyrolysis and gasification at all feedstock moisture contents. The results indicate that efficiency has an effect on economics, but this is not as strong as the effects of feedstock cost and scale (Figure 7.5).

Project life Project lives of 3, 5, 10 and 15 years were considered. Project life has no effect on payback times, as long as the payback time is less than the project lives being considered. This is because payback time only measures short-term profitability. However, project life affects

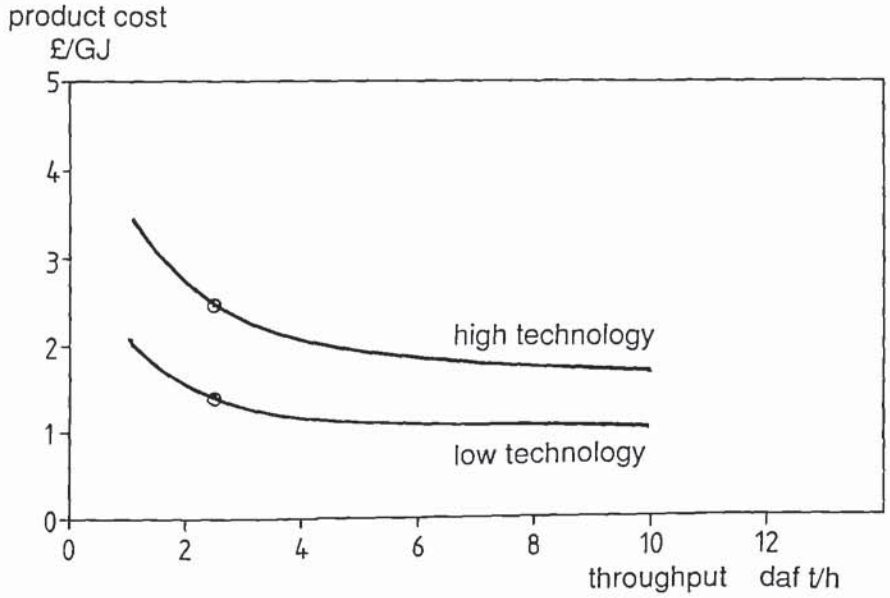


Figure 7.4 Product cost against gasifier throughput

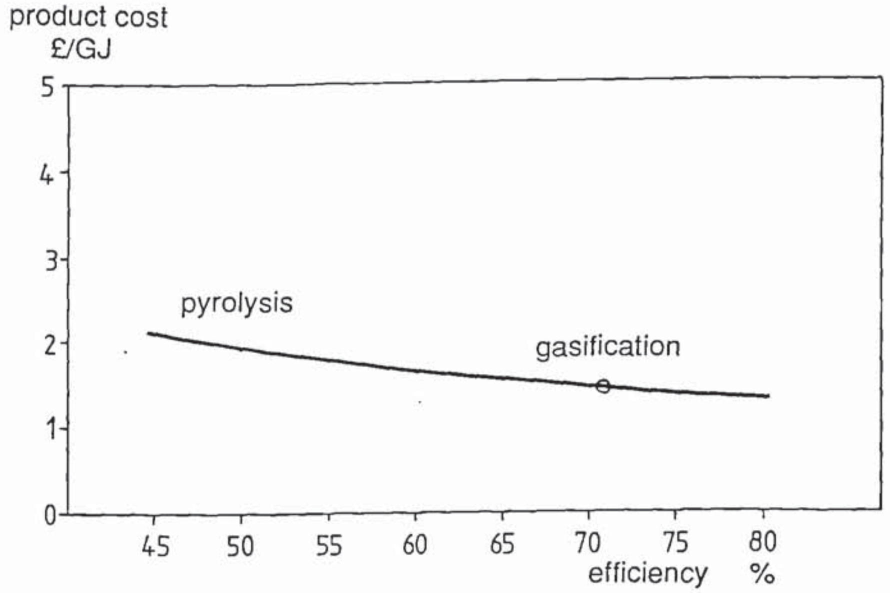


Figure 7.5 Product cost against gasifier efficiency

both product cost, as the capital cost is amortised over a different period, and also DCFRR, as the income of the later years of a project is lost by shortening a project lifetime.Project life is not one of the important factors in project profitability.

Operating hours The base case assumes continuous operation at a 70% load factor, ie 6000 hours/year. This is in line with combustion plants. The effect of higher and lower load factors have been investigated by considering operating hours of 4000 and 8000 h/y, while still assuming 4 shift operation. These variations have a moderate effect on productivity, although gasification is still attractive at 4000 h/y.

The effect of three shift (gasifier shut down at weekends), two shift and single shift operation are considered. Three shift and two shift operation are still attractive, but single shift operation must be considered uneconomic. However, the economic model used did not include the extra costs associated with start-up and shut-down, so these results may not be valid if these costs were found to be appreciable.

Feedstock escalation The base case included no differential inflation, all real escalation rates being zero. However, the effect of feedstock cost escalation rate was investigated by performing calculations at -5%, -2.5%, 0%, 2.5% and 5% real escalation rates. The negative escalation rates cover the case where disposal credit for waste is increasing, but pretreatment costs are constant - the cost of the processed waste being above zero, but being reduced by the increase in disposal credit. The positive escalation rates cover possible increases in the cost of feedstocks if a market in them is established, forcing up prices.

Feedstock cost escalation does have an effect on gasifier profitability, the profitability being reduced with increasing real escalation, but it is not a dominating factor in the economics.

Product price escalation The effect of product price escalation was tested by using real escalation rates of -5%, -2.5%, 0%, 2.5% and 5 %. There is obviously no effect on product cost, but increasing energy escalation increases profitability , the order of magnitude of the effect being similar to that for feedstock escalation .

Table 7.7 Results of Economic Calculations

Conditions	Mean Real Product Cost £/GJ	Real DCFrr %	Nominal Payback Years
Straw, 17 % moisture, £17/t			
1 t/h gasifier	2.83	<-90	>10
2.5 t/h gasifier	2.27	15.5	4.31
5 t/h gasifier	2.03	35.1	2.48
10 t/h gasifier	1.89	54.5	1.69
Straw, 17 % moisture, £22/t			
1 t/h gasifier	3.22	<-90	>10
2.5 t/h gasifier	2.65	-26.7	>10
5 t/h gasifier	2.42	4.4	6.57
10 t/h gasifier	2.27	20.4	3.69
Refuse, 35 % moisture, £0/t disposal credit			
1 t/h gasifier	3.44	<-90	>10
2.5 t/h gasifier	2.82	<-90	>10
5 t/h gasifier	2.56	-12.7	>10
10 t/h gasifier	2.40	6.1	6.12
Refuse, 35 % moisture, £5/t disposal credit			
1 t/h gasifier	2.74	-19.2	>10
2.5 t/h gasifier	2.11	23.3	3.37
5 t/h gasifier	1.86	47.7	2.06
10 t/h gasifier	1.70	64.9	1.43
Refuse, 35 % moisture, £10/t disposal credit			
1 t/h gasifier	2.03	22.2	3.48
2.5 t/h gasifier	1.41	57.7	1.60
5 t/h gasifier	1.15	85.3	1.11
10 t/h gasifier	0.99	117.0	0.81

Table 7.7 (continued)

Conditions	Mean Real Product Cost £/GJ	Real DCFrr %	Nominal Payback Years
Refuse, 35 % moisture, £15/t disposal credit			
1 t/h gasifier	1.33	47.8	1.91
2.5 t/h gasifier	0.71	90.0	1.05
5 t/h gasifier	0.45	126.0	0.76
10 t/h gasifier	0.29	168.8	0.56
Refuse, 35 % moisture dried to 17 % moisture, £10/t disposal credit			
1 t/h gasifier	2.51	-0.8	8.27
2.5 t/h gasifier	1.95	34.3	2.53
5 t/h gasifier	1.71	56.5	1.63
10 t/h gasifier	1.56	80.8	1.17
Wood, 50 % moisture, £13/t			
1 t/h gasifier	3.99	<-90	>10
2.5 t/h gasifier	3.27	<-90	>10
5 t/h gasifier	2.98	<-90	>10
10 t/h gasifier	2.79	<-90	>10
Wood, 50 % moisture, £17/t			
1 t/h gasifier	4.63	<-90	>10
2.5 t/h gasifier	3.92	<-90	>10
5 t/h gasifier	3.62	<-90	>10
10 t/h gasifier	3.44	<-90	>10

Table 7.8 Results of the Sensitivity Analysis

Conditions	Mean Real Product Cost £/GJ	Real DCFrr %	Nominal Payback Years
Base Case	**1.41**	**57.7**	**1.60**
1 t/h low capital cost	2.03	22.2	3.48
2.5 t/h low capital cost (base)	**1.41**	**57.7**	**1.60**
5 t/h low capital cost	1.15	85.3	1.11
10 t/h low capital cost	0.99	117.0	0.81
1 t/h high capital cost	3.50	-24.1	>10
2.5 t/h high capital cost	2.47	4.41	6.58
5 t/h high capital cost	1.98	17.4	4.05
10 t/h high capital cost	1.64	29.9	2.82
45% efficiency (pyrolysis)	2.09	18.5	3.90
50% efficiency (pyrolysis)	1.90	26.8	3.06
55% efficiency (pyrolysis)	1.75	34.5	2.51
60% efficiency	1.62	42.0	2.14
65% efficiency	1.52	49.2	1.86
70% efficiency	1.43	56.4	1.64
75% efficiency	1.35	63.4	1.47
80% efficiency	1.28	70.4	1.33
3 year project life	1.94	34.3	1.60
5 year project life	1.64	50.9	1.60
15 year project life	1.33	58.3	1.60
4000 hours/year operation, 4 shift	1.81	28.7	2.91
8000 hours/year operation, 4 shift	1.21	85.4	1.11
3 shift operation, 4300 hours/year	1.63	36.8	2.39
2 shift operation, 4000 hours/year	1.60	35.8	2.44
1 shift operation, 2000 hours/year	2.41	5.9	6.18

Table 7.8 (continued)

Conditions	Mean Real Product Cost £/GJ	Real DCFrr %	Nominal Payback Years
-5% real escalation of feed cost	1.32	59.7	1.57
-2.5% real escalation of feed cost	1.36	58.8	1.59
2.5% real escalation of feed cost	1.46	56.6	1.62
5% real escalation of feed cost	1.52	55.4	1.63
-5% real escalation of prod. price	1.41	41.8	1.84
-2.5% real escalation of prod. price	1.41	50.2	1.71
2.5% real escalation of prod. price	1.41	64.9	1.51
5% real escalation of prod. price	1.41	71.6	1.43

7.4.4 Product Price and Feedstock Cost

Product price and feedstock cost both have a strong influence on profitability, but they are also the factors which are most likely to be site specific. Therefore, it was decided to examine the relationship between them, given a specified level of profitability.

In many areas of industry, a project with a payback time of greater than three years would not be considered for implementation, even if other measures of profitability were favourable. It was thus decided to study the relationships between feedstock cost and the product selling price, if a 3 year payback time were required, at a range of throughputs.

Figure 7.6 shows the relationship between product selling price and feedstock cost, for a three year payback period. The capital costs have been estimated using the lower correlation of Section 7.2.2, at throughputs of 1, 2.5, 5 and 10 t/h. All other conditions are as for the

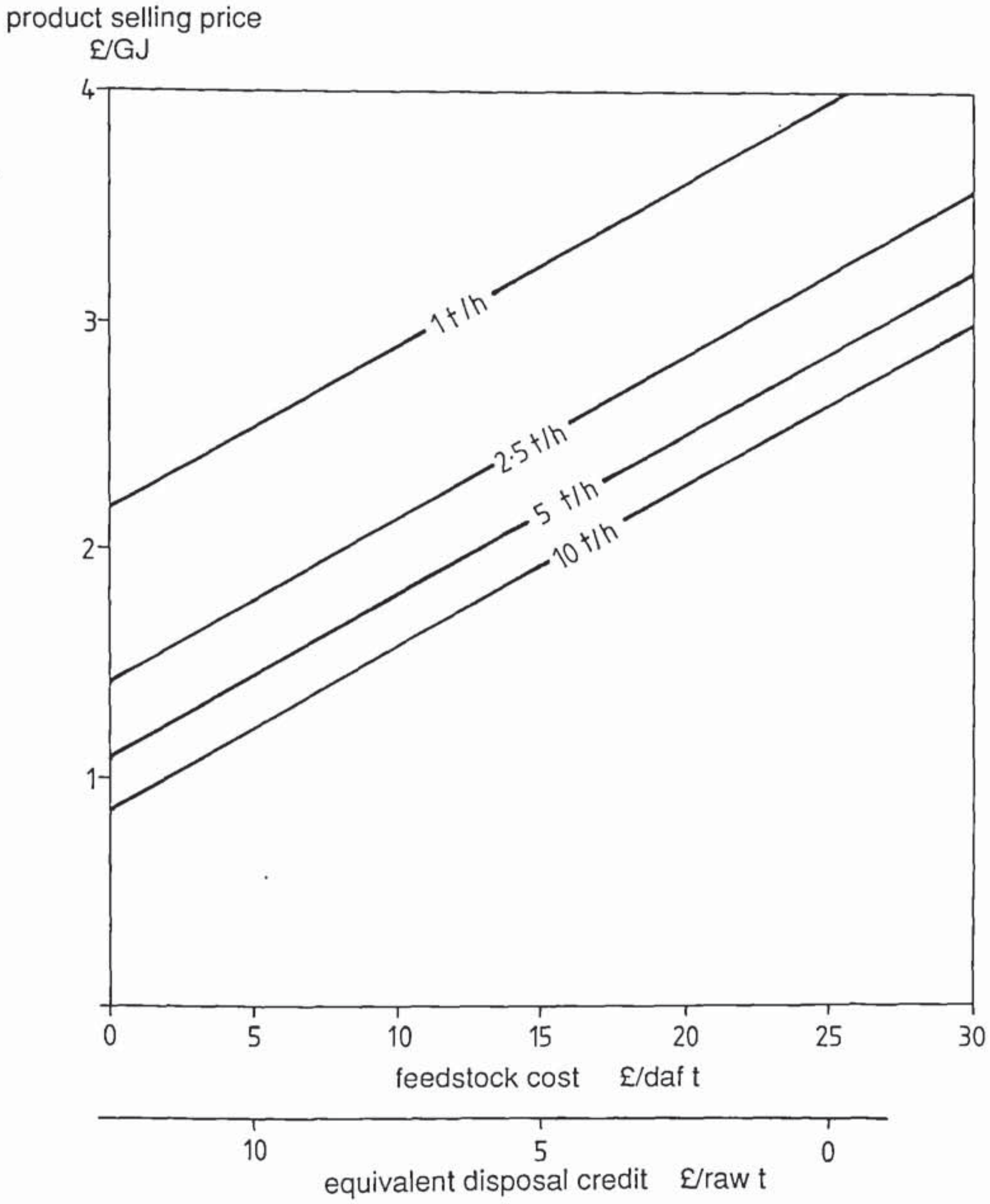


Figure 7.6 **Relationship between product selling price and feedstock cost for a 3 year payback time**

base case (Table 7.6). This diagram enables the required product selling price to be estimated if the feedstock cost is known, given that a three year payback time is required. Conversely, if a product price has been fixed, the cost of feedstock required to attain a three year payback can be estimated. Similar diagrams could be constructed for other methods of profitability if required.

7.4.5 Production Cost Analysis

A detailed analysis of production costs of fuel gas for a range of cases is shown in Table 7.9. This shows the significance of feed cost in most situations.

Table 7.9 Production Cost Analysis

Scope: 2.5 tonnes/h feed daf basis; unspecified conditions as base case.

	Total cost £/GJ	Cap cost £/GJ	Feed cost £/GJ	Lab cost £/GJ	Util cost £/GJ	O/h & maint £/GJ
STRAW (78% conversion efficiency at 17% water)						
£22/t delivered (£26/t daf)	**2.65**	0.28	1.67	0.26	0.24	0.21
£17/t on-farm (£20/t daf)	**2.27**	0.28	1.28	0.26	0.24	0.21
REFUSE -WET (71% conversion efficiency at 35% water)						
£10/t raw refuse disposal credit (£5/t daf, shredded, screened, classified)	**1.41**	0.30	0.35	0.28	0.24	0.23
£5/t raw refuse disposal credit (£15/t daf, shredded, screened, classified)	**2.11**	0.30	1.06	0.28	0.24	0.23
REFUSE-DRY (78% conversion efficiency at 17% water)						
£10/t raw refuse disposal credit (£15/t daf, shredded, screened, classified, dried)	**1.95**	0.28	0.96	0.26	0.24	0.21
WOOD (62% conversion efficiency at 50% water)						
£17/t delivered (£34/t daf)	**3.92**	0.35	2.74	0.32	0.24	0.27
£13/t on-site (£26/t daf)	**3.27**	0.35	2.10	0.32	0.24	0.27
Breakeven cost:£7.5/t (£15/tdaf)	**2.40**	0.35	1.22	0.32	0.24	0.27

7.5 CONCLUSIONS

Gasifier capital costs have been found to be independent of gasifier geometry. The capital costs of gasifiers of the same throughput can vary by a factor of three, depending on the level of technology and the size of company constructing the gasifier. Gasifier capital costs were found to follow conventional economies of scale, with a scale factor of 0.65.

The most important influences on gasifier profitability have been found to be feedstock cost, product selling price and throughput. Given a throughput of 2.5 tonnes/h, dry basis, the largest element of product gas cost is the feedstock cost. This is also the cost element which is the most variable , varying from £ -5/tonne to £ +34/tonne for prepared feedstock on a dry basis. [4]

Given the economic conditions of late 1985, refuse gasification for the production of producer gas is viable in the UK at throughputs of about 2.5 tonnes/h dry basis and above, in the metropolitan counties. It would be viable at throughputs of about 5 tonnes/h in other areas where refuse disposal costs are lower, if there were sufficient arisings of refuse. Straw gasification was found to be of marginal profitability, and wood gasification was found to be unprofitable, but this would change if there was a change in the relationship between feedstock cost and product gas price.

CHAPTER 8
CONCLUSIONS AND RECOMMENDATIONS

8.1 MODELLING AND DESIGN OF GASIFIERS

A new type of equilibrium model of gasification has been constructed, the adiabatic carbon boundary model (Chapter 5). This represents an ideal gasifier working at the point where the carbon in the feedstock is completely gasified. The computer program "MODEL" has been written to perform the calculations of this model, as well as those of other equilibrium models of gasification.

The adiabatic carbon boundary model has been used in a sensitivity analysis of the parameters which influence gasification. Of the parameters which can be chosen by the designer, the biomass moisture content and gasifying agent composition were found to have the largest effects on performance, whilst pressure and heat loss had a lesser effect and gasifying agent inlet temperature had a negligible effect.

The results of the carbon boundary model have been compared with the performance of a number of real gasifiers and used to construct a plot of real against ideal product heating value. Most gasifiers show results which are close to, but slightly worse than, ideality.

Fluid bed gasifiers have been modelled by using a stagewise approach, using both equilibrium modelling of gasification and experimentally derived pyrolysis data (Chapter 6). The computer program "FLMOD" has been written and used to test four stagewise models which differ in their treatment of pyrolysis. It was shown that if the pyrolysis products pass straight to the product gas without reaction then there is a large effect on the product gas composition. If only the hydrocarbon products bypass the gasification reactions then the resulting model predicts well the performance of some fluid bed gasifiers, including the Lurgi CFB gasifier. This model is simple in concept and straightforward to program on the computer.

It has been concluded that pyrolysis and the reactions of pyrolysis products play an important part in the process of fluidised bed gasification, confirming the results of other workers. Yet, even the sophisticated kinetic models of gasification developed by other workers (see Chapter 4) only take account of pyrolysis in a primitive manner.

Sensitivity analyses carried out using the stagewise model show similar trends to those predicted by the adiabatic carbon boundary model.

Both equilibrium and stagewise models can be used in the evaluation of results of real gasifiers, as shown in Chapters 5 and 6. The stagewise model may also be used in the calculation of fluid bed gasifier mass balances for design purposes, assuming that the design of the gasifier allows good mixing of feed into the fluid bed so that the volatile pyrolysis products may undergo reaction. For the sizing of gasification reactors, char gasification is the rate-limiting step and the model of Groeneveld can be used.

8.2 GASIFICATION ECONOMICS

Correlations of gasifier capital costs have been derived, based on actual cost data. The capital costs of gasifiers of the same throughput can vary by a factor of three, depending on the level of technology and the size of company constructing the gasifier. These correlations exhibit conventional economies of scale, with a scale factor of 0.65.

Two computer programs have been written to model gasifier economics. The most important influences on gasifier profitability have been found to be feedstock cost, product selling price and throughput.

Given the economic conditions of late 1985, refuse gasification for the production of producer gas is viable at throughputs of about 2.5 tonnes/h dry basis and above, in the metropolitan counties of the United Kingdom. It would be viable at throughputs of about 5 tonnes/h in other areas of the UK where refuse disposal costs are lower, if there were sufficient arisings of refuse. Straw gasification was found to be of marginal profitability, and wood gasification was found to be unprofitable.

Given a throughput of above 2.5 tonnes/h dry basis, the largest element of product gas cost is the feedstock cost. This is also the cost element which is the most variable, varying from £-5/tonne to £+34/tonne for prepared feedstock on a dry basis.

8.3 RECOMMENDATIONS FOR FUTURE WORK

The most important recommendation is that future work in gasifier modelling should concentrate on the the construction of robust models of both pyrolysis and the reactions of tar and hydrocarbon degradation. These could then be incorporated in a stagewise model of the whole gasification process.

Gasification processes are complex to analyse and difficult to understand. Consequently, it would be interesting to assess the applicability of equilibrium-based models in the evaluation of gasification processes based on feedstocks other than biomass, in particular coal, coke, oil and gas. Apart from the evaluation of the performance of existing reactors, such modelling could be used in the development of new processes by providing information about feasible operating regions.

In the field of gasifier economics, there are several areas where work would be valuable:

i. The construction of a capital cost model of gasification systems from first principles by using a conventional costing method, for example process step scoring or factorial estimation.

ii. The comparison of such a model with both existing data and the correlations developed by the author. This would enable the large variations in gasifier capital costs to be better understood.

iii. The extension of the operating cost models of gasification used by the author to include the generation and use of synthesis gas to produce liquid fuels and chemicals.

APPENDIX A

CALCULATION OF EQUILIBRIUM WITH NO SOLID CARBON

The chemical reactions assumed to be in equilibrium are:

$$CO + H_2O = CO_2 + H_2 \quad [1]$$

$$2\,H_2 + 2\,CO = CH_4 + CO_2 \quad [2]$$

Assuming ideal gas behaviour, i.e. at below 50 atmospheres approximately, these chemical reactions give rise to the following equilibrium relationships:

$$K_4 = \frac{p_{CO2}\ p_{H2}}{p_{CO}\ p_{H2O}} = \frac{y_{CO2}\ y_{H2}}{y_{CO}\ y_{H2O}} \quad [3]$$

$$K_5 = \frac{p_{CO2}\ p_{CH4}}{p^2_{CO}\ p^2_{H2}} = \frac{y_{CO2}\ y_{CH4}}{P^2\ y^2_{CO}\ y^2_{H2}} \quad [4]$$

The equilibrium equations are solved together with the element balances in order to calculate the gas composition. One mole of a gas which is not at equilibrium is assumed to react and to reach equilibrium. The molar composition of this gas is specified as $x_{0,H2}$, $x_{0,CO}$, $x_{0,CO2}$, $x_{0,CH4}$, $x_{0,N2}$, $x_{0,H2O}$. In order to reach equilibrium α moles of CO and α moles of H_2O react to form α moles of CO_2 and α moles of H_2, according to equation [1]; and 2β moles of H_2 and 2β moles of CO react to form β moles of CO_2 and β moles of CH_4, according to equation [2]. Whilst reaction [1] does not change the total number of moles of gas, equation [2] does, so it is appropriate to first estimate a value of β and then adjust the gas compositions accordingly:

$$M = 1 - 2\,\beta \quad [5]$$

$$x_{1,H2} = (\,x_{0,H2} - 2\,\beta\,)\,/\,M \quad [6]$$

$$x_{1,CO} = (\,x_{0,CO} - 2\,\beta\,)\,/\,M \quad [7]$$

$$x_{1,CO2} = (\,x_{0,CO2} + \beta\,)\,/\,M \quad [8]$$

$$x_{1,CH4} = (\,x_{0,CH4} + \beta\,)\,/\,M \quad [9]$$

$$x_{1,N2} = (\,x_{0,N2}\,)\,/\,M \quad [10]$$

$$x_{1,H2O} = (\,x_{0,H2O}\,)\,/\,M \quad [11]$$

Reaction [1] is now assumed to take place, and to reach equilibrium. The gas composition will then be:

$$y_{H2} = x_{1,H2} + \alpha \qquad [12]$$

$$y_{CO} = x_{1,CO} - \alpha \qquad [13]$$

$$y_{CO2} = x_{1,CO2} + \alpha \qquad [14]$$

$$y_{CH4} = x_{1,CH4} \qquad [15]$$

$$y_{N2} = x_{1,N2} \qquad [16]$$

$$y_{H2O} = x_{1,H2O} - \alpha \qquad [17]$$

Substituting [12] to [17] in [3] gives the following equation in α:

$$K_4 = \frac{(x_{1,CO2} + \alpha)(x_{1,H2} + \alpha)}{(x_{1,H2O} - \alpha)(x_{1,CO} - \alpha)} \qquad [18]$$

Rearranging gives a quadratic in α:

$$0 = (1 - K_4)\,\alpha^2 + (x_{1,CO2} + x_{1,H2} + K_4\, x_{1,CO} + K_4\, x_{1,H2O})\,\alpha + (x_{1,CO2}\, x_{1,H2} + K_4\, x_{1,CO}\, x_{1,H2O}) \qquad [19]$$

This quadratic is easily solved for α and the correct solution selected by calculating both possible gas compositions, and choosing the value of α which gives a gas composition with all mole fractions between 0 and +1. The calculated gas composition is then back-substituted in equation [4], giving a trial value of K_5. This trial value of K_5 is then compared with the actual value calculated from the Gibbs free energies at the required temperature (see Chapter 5, equation [5.9], p 76). The value of β can then be changed and the calculation repeated iteratively until the trial value of K_5 agrees with the calculated value, giving the equilibrium gas composition.

Nomenclature, Appendix A

K_i	Equilibrium constant of chemical reaction i
M	Total moles of gas after adjustment for reaction [2]
P	Total pressure
p_i	Partial pressure of component i
$x_{0,i}$	Initial mole fraction of component i
$x_{1,i}$	Mole fraction of component i after adjustment for reaction [2]
y_i	Mole fraction of component i at equilibrium
α	Moles of CO converted to CO_2 in reaction [1] in order to attain equilibrium of reaction [1]
β	Moles of CH_4 created by reaction [2] from CO and H_2 in order to attain equilibrium of reaction [2]

APPENDIX B

FLOW DIAGRAMS, LISTINGS AND EXAMPLE OUTPUT OF COMPUTER PROGRAMS

Program MODEL

The basic flow diagram of this program is shown as Figure B1. A complete list of subroutines and functions is given below:

Subroutines

CEPP	Subroutine used by EQCONC.
CBOUND	A dummy subroutine to allow SIMP2D to call FCARB
EQRC	Subroutine which calculates the equilibrium concentrations of a gas, with a restricted amount of carbon present.
EQRC2B	Subroutine used by EQRC.
FCARB	Subroutine which calculates the heat and mass balances of the model where excess carbon is present. Calls EQCONC to calculate equilibrium gas concentrations.
GPHASE	Subroutine which performs the restricted carbon calculation by iteration. Heat and mass balances are perfomed by calling STAGE2.
NORM	A general purpose subroutine. Takes an array, and normalises it so that the sum of the elements equals 1.
EQCONC	Subroutine which calculates the equilibrium concentrations of a gas, with excess carbon present.
SIMP2D	Subroutine for performing simplex iteration in 2 dimensions. Used to perform the iterations of the carbon boundary calculation, where the variables are the temperature (Tcb) and gasifying agent flow (Qga).
STAGE2	Subroutine to perform the heat and mass balances where the amount of carbon is limited. Calls EQRC to calculate equilibrium gas concentration. Called by GPHASE.

Functions

ASHEAT	Function which returns the sensible heat of ash at a given temperature. Used in heat balance calculations.
CALVAL	Function which calculates the HHV of a gas, to IGU standards.
DELHF	Function to calculate the heat of formation of a C-H-O material, given its HHV.
HHV	Function to calculate the HHV of a biomass, using the IGT equation.
SENSH	Function which returns the sensible heat of a given component at a given temperature. Used in heat balance calculations.

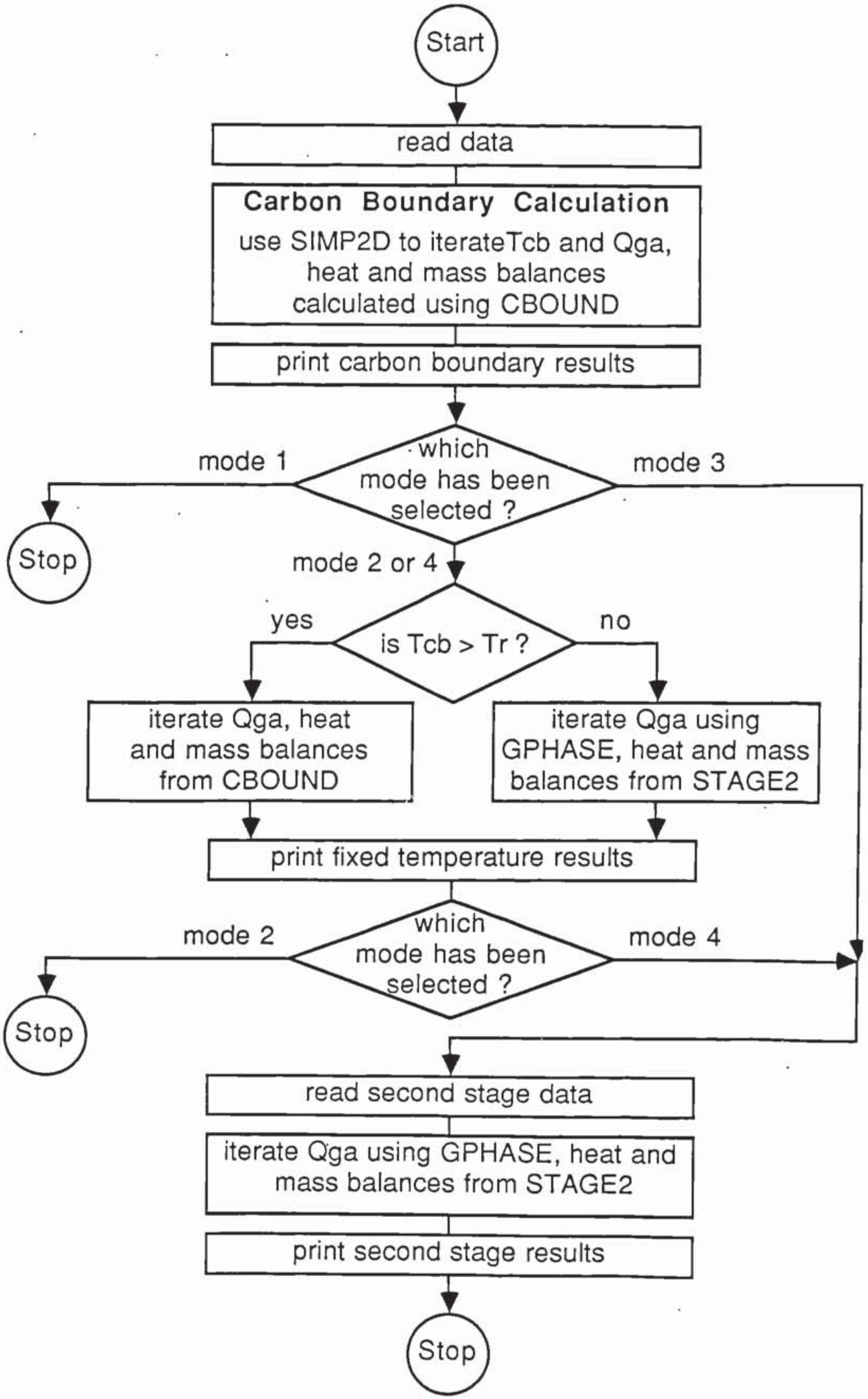


Figure B1 **Flow Diagram of Program MODEL**

```
      PROGRAM MODEL

C     PROGRAM TO FIND THE QUANTITY OF GAS REQUIRED
C     TO GASIFY 1 KMOL OF BIOMASS AT A GIVEN TEMP.
C     ASSUMING AN ADIABATIC REACTOR AND
C     EQUILIBRIUM CONCENTRATION OF OUTLET GAS
C     WITH SECOND STAGE FOR PARTIAL OXIDATION OF OUTLET GAS
C
C
C
C
C     TO USE INTERACTIVELY ASSIGN:
C                                 20  =  NEAT RESULTS FILE
C                                 21  =  *
C
C
C
C
C
C
C     DATA REQUIRED:      FOLLOW PROMPTS ON SCREEN,
C                         DATA REQUIRED VARIES ON MODE OF OPERATION
C
C

      REAL V(3,2),VRES(2),LOSSES
      REAL NITR,MWBIO,MOLFR(6),XOUT(6)
      REAL GASIN(8),DHF(8)
      REAL DHC(8),DRYMF(5)
      CHARACTER *3 COMP(8)
      CHARACTER *6 NORUN
      COMMON /FUEL/ X,Y,Z,RMB,MWBIO,BIOHF,HTCOM,ASHWT
      COMMON /SYST/ P,LOSSES,MOLFR,TOTMOL
      COMMON /INLET/ GASIN,DHF,GAST,ENTHS
      COMMON /RCARB/ XOUT,FOUT
      EXTERNAL CBOUND

C     HEATS OF FORMATION OF COMPONENTS
      DATA (DHF(I),I=1,8,1) / 0.0,110.5,393.5,74.9,0.0,241.8,
     *0.0,0.0/
C     NAMES OF COMPONENTS
      DATA (COMP(I),I=1,8,1) /"H2 ","CO ","CO2","CH4","N2 ","H2O",
     *" C ","O2 "/
C     HEATS OF COMBUSTION OF COMPONENTS
      DATA (DHC(I),I=1,8,1) /241.8,283.0,0.,802.2,0.,0.,393.5,0./

      X=1.0
      Y=1.5
      Z=0.7
      ASH=1.0

C     SET UP RESULTS FILE, AND WRITE INTRODUCTION TO
C     INTERACTIVE TERMINAL
```

```
C     N.B. NOTE THE FOLLOWING FILE NUMBERS:
C        3      TERMINAL SCREEN (DEFAULT SETTING ON THE HARRIS)
C        20      RESULTS FILE
C        21      INTERACTIVE INPUT, OR DATA FILE
      WRITE(20,'(1H1)')
      WRITE(3,'(20(/),"  ")')
      WRITE(3,'(5X,"     *** PROGRAM MODEL ***")')
      WRITE(3,'(/," This program is a thermodynamic model of a"
     *," Biomass gasifier.",/," The calculation is performed in "
     *"several ways.")')
47198 WRITE(3,'(/," Please enter the number of the ",
     *"calculation required",/,/,
     *"  1. Carbon Boundary calculation only ",/,
     *"  2. Fixed temperature calculation only",/,
     *"  3. Carbon Boundary and gas phase calculations",/,
     *"  4. Fixed temperature and gas phase calculations",/
     *"  5. Stop the program")')
C     READ THE TYPE OF CALCULATION REQUIRED, FROM
C     THE INTERACTIVE TERMINAL OR DATA FILE
      READ(21,*) MODE
      IF (MODE.LE.0.OR.MODE.GE.5) STOP

      ITOUT=0
47199 WRITE(3,'(" ENTER RUN NUMBER")')
      READ(21,'(A6)') NORUN

C     INTERACTIVE DATA ENTRY SECTION
C     (DATA MAY ALSO BE READ FROM A DATA FILE ASSIGNED TO LFN 21)
      WRITE(3,'(" ENTER DATA.... ")')
      WRITE(3,'(" ENTER PRESSURE")')
      READ(21,*) P
      WRITE(3,'(" ENTER % HEAT LOSS FROM GASIFIER")')
      READ(21,*) LOSSES
      WRITE(3,*) 'THE CURRENT BIOMASS SPECIFICATION IS:'
      WRITE(3,*) 'C',X,' H',Y,' O',Z
      WRITE(3,*) 'ASH ',ASH,'%'
      WRITE(3,*) 'ENTER 1 IF YOU WANT TO CHANGE THIS'
      READ(21,*) IFLAG
      IF (IFLAG.NE.1) GOTO 47201
      WRITE(3,'(" ENTER COMPOSITION OF CxHyOz",/
     *," ENTER x")')
      READ(21,*) X
      WRITE(3,'(" ENTER y")')
      READ(21,*) Y
      WRITE(3,'(" ENTER z")')
      READ(21,*) Z
      WRITE(3,'(" ENTER % ASH IN BIOMASS DRY ASH FREE BASIS")')
      READ(21,*) ASH
47201 MWBIO=X*12.0+Y+Z*16.0
      WRITE(3,'(" ENTER % H20 IN BIOMASS DRY ASH FREE BASIS")')
      READ(21,*) PCMOIST
      RMB=PCMOIST*MWBIO/1800.
      DO 47999 JK=1,8
       GASIN(JK)=0.0
47999 CONTINUE
```

```
      WRITE(3,'(" FIRST STAGE DATA ")')
      WRITE(3,*) 'ENTER A NUMBER TO INDICATE GASIFYING AGENT'
      WRITE(3,*) '  1    FOR AIR'
      WRITE(3,*) '  2    FOR OXYGEN'
      WRITE(3,*) '  3    TO SPECIFY ANOTHER GASIFYING AGENT'
      READ(21,*) IGATYP
      IF (IGATYP.EQ.3) THEN
      GOTO 48001
      ELSE IF (IGATYP.EQ.2)
       GASIN(8)=1.0
      ELSE IF (IGATYP.EQ.1)
       GASIN(5)=.79
       GASIN(8)=0.21
      ELSE
       STOP ": ERROR, INCORRECT IGATYP"
      END IF
      GOTO 48002
48001 WRITE(3,'(" ENTER RELATIVE QUANTITIES OF INLET GASES"
     *" BY MOLES !")')
      DO 1 I=1,8,1
      IF (I.EQ.7) THEN
        GASIN(I)=0.0
      ELSE
        WRITE(3,'(" ENTER FRACTION OF ",A3)') COMP(I)
        READ(21,*) GASIN(I)
      END IF
1     CONTINUE
48002 CALL NORM (GASIN,8)
      WRITE(3,'(" ENTER TEMPERATURE OF INLET GASES")')
      READ(21,*) GAST
      WRITE(3,'(" ENTER ENTHALPY OF INLET STEAM IN kJ/kg"
     *," FROM STEAM TABLES")')
      READ(21,*) ENTHS
      IF (MODE.EQ.2.OR.MODE.EQ.4)
        WRITE(3,'(" ENTER GASIFIER TEMPERATURE ")')
        READ(21,*) T
      END IF

C     PERFORM CALCULATIONS ON RAW DATA, TO ALLOW IT TO BE USED
C     BY LATER CALCULATIONS
      ENTHS=ENTHS-2546.6
      HTCOM=HHV(X,Y,Z,ASH)
      BIOHF=DELHF(X,Y,-MWBIO*HTCOM)
C     CALCULATE CHEMICAL POTENTIAL ENERGY ENTERING THE GASIFIER
      PCEIN=MWBIO*HTCOM
      ASHWT=ASH*MWBIO/100.0

C     CARBON BOUNDARY CALCULATION

C     SET INITIAL POINTS FOR CORNERS OF THE SIMPLEX IN THE
C     CARBON BOUNDARY CALCULATION
      V(1,1)=0.3
      V(1,2)=1193.0
      V(2,1)=0.5
      V(2,2)=798.0
```

```
      V(3,1)=0.9
      V(3,2)=673.0
      TOL=1.0
       LAB=0
C     CALL THE SIMPLEX SUBROUTINE, I.E. PERFORM THE CARBON
C     BOUNDARY ITERATION
      CALL SIMP2D(V,VRES,YRES,TOL,LAB,CBOUND)

C     OUTPUT OF INITIAL DATA AND RESULTS OF PRELIMINARY
C     CALCULATIONS
      WRITE(20,'(/,10X,18("*"))')
      WRITE(20,'(9X," RESULTS FROM MODEL ")')
      WRITE(20,'(9X," RUN NUMBER  ",A6)') NORUN
      WRITE(20,'(10X,18("*"),/,/)')
      WRITE(20,'(10X,"INPUT CONDITIONS")')
      WRITE(20,'(1H+,9X,16("_"),/)')
      WRITE(20,'(10X,"PRESSURE IS ",F5.1," atm",/)') P
      WRITE(20,'(10X,"HEAT LOSS IS",F5.1," %",/)') LOSSES
      WRITE (20,'(9X," COMPOSITION OF DRY BIOMASS ")')
      WRITE (20,'(9X," C ",F5.3," H ",F5.3," O ",F5.3,/)') X,Y,Z
      WRITE(20,'(9X," MOISTURE CONTENT OF BIOMASS ",F8.2,
     *"% dry ash free weight")') PCMOIST
      WRITE(20,'(9X," ASH CONTENT OF BIOMASS      ",F8.2,
     *"% dry ash free weight",/)') ASH
      WRITE(20,'(9X," HEAT OF COMBUSTION IS ",F8.2," MJ/kg")')
     *HTCOM
      WRITE(20,'(/,/,10X,
     *"FIRST STAGE CONDITIONS",/,1H+,9X,22("_"))')
      WRITE(20,90003) GAST,ENTHS+2546.6
      WRITE(20,90001) (COMP(I),I=1,6),COMP(8)
     *,(GASIN(I),I=1,6),GASIN(8)

C     IF ERROR HAS OCCURRED, WRITE ERROR MESSAGE TO TERMINAL
      IF (LAB.EQ.1) THEN
      WRITE(3,'(" TOO MANY ITERATIONS IN CARBON ",
     *"BOUNDARY CALCULATION")')
      WRITE(20,'(/,10X,"TOO MANY ITERATIONS IN CARBON ",
     *"BOUNDARY CALCULATION")')
      STOP
      END IF

C     CALCULATE OUTLET GAS CALORIFIC VALUE, OUTLET CHEMICAL
C     POTENTIAL ENERGY, AND HENCE EFFICIENCY OF CHEMICAL ENERGY
C     CONVERSION
      CVIGU=CALVAL(MOLFR)
      DO 86204 I=1,6
        PCEIN=PCEIN+DHC(I)*GASIN(I)*VRES(1)
86204 CONTINUE
      PCEOUT=0.0
      DO 86205 I=1,6
        PCEOUT=PCEOUT+TOTMOL*MOLFR(I)*DHC(I)
86205 CONTINUE
      CHEMEFF=100.0*PCEOUT/PCEIN
      DO 45144 I=1,5
        DRYMF(I)=MOLFR(I)
45144 CONTINUE
```

```
      CALL NORM(DRYMF,5)

C     CARBON BOUNDARY RESULTS
      WRITE(20,'(/,/,10X,"CARBON BOUNDARY CALCULATION",/,
     *1H+,9X,27("_"))')
      WRITE(20,90004) VRES(2)
      WRITE(20,90000) VRES(1),TOTMOL,TOTMOL*(1.0-MOLFR(6))
      WRITE(20,90005)
      WRITE(20,90002)  (COMP(I),I=1,6),(MOLFR(I),I=1,6)
      WRITE(20,90008)  (COMP(I),I=1,5),(DRYMF(I),I=1,5)
      WRITE(20,90006)
      WRITE(20,90002)  (COMP(I),I=1,6),(MOLFR(I)*TOTMOL,I=1,6)

      WRITE(3,'(/,/,10X,"CARBON BOUNDARY CALCULATION")')
      WRITE(3,90004)  VRES(2)
      WRITE(3,90002)  (COMP(I),I=1,6),(MOLFR(I),I=1,6)
      WRITE(20,90007) CVIGU
      WRITE(20,'(/,10X,"CHEMICAL POTENTIAL ENERGY CONVERSION ",
     *"EFFICIENCY ",F5.1,"%")') CHEMEFF

C     CHECK MODE OF CALCULATION, AND GO TO NEXT STEP
      IF (MODE.EQ.1) GOTO 47200
      WRITE(20,'(/,/,10X,"continued on next page...",/1H1,/,/)')
      WRITE(20,'(9X," RESULTS FROM MODEL ")')
      WRITE(20,'(9X," RUN NUMBER  ",A6," contd.")') NORUN
      IF (MODE.EQ.3) THEN
      TIN=VRES(2)
      GOTO 58001
      END IF

C     CONSTANT TEMPERATURE FIRST STAGE CALCULATION

      WRITE(20,'(/,/,10X,"FIXED TEMPERATURE CALCULATION ",/,
     *1H+,9X,29("_"))')
      WRITE(20,90004) T
      WRITE(3,'(/,/,10X,"FIRST STAGE CALCULATION")')
      WRITE(3,90004) T
      IF (VRES(2).EQ.T) THEN
        FLOW=VRES(2)
        GOTO 56001
      END IF
C     TEST WHETHER FREE CARBON OF GAS PHASE CALCULATION IS TO BE
C     USED
      IF (VRES(2).GT.T) THEN
C       FREE CARBON CALCULATION
        QU=VRES(1)
        QL=0.0
        CALL FCARB(T,QL,ERRORL,FREEC)
        CALL FCARB(T,QU,ERRORU,FREEC)
        J=0
C       MAIN ITERATION FOR FREE CARBON CALCULATION
        DO
C       CHECK NUMBER OF ITERATIONS
        IF (J.GT.50) THEN
          WRITE(3,'(" TOO MANY ITERATIONS IN FREE CARBON ",
```

```
     *    "CALCULATION")')
          WRITE(20,'(/,10X,"TOO MANY ITERATIONS IN FREE ",
     *    "CARBON CALCULATION")')
          GOTO 47200
        END IF
C       SPLIT RANGE OF ITERATION IN TWO
        FLOW=(QU+QL)/2.0
        CALL FCARB(T,FLOW,ERROR,FREEC)
C       CHECK WHERE SOLUTION LIES, AND SET NEW, SMALLER RANGE OF
C       ITERATION
        IF (ERRORU*ERROR.GT.0.0) THEN
          QU=FLOW
          ERORRU=ERROR
        ELSE
          QL=FLOW
          ERRORL=ERROR
        END IF

C       CHECK FOR CONVERGENCE
        ERROR=ABS(ERROR)
        UNTIL (ERROR.LT.5.0)

        WRITE(20,'(/,10X,"CHAR PRODUCED ",11X,F8.4,
     *   " mol/mol Biomass")') FREEC
        WRITE(3,'(/,10X,"CHAR PRODUCED ",F8.4,
     *  " mol/mol Biomass")') FREEC
      ELSE
C       GAS PHASE CALCULATION
C       CALL "GPHASE" TO PERFORM THIS CALCLATION
        CALL GPHASE(VRES(2),T,FLOW,LABEL)
        IF (LABEL.NE.0) THEN
         WRITE(20,'(/,10X,"ERROR ",I3," IN 1ST STAGE GAS PHASE")')
     *    LABEL
         WRITE(3,'(" ERROR ",I3," IN FIRST STAGE GAS PHASE")')
     *    LABEL
          GOTO 47200
        END IF
        DO 50007 I=1,6,1
        MOLFR(I)=XOUT(I)
50007   CONTINUE
        TOTMOL=FOUT
      FLOW=FLOW+VRES(1)
      FREEC=0.0
      END IF
56001 TIN=T
C     CALCULATE CHEMICAL ENERGY CONVERSION EFFICIENCY,AND
C     SET UP VARIABLES FOR POSSIBLE NEXT STEP OF CALCULATION
      PCEIN=MWBIO*HTCOM
      DO 86206 I=1,6
        PCEIN=PCEIN+DHC(I)*GASIN(I)*FLOW
86206 CONTINUE
      PCEOUT=0.0
      DO 86207 I=1,6
        PCEOUT=PCEOUT+TOTMOL*MOLFR(I)*DHC(I)
86207 CONTINUE
```

```
      PCEOUT=PCEOUT+FREEC*DHC(7)
      CHEMEFF=100.0*PCEOUT/PCEIN
      DO 45145 I=1,5
        DRYMF(I)=MOLFR(I)
45145 CONTINUE
      CALL NORM(DRYMF,5)

C     OUTPUT OF RESULTS FROM KNOWN TEMPERATURE CALCULATION
      WRITE(20,90000) FLOW,TOTMOL,TOTMOL*(1.0-MOLFR(6))
      WRITE(20,90005)
      WRITE(20,90002) (COMP(I),I=1,6),(MOLFR(I),I=1,6)
      WRITE(20,90008) (COMP(I),I=1,5),(DRYMF(I),I=1,5)
      WRITE(20,90006)
      WRITE(20,90002) (COMP(I),I=1,6),(MOLFR(I)*TOTMOL,I=1,6)
      CVIGU=CALVAL(MOLFR)
      WRITE(20,90007) CVIGU
      WRITE(20,'(/,10X,"CHEMICAL POTENTIAL ENERGY CONVERSION ",
     *"EFFICIENCY ",F5.1,"%")') CHEMEFF

      WRITE(3,90002) (COMP(I),I=1,6),(MOLFR(I),I=1,6)
      IF (MODE.EQ.2) GOTO 47200


C     ENTER DATA FOR SECOND STAGE CALCULATION

58001 WRITE(3,'(" ENTER TEMPERATURE FOR GAS PHASE")')
      READ(21,*) T
C     CHECK THAT PARTIAL OXIDATION TEMPERATURE IS ABOVE THE
C     CURRENT TEMPERATURE OF THE GAS STREAM
      IF (T.LT.TIN) THEN
      WRITE(3,'(" GAS PHASE CALCULATION TEMP IS TOO LOW ")')
      WRITE(20,'(/,9X" GAS PHASE CALCULATION TEMP IS TOO LOW ")')
      GOTO 47200
      END IF
C
C     ENTER GASIFYING AGENT DATA FOR GAS PHASE
C
      DO 48099 JK=1,8
       GASIN(JK)=0.0
48099 CONTINUE
      WRITE(3,*) 'ENTER A NUMBER TO INDICATE GASIFYING AGENT'
      WRITE(3,*) '  1    FOR AIR'
      WRITE(3,*) '  2    FOR OXYGEN'
      WRITE(3,*) '  3    TO SPECIFY ANOTHER GASIFYING AGENT'
      READ(21,*) IGATYP
      IF (IGATYP.EQ.3) THEN
      GOTO 48101
      ELSE IF (IGATYP.EQ.2)
       GASIN(8)=1.0
      ELSE IF (IGATYP.EQ.1)
       GASIN(5)=.79
       GASIN(8)=0.21
      ELSE
       STOP ": ERROR, INCORRECT IGATYP"
      END IF
      GOTO 48102
```

```
48101 WRITE(3,'(" ENTER RELATIVE QUANTITIES OF INLET GASES"
     *" BY MOLES !")')
      DO 48201 I=1,8,1
      IF (I.EQ.7) THEN
        GASIN(I)=0.0
      ELSE
        WRITE(3,'(" ENTER FRACTION OF ",A3)') COMP(I)
        READ(21,*) GASIN(I)
      END IF
48201     CONTINUE
48102 CALL NORM (GASIN,8)


      WRITE(3,'(" ENTER TEMPERATURE OF INLET GASES")')
      READ(21,*) GAST
      WRITE(3,'(" ENTER ENTHALPY OF INLET STEAM IN kJ/kg"
     *," FROM STEAM TABLES")')
      READ(21,*) ENTHS
      ENTHS=ENTHS-2546.6

C     SECOND STAGE CALCULATION
C     USE "GPHASE" TO PERFORM CALCULATION
      CALL GPHASE(TIN,T,FLOW,LABEL)

C     ECHO PARTIAL OXIDATION DATA TO RESULTS FILE
      WRITE(20,9002)
9002  FORMAT(/,/,10X,"GAS PHASE RESULTS",/,1H+
     *,9X,17("_"))
      WRITE(20,90001) (COMP(I),I=1,6),COMP(8)
     *,(GASIN(I),I=1,6),GASIN(8)
      WRITE(20,90003) GAST,ENTHS+2546.6
      WRITE(20,90004) T
C     IF CALCULATION IS NOT SUCCESSFUL, WRITE ERROR MESSAGE
      IF (LABEL.NE.0) THEN
        WRITE(20,'(/,10X,"ERROR ",I3," IN 2ND STAGE GAS PHASE")')
     *   LABEL
        WRITE(3,'(" ERROR ",I3," IN 2ND STAGE GAS PHASE")') LABEL
        GOTO 47200
      END IF
C     CALCULATE EFFICIENCY ETC.
      PCEIN=MWBIO*HTCOM
      DO 86208 I=1,6
        PCEIN=PCEIN+DHC(I)*GASIN(I)*FLOW
86208 CONTINUE
      PCEOUT=0.0
      DO 86209 I=1,6
        PCEOUT=PCEOUT+FOUT*XOUT(I)*DHC(I)
86209 CONTINUE
      CHEMEFF=100.0*PCEOUT/PCEIN
      DO 45146 I=1,5
        DRYMF(I)=MOLFR(I)
45146 CONTINUE
      CALL NORM(DRYMF,5)
C     WRITE RESULTS OF PARTIAL OXIDATION TO OUTPUT FILE
      WRITE(20,90000) FLOW,FOUT,FOUT*(1.0-XOUT(6))
```

```
      WRITE(20,90005)
      WRITE(20,90002) (COMP(I),I=1,6),(XOUT(I),I=1,6)
      WRITE(20,90008) (COMP(I),I=1,5),(DRYMF(I),I=1,5)
      WRITE(20,90006)
      WRITE(20,90002) (COMP(I),I=1,6),(XOUT(I)*FOUT,I=1,6)
      CVIGU=CALVAL(XOUT)
      WRITE(20,90007) CVIGU

      WRITE(3,'(/,/,10X,"SECOND STAGE CALCULATION")')
      WRITE(3,90004) T
      WRITE(3,90002) (COMP(I),I=1,6),(XOUT(I),I=1,6)
      WRITE(20,'(/,10X,"CHEMICAL POTENTIAL ENERGY CONVERSION ",
     *"EFFICIENCY ",F5.1,"%")') CHEMEFF

C     RETURN TO START TO ALLOW ANOTHER RUN
47200 WRITE(20,'(1H1)')
      GOTO 47198


C     FORMAT STATEMENTS
90000 FORMAT(/,10X,"GASIFYING AGENT FLOWRATE ",F8.4,
     *" mol/mol Biomass",/,10X,"OUTLET GAS FLOWRATE",6X,F8.4,
     *" mol/mol Biomass",/,10X,"OUTLET FLOW (DRY BASIS)  ",
     *F8.4," mol/mol Biomass")
90001 FORMAT(/,10X,"GASIFYING AGENT COMPOSITION ",/,11X,
     *7(A3,5X),/,10X,7(F6.4,2X))
90002 FORMAT(11X,6(A3,5X),/,10X,6(F6.4,2X))
90003 FORMAT(/,10X,"GASIFYING AGENT TEMPERATURE ",F8.2," K",/,
     *10X,"GASIFYING STEAM ENTHALPY ",3X,F8.2," kJ/kg")
90004 FORMAT(/,10X,"OUTLET TEMPERATURE",10X,F8.2," K")
90005 FORMAT(/,10X,"OUTLET GAS COMPOSITION ")
90006 FORMAT(/,10X,"OUTLET GAS FLOWRATES, mol/mol Biomass")
90007 FORMAT(/,10X,"OUTLET GAS CALORIFIC VALUE (IGU standard)",
     *F8.4," MJ/m**3")
90008 FORMAT(/,10X,"OUTLET GAS COMPOSITION, DRY BASIS",/,
     *11X,5(A3,5X),/,10X,5(F6.4,2X))
      END
C
C
C
C
C
C
      SUBROUTINE FCARB(T,Q,ERROR,FREEC)
      REAL NOR,MOLFR(6),LHM,LOSSES
      REAL GASIN(8),DHF(8)
      REAL MWBIO
      COMMON /FUEL/ X,Y,Z,RMB,MWBIO,BIOHF,HTCOM,ASHWT
      COMMON /SYST/ P,LOSSES,MOLFR,TOTMOL
      COMMON /INLET/ GASIN,DHF,GAST,ENTHS
C     SUBROUTINE TO DO HEAT BALANCING, FOR THE CASE WHERE BIOMASS
C     IS GASIFIED, ASSUMING THAT THERE IS SOME FREE CARBON IN THE
C     EQUILIBRIUM CALCULATION
C     Q & T ARE TRANSFERRED AS ARGUMENTS, OTHER DATA ON BIOMASS
C     AND GASIFYING AGENT ARE TRANSFERRED BY COMMON BLOCKS
```

```
C     Q=GAS FLOW PER MOLE DRY BIOMASS
C     T=TEMPERATURE
C
C     CALCULATE TOTAL MOLES OF C,H,O,N
      TOTO=Z+Q*(GASIN(2)+2.0*GASIN(3)+GASIN(6)+2.0*GASIN(8))+RMB
      TOTC=X+Q*(GASIN(2)+GASIN(3)+GASIN(4))
      TOTN=Q*GASIN(5)*2.0
      TOTH=Y+Q*2.0*(GASIN(1)+GASIN(4)*2.0+GASIN(6))+2.0*RMB
C
C     CALCULATE H:O & N:O RATIOS, AND CALL EQCONC TO CALCULATE
C     EQUILIBRIUM
      NOR=TOTN/TOTO
      HOR=TOTH/TOTO
      CALL EQCONC(HOR,NOR,P,T,MOLFR,LAB)
C     PERFORM MASS BALANCE TO CALCULATE TOTAL GAS MOLES, AND FREE
C     CARBON
      TOTMOL=TOTO/(MOLFR(2)+2.0*MOLFR(3)+MOLFR(6))
      FREEC=TOTC-TOTMOL*(MOLFR(2)+MOLFR(3)+MOLFR(4))
C
C     CALCULATION OF HEAT BALANCE TERMS
C
C     SENSIBLE HEAT OF INLET GASES
      SHIG=0.0
      DO 1 I=1,8,1
      IF (I.EQ.6) THEN
      SHIG=SHIG+(GASIN(6)*18.0*ENTHS*Q)
      ELSE
      SHIG=SHIG+SENSH(GAST,I)*Q*GASIN(I)
      END IF
1     CONTINUE
C
C     HEAT OF REACTION
      PRODHR=0.0
      DO 2 I=1,6,1
      PRODHR=PRODHR-TOTMOL*MOLFR(I)*DHF(I)
2     CONTINUE
      REACTHR=BIOHF
      DO 3 I=1,8,1
      REACTHR=REACTHR-Q*GASIN(I)*DHF(I)
3     CONTINUE
      REACTHR=REACTHR-RMB*DHF(6)
      DELTHR=1000.0*(PRODHR-REACTHR)
C
C     LATENT HEAT OF MOISTURE IN BIOMASS
      LHM=RMB*18.0*2441.8
C
C     SENSIBLE HEAT OF OUTLET GAS
      SHOG=0.0
      DO 37180 I=1,6,1
      SHOG=SHOG+MOLFR(I)*SENSH(T,I)
37180 CONTINUE
      SHOG=SHOG*TOTMOL
C
C     ADD SENSIBLE HEAT OF CHAR TO THAT OF GAS
      IF (FREEC.GT.0.0) THEN
      SHOG=SHOG+FREEC*SENSH(T,7)
```

```
      END IF
C
C     ADD SENSIBLE HEAT OF ASH TO THAT OF GAS
      SHOG=ASHWT*ASHEAT(T)+SHOG
C
C     ALLOW FOR % HEAT LOSS
      HLOSS-HTCOM*MWBIO*LOSSES*10.0
C
C     OVERALL HEAT BALANCE, TO CALCULATE ERROR
      ERROR=DELTHR-SHIG+SHOG+LHM+HLOSS
      RETURN
      END
C
C
C
C
C
      FUNCTION ASHEAT(T)
      REAL T,A,B,C,D

      DATA A,B,C,D /0.81168,0.000171816,8081.5,284.381/
      ASHEAT=A*T+B*T*T+C/T-D
      RETURN
      END
C
C
C
C
      SUBROUTINE EQRC(MOLIN,T,P,MOLOUT,LABEL)
      REAL MOLIN(6),MOLOUT(6),K(2),KOUT,B(5,6),G(6)
C
C     SUBROUTINE TO CALCULATE C-H-O EQUILIBRIA
C     WITH LIMITED CARBON
C
      DATA ((B(J,I),J=1,5,1),I=1,6,1) /
     *0.00000, 0.00000, 0.00000, 0.00000, 0.00000,
     *13.61200E3, 1.83170,-2.75840E-3, 0.65360E-6,-0.78772E-10,
     *47.28000E3, 0.13220,-0.94025E-3, 0.45112E-6,-0.91901E-10,
     * 8.37220E3,-1.07690,-5.64350E-3, 2.90460E-6,-5.23510E-10,
     * 0.00000, 0.00000, 0.00000, 0.00000, 0.00000,
     * 28.78000E3,-0.69477,-1.42830E-3, 0.74925E-6,-1.37850E-10/
C
C
      DO 50 I=1,6,1
      G(I)=B(1,I)/T+B(2,I)*LOG(T)+B(3,I)*T+B(4,I)*T*T+B(5,I)*T**3
50    CONTINUE
C     CALCULATE EQUILIBRIUM CONSTANTS K(1) & K(2)
      K(1)=EXP(G(3)+G(1)-G(6)-G(2))
      K(2)=EXP(G(3)+G(4)-2.0*(G(1)+G(2)))
C
      LABEL=0
      J=0
C     SET UPPER AND LOWER BOUNDS FOR THE POSSIBLE VALUES OF BETA
      BETAL=-MOLIN(4)
      BETAU=(MOLIN(1)+MOLIN(2))/2.0
```

```
300   J=J+1
C     IF THERE ARE TOO MANY ITERATIONS, SET LABEL TO SHOW
C      ERROR,AND RETURN
      IF (J.GT.25) THEN
      LABEL=LABEL+100
      RETURN
      END IF
      CALL EQRC2B(MOLIN,MOLOUT,ALPHA,BETAL,K(1),KOUT,P,LABEL)
C     ADJUST LOWER BOUND SO THAT CALCULATION IS POSSIBLE
      IF (LABEL.NE.0) THEN
      BETAL=BETAL*.95
      GOTO 300

      END IF
C     ERRORL IS THE ERROR BETWEEN THE CALCULATED &
C     TRIAL EQUILIBRIUM CONSTANTS
      ERRORL=K(2)-KOUT
C
      J=0
400   J=J+1
      IF (J.GT.25) THEN
      LABEL=LABEL+200
      RETURN
      END IF
      CALL EQRC2B(MOLIN,MOLOUT,ALPHA,BETAU,K(1),KOUT,P,LABEL)
C     ADJUST UPPER BOUND SO THAT THE CALCULATION WILL WORK
      IF (LABEL.NE.0) THEN
      BETAU=BETAU*.9
      GOTO 400
      END IF
      ERRORU=K(2)-KOUT
C     CHECK THAT BETAU & BETAL ARE EITHER SIDE OF THE SOLUTION,
C     AND ADJUST IF THEY ARE NOT
      IF (ERRORU*ERRORL.GT.0.0) THEN
      BETAU=BETAU*1.05
      GOTO 400
      END IF
C
      J=0
C
C     MAIN ITERATION LOOP
C     THIS WORKS BY SPLITTING THE INTERVAL BETWEEN THE UPPER &
C     LOWER BOUNDS, AND THEN DECIDING WHICH HALF THE SOLUTION LIES
C     IN THE INTERVAL IS THEN REDUCED TO HALF THE ORIGINAL, AND
C     THE PROCESS REPEATED TO THE DESIRED ACCURACY
      DO
      J=J+1
      IF (J.GT.75) THEN
      LABEL=300
      RETURN
      END IF
C     FIND THE MID-POINT OF THE INTERVAL
      BETA=(BETAL+BETAU)/2.0
C     PERFORM THE HEAT BALANCE FOR THE MID POINT
      CALL EQRC2B(MOLIN,MOLOUT,ALPHA,BETA,K(1),KOUT,P,LABEL)
```

```
      IF (LABEL.NE.0) THEN
      LABEL=400+LABEL
      RETURN
      END IF
      ERROR=K(2)-KOUT
C
C     TEST FOR WHICH HALF OF THE INTERVAL THE SOLUTION IS IN
C     THEN SET THE NEW UPPER & LOWER BOUNDS
      IF (ERROR*ERRORU.GT.0.0) THEN
      ERRORU=ERROR
      BETAU=BETA
      ELSE
      ERRORL=ERROR
      BETAL=BETA
      END IF
C
C     CHECK IF REQUIRED ACCURACY HAS BEEN REACHED
      UNTIL (ABS(K(2)-KOUT)/K(2).LT.0.0001)
      RETURN
      END
C
C
C
C
      SUBROUTINE EQRC2B(MOLIN,MOLOUT,ALPHA,BETA,K1,K2,P,LABEL)
      REAL MOLIN(6),MOLOUT(6),K1,K2,X(6,2),ALP(2)
      INTEGER IFLAG(2)
C     SUBROUTINE TO PERFORM QUADRATIC SOLUTION OF EQUILIBRIUM
C     EQUATIONS FOR SUBROUTINE EQRC
      IFLAG(1)=0
      IFLAG(2)=0
      ICORR=0
C     SET UP INITIAL CONCENTRATIONS, ALLOWING FOR CONVERSION OF
C     BETA IN SECOND EQUILIBRIUM REACTION
      A1=MOLIN(3)+BETA
      A2=MOLIN(1)-2.0*BETA
      A3=MOLIN(2)-2.0*BETA
      A4=MOLIN(6)
      TOTAL=1.0-2.0*BETA
C     CALCULATE QUADRATIC COEFFICIENTS
      A=1.0-K1
      B=A1+A2+K1*(A3+A4)
      C=A1*A2-A3*A4*K1
      BRCKT=B*B-4.0*A*C
C     CHECK QUADRATIC HAS A REAL SOLUTION, OTHERWISE SET ERROR
C     FLAG, AND RETURN
      IF (BRCKT.LT.0.0) THEN
      LABEL=1
      RETURN
      END IF
      BRCKT=SQRT(BRCKT)
C     CALCULATE BOTH POSSIBLE QUADRATIC SOLUTIONS
      ALP(1)=(-B+BRCKT)/(2.0*A)
      ALP(2)=(-B-BRCKT)/(2.0*A)
C     CALCULATE A SET OF POSSIBLE COMPOSITIONS FROM EACH SOLUTION
      DO 1 I=1,2
```

```
      X(1,I)=MOLIN(1)+ALP(I)-2.0*BETA
      X(2,I)=MOLIN(2)-ALP(I)-2.0*BETA
      X(3,I)=MOLIN(3)+ALP(I)+BETA
      X(4,I)=MOLIN(4)+BETA
      X(5,I)=MOLIN(5)
      X(6,I)=MOLIN(6)-ALP(I)
C
C     TEST EACH SET OF COMPOSITIONS TO FIND THE CORRECT ONE
      DO 2 J=1,6
      X(J,I)=X(J,I)/TOTAL
      IF (X(J,I).LT.0.0.OR.X(J,I).GT.1.0) IFLAG(I)=1
2     CONTINUE
1     CONTINUE
      IF (IFLAG(1).EQ.1) THEN
      IF (IFLAG(2).EQ.1) THEN
      LABEL=2
      RETURN
      ELSE
      ICORR=2
      END IF
      ELSE
      IF (IFLAG(2).EQ.1) THEN
      ICORR=1
      ELSE
      LABEL=3
      RETURN
      END IF
      END IF
C
C     SET ALPHA AND OUTLET MOLE FRACTIONS TO THE CORRECT SOLUTION,
C     AS FOUND BY THE ABOVE TESTS
      DO 3 I=1,6
      MOLOUT(I)=X(I,ICORR)
3     CONTINUE
      ALPHA=ALP(ICORR)
C
C     CALCULATE A TRIAL VALUE OF K2 FOR USE BY EQRC
      K2=(MOLOUT(4)*MOLOUT(3))/(MOLOUT(1)*MOLOUT(2)*P)**2
      LABEL=0
      RETURN
      END
C
C
C
C
C
C
C
      SUBROUTINE CBOUND(V,YOUT)
      REAL V(2),ERROR,FREEC
C     DUMMY SUBROUTINE TO ALLOW "SIMP2D" TO CALL "FCARB"
      CALL FCARB(V(2),V(1),ERROR,FREEC)
      YOUT=ABS(ERROR)+1.0E5*ABS(FREEC)
      RETURN
      END
```

```
C
C
      SUBROUTINE STAGE2(TIN,T,Q,ERROR,LABEL)
      REAL XOUT(6),MOLFR(6),INFLOW(6),LOSSES
      REAL GASIN(8),DHF(8)
      COMMON /SYST/ P,LOSSES,MOLFR,TOTMOL
      COMMON /INLET/ GASIN,DHF,GAST,ENTHS
      COMMON /RCARB/ XOUT,FOUT
C
C     SUBROUTINE TO DO ENERGY BALANCING FOR RESTRICTED CARBON
C     EQUILIBRIA, USING SUBROUTINE EQRC TO CALCULATE EQUILIBRIA
C
C     CALCULATE TOTAL MOLES OF EACH COMPONENT IN
      DO 1 I=1,6,1
      INFLOW(I)=MOLFR(I)*TOTMOL
1     CONTINUE
C
C     INLET GAS SENSIBLE HEAT
      SHIG=0.0
      DO 3 I=1,6,1
      SHIG=SHIG+SENSH(TIN,I)*INFLOW(I)
3     CONTINUE
C
C     HEAT OF FORMATION OF INLET GAS
      HFIG=0.0
      DO 101 I=1,6,1
      HFIG=HFIG-INFLOW(I)*DHF(I)
101   CONTINUE
C
C     ADD GASIFYING AGENT TO GAS
      DO 201 I=1,6,1
      INFLOW(I)=INFLOW(I)+GASIN(I)*Q
201   CONTINUE
C     ADD O2 BY REACTING IT WITH CO TO FORM CO2
      INFLOW(3)=INFLOW(3)+2.0*GASIN(8)*Q
      INFLOW(2)=INFLOW(2)-2.0*GASIN(8)*Q
C     CALCULATE TOTAL MOLES OF CARBON ENTERING
      CARBIN=INFLOW(2)+INFLOW(3)+INFLOW(4)
C
C     NORMALISE REVISED INLET FLOWS TO GIVE REVISED INLET
C     COMPOSITION
      CALL NORM(INFLOW,6)
C
C     CALCULATE EQUILIBRIUM
      LABEL=0
      CALL EQRC(INFLOW,T,P,XOUT,LABEL)
C     IF EQUILIBRIUM CALCULATION HAS FAILED, SET ERROR FLAG, AND
C     RETURN
      IF (LABEL.NE.0) THEN
      LAB=LABEL
      RETURN
      END IF
C
C     MASS BALANCE TO CALCULATE TOTAL GAS MOLES OUT
C
```

```
C       FOUT*CARBOUT=CARBIN
        CARBOUT=XOUT(2)+XOUT(3)+XOUT(4)
        FOUT=CARBIN/CARBOUT
C
C       OUTLET GAS SENSIBLE HEAT
        SHOG=0.0
        DO 4 I=1,6,1
        SHOG=SHOG+SENSH(T,I)*FOUT*XOUT(I)
4       CONTINUE
C
C       GASIFYING AGENT SENSIBLE HEAT
        SHGA=0.0
        DO 102 I=1,8,1
        IF (I.EQ.6) THEN
        SHGA=SHGA+Q*GASIN(6)*18.0*ENTHS
        ELSE
        SHGA=SHGA+Q*GASIN(I)*SENSH(GAST,I)
        END IF
102     CONTINUE
C
C       HEAT OF FORMATION OF OUTLET GAS
        HFOG=0.0
        DO 103 I=1,6,1
        HFOG=HFOG-XOUT(I)*DHF(I)*FOUT
103     CONTINUE
C
C       HEAT OF FORMATION OF GASIFYING AGENT
        HFGA=0.0
        DO 104 I=1,8,1
        HFGA=HFGA-Q*DHF(I)*GASIN(I)
104     CONTINUE
C
C       HEAT OF REACTION
        DELTHR=1000.0*(HFOG-HFIG-HFGA)
C
C       CALCULATION OF ERROR IN HEAT BALANCE
        ERROR=DELTHR-SHIG-SHGA+SHOG
C
        LAB=0
        RETURN
        END
C
C
C
C
        SUBROUTINE NORM(X,N)
        REAL X(N),TOTAL
C
C       SUBROUTINE TO NORMALISE COMPOSITIONS
C
        TOTAL=0.0
        DO 1000 I=1,N,1
        TOTAL=TOTAL+X(I)
1000    CONTINUE
        DO 1001 I=1,N,1
        X(I)=X(I)/TOTAL
```

```
1001  CONTINUE
      RETURN
      END
C
C
      FUNCTION CALVAL(X)
      REAL X(6),CV(5)
C     A FUNCTION TO FIND THE CALORIFIC VALUE OF A GAS IN MJ/m**3
C     DRY BASIS, INTERNATIONAL GAS UNION STANDARD
C     BY MULTIPLYING THE I.G.U. CALORIFIC VALUE FOR EACH COMPONENT
C     BY ITS MOLE FRACTION, DRY BASIS
      DATA (CV(I),I=1,5,1) /12.10,11.97,0.0,37.69,0.0/
      TOTAL=0.0
      CALVAL=0.0
      DO 1 I=1,5,1
      TOTAL=TOTAL+X(I)
1     CONTINUE
      DO 2 I=1,5,1
      CALVAL=CALVAL+X(I)*CV(I)/TOTAL
2     CONTINUE
      RETURN
      END
C
C
C
C
      FUNCTION HHV(X,Y,Z,ASH)
      REAL X,Y,Z,TOTM,MASSX,MASSY,MASSZ
C
C     FUNCTION TO CALCULATE HIGHER HEATING VALUE OF BIOMASS
C     USING IGT EQUATION
C
      TOTM=X*12.0+Y+Z*16.0
      A=ASH/100.0
C     CONVERT CxHyOz FORM OF COMPOSITION TO MASS %
      MASSX=(1.0-A)*1200.0*X/TOTM
      MASSY=(1.0-A)*100.0*Y/TOTM
      MASSZ=(1.0-A)*1600.0*Z/TOTM
C
C     USE IGT EQUATION TO CALCULATE HIGHER HEATING VALUE
      HHV=146.58*MASSX+568.78*MASSY+29.45-51.53*MASSZ-6.58*ASH
C     CONVERT FROM BTU/LB TO MJ/KG
      HHV=HHV*2.326E-3
      RETURN
      END
C
      FUNCTION DELHF(X,Y,HC)
C     FUNCTION TO CALCULATE HEAT OF FORMATION
C     FOR A CxHyOz MATERIAL
C     GIVEN HEAT OF COMBUSTION
      REAL X,Y,HC
      DELHF=-Y*285.8/2-X*393.5-HC
      RETURN
      END
C
```

```
C
C
      FUNCTION SENSH(T,I)
C     FUNCTION TO CALCULATE SENSIBLE HEAT
C     FOR ONE MOLE OF COMPONENT I
      REAL C(4,8)
      DATA ((C(J,I),J=1,4,1),I=1,8,1) /
     * 6.62000, 0.00081, 0.00000, 0.00000,
     *  6.60000, 0.00120, 0.00000, 0.00000,
     * 10.34000, 0.00274, 0.00000, 195500.0,
     *  5.34000, 0.01150, 0.00000, 0.00000,
     * 6.50000, 0.00100, 0.00000, 0.0000,
     * 8.2200,  0.00015, 0.00000134, 0.00000,
     * 2.67300, 0.002617, 0.0000, 116900.0,
     *  8.27000, 0.000258,0.00000, 187700.0/
      TD=298.15
C     CALCULATE ENTHALPY AT T
      HT=C(1,I)*T+C(2,I)*T*T/2.0+C(3,I)*T*T*T/3.0+C(4,I)/T
C     SUBTRACT ENTHALPY AT BASIS TEMPERATURE, 298.15 KELVIN
      HT=HT-C(1,I)*TD-C(2,I)*TD*TD/2.0-C(3,I)*TD*TD*TD/3.0
     *  -C(4,I)/TD
C     CONVERT FROM CALORIES/MOL TO KJ/KMOL
      SENSH=HT*4.1868
      RETURN
      END
C
C
C
      SUBROUTINE EQCONC(HORR,NOR,P,T,X,LABEL)
C     SUBROUTINE TO CALCULATE EQUILIBRIUM CONCENTRATIONS
C     FROM CARBON HYDROGEN OXYGEN SYSTEMS
C     USING METHOD OF BARON, PORTER, & HAMMOND
      REAL K(3),PP(6),X(6),G(6)
      REAL B(5,6)
      REAL PPL(6),PPU(6),PPM(6),NOR
C
C
      DATA ((B(J,I),J=1,5,1),I=1,6,1) /
     *0.00000, 0.00000, 0.00000, 0.00000, 0.00000,
     *13.61200E3, 1.83170,-2.75840E-3, 0.65360E-6,-0.78772E-10,
     *47.28000E3, 0.13220,-0.94025E-3, 0.45112E-6,-0.91901E-10,
     * 8.37220E3,-1.07690,-5.64350E-3, 2.90460E-6,-5.23510E-10,
     * 0.00000, 0.00000, 0.00000, 0.00000, 0.00000,
     * 28.78000E3,-0.69477,-1.42830E-3, 0.74925E-6,-1.37850E-10/
C
C
C
C
C     CALCULATION OF GIBBS FREE ENERGIES
      DO 1000 I=1,6,1
      G(I)=B(1,I)/T+B(2,I)*LOG(T)+B(3,I)*T+B(4,I)*T*T+B(5,I)*T**3
1000  CONTINUE
C     CALCULATE EQUILIBRIUM CONSTANTS USING FREE ENERGIES
      K(1)=EXP(G(1)+G(2)-G(6))
      K(2)=EXP(G(1)+G(3)-G(2)-G(6))
```

```
      K(3)=EXP(G(4)-2*G(1))
      LABEL=0
      J=0
C
C     SET UPPER AND LOWER BOUNDS FOR ITERATION
      HRU=1.0E6
      HRL=1.0E-6
      CALL CEPP(HRL,NOR,K,P,PPL,HORL,L)
      IF (L.EQ.1) THEN
      LABEL=1
      RETURN
      END IF
100   CONTINUE
      CALL CEPP(HRU,NOR,K,P,PPU,HORU,L)
C     CHECK THAT UPPER BOUND HAS BEEN FOUND IN LESS THAN 50
C     ITERATIONS
      IF (L.EQ.1) THEN
      J=J+1
      IF (J.GT.50) THEN
      LABEL=1
      RETURN
      END IF
C     ADJUST VALUE OF UPPER BOUND SO THAT CALCULATION "CEPP"
C     MAY BE PERFORMED
110   HRU=HRU/10.0
      GOTO 100
      END IF
C     CHECK THAT UPPER AND LOWER BOUNDS ARE EITHER SIDE OF
C     SOLUTION
      IF ((HORU-HORR)*(HORL-HORR).GT.0.0) THEN
      HRU=HRU*9.0
      GOTO 100
      END IF
C
C     MAIN ITERATION TO GIVE CORRECT H TO O RATIO
C
      DO
C     SPLIT RANGE OF ITERATION IN TWO, BY TAKING GEOMETRIC
C     MEAN OF UPPER & LOWER BOUNDS
      HRM=SQRT(HRU*HRL)
      CALL CEPP(HRM,NOR,K,P,PPM,HORM,L)
      IF (L.EQ.1) THEN
      LABEL=1
      RETURN
      END IF
C     CHECK WHETHER SOLUTION IS ABOVE OR BELOW THE MID-POINT,
C     IN ORDER TO SET BOUNDS ON NEW, SMALLER RANGE
      IF ((HORM-HORR)*(HORL-HORR).LT.0.0) THEN
      HRU=HRM
      HORU=HORM
      ELSE
      HRL=HRM
      HORL=HORM
      END IF
C
C     CHECK IF REQUIRED ACCURACY HAS BEEN OBTAINED
```

```
      TOL=ABS((HORU-HORL)/HORR)
      UNTIL (TOL.LT.0.00001)
      HRM=SQRT(HRU*HRL)
      CALL CEPP(HRM,NOR,K,P,PP,HOR,L)
C
C     CONVERSION OF PARTIAL PRESSURES TO MOLE FRS.
4000  DO 4010 I=1,6,1
      X(I)=PP(I)/P
4010  CONTINUE
      RETURN
      END
C
C
      SUBROUTINE CEPP(H,NR,EK,P,PP,OR,L)
C     SOLUTION OF QUADRATIC EQUATION IN P(H2) TO GENERATE A SET OF
C     PARTIAL PRESSURES, GIVEN A H2:H2O RATIO
C     THIS IS USED BY EQCONC
      REAL EK(3),PP(6),NR
      L=0
C     CALCULATE COEFFICIENTS OF QUADRATIC EQUATION
      A=EK(3)
      B=1+H*(1+NR/2)
      C=EK(1)*H*(1+NR/2)+EK(1)*EK(2)*H*H*(1+NR)-P
      BRCKT=B*B-4*A*C
      IF (BRCKT.LT.0.0) THEN
      L=1
      OR=0
      GOTO 6000
      END IF
C     CALCULATE BOTH POSSIBLE SOLUTIONS TO QUADRATIC
      XP=(-B+SQRT(BRCKT))/(2*A)
      XM=(-B-SQRT(BRCKT))/(2*A)
C     CHOOSE THE REQUIRED SOLUTION
      IF (XP.GT.0.0.AND.XP.LE.P) THEN
      PP(1)=XP
      ELSE IF (XM.GE.0.0.AND.XM.LE.P)
      PP(1)=XM
      ELSE
      L=1
      OR=0
      END IF
C     FROM THE SOLUTION, CALCULATE OTHER PARTIAL PRESSURES
      PP(2)=EK(1)*H
      PP(3)=EK(1)*EK(2)*H*H
      PP(4)=EK(3)*PP(1)*PP(1)
      PP(6)=H*PP(1)
      PP(5)=NR*(PP(2)+PP(6)+2*PP(3))/2
C     CALCULATE THE TRIAL H:O RATIO
      OR=2*(PP(1)+PP(6)+2*PP(4))/(PP(2)+PP(6)+2*PP(3))
6000  CONTINUE
      RETURN
      END
C
C
C
```

```
      SUBROUTINE SIMP2D(V,VRES,YRES,TOL,LABEL,FUNCT)
      REAL V(3,2),Y(3),VRES(2)
      REAL YRES,VCENT(2),VR(2),VE(2),VDUM(2),VC(2)
      INTEGER LABEL
C     SUBROUTINE TO PERFORM MODIFIED SIMPLEX ITERATION IN TWO
C     DIMENSIONS. LOOKS FOR A MINIMUM, WHERE THE VALUE OF
C     THE DEPENDENT VARIABLE EQUALS ZERO
C
C     FIND THE VALUE OF THE DEPENDENT VARIABLE AT THE THREE
C     CORNERS OF THE SIMPLEX, BY CALLING THE SUBROUTINE SUPPLIED
C     AS AN ARGUMENT
      DO 1 I=1,3,1
      DO 2 J=1,2,1
      VDUM(J)=V(I,J)
2     CONTINUE
      CALL FUNCT(VDUM,Y(I))
1     CONTINUE
      ILOOP=0
C
C     MAIN ITERATIVE LOOP
      DO
      ILOOP=ILOOP+1
C     CHECK CONVERGENCE IS NOT TOO SLOW
      IF (ILOOP.GT.100) THEN
      LABEL=1
      RETURN
      END IF
      IM=1
      IT=1
      IS=1
C     SORT THE THREE POINTS TO FIND THE HIGHEST AND LOWEST
C     VALUES OF THE DEPENDENT VARIABLE
      DO 3 I=2,3,1
      IF (Y(I).LT.Y(IM)) THEN
      IM=I
      ELSE IF (Y(I).GT.Y(IS))
      IS=I
      END IF
3     CONTINUE
      DO 35 I=1,3,1
      IF (I.NE.IS.AND.I.NE.IM) THEN
      IT=I
      END IF
35    CONTINUE
C     IF PRINTING OF EACH ITERATION IS REQUIRED, THEN WRITE TO LFN
6
      IF (LABEL.EQ.2) THEN
      WRITE(6,'("SIMP2D...   ITERATION ",I3)') ILOOP
      WRITE (6,'(" IM",I2," IS",I2," IT",I2)') IM,IS,IT
      WRITE(6,'(" I   V(I,1)  ",3(I3,"  ",E12.4))')
     * (I,V(I,1),I=1,3,1)
      WRITE(6,'(" I   V(I,2)  ",3(I3,"  ",E12.4))')
     * (I,V(I,2),I=1,3,1)
      WRITE(6,'(" I   Y(I)    ",3(I3,"  ",E12.4))')
     * (I,Y(I),I=1,3,1)
```

```
      WRITE(6,'(/)')
      END IF
C     FIND THE CENTROID OF THE SIMPLEX
      DO 4 I=1,2,1
      VCENT(I)=(V(IM,I)+V(IT,I))/2.0
4     CONTINUE
C
C     REFLECTION, CONTRACTION, OR EXPANSION OF SIMPLEX, AS
C     REQUIRED
C
      DO 5 I=1,2,1
      VR(I)=2.0*VCENT(I)-V(IS,I)
5     CONTINUE
      CALL FUNCT(VR,YR)
      IF (YR.GT.Y(IM).AND.Y(IT).GT.YR) THEN
      DO 6 I=1,2,1
      V(IS,I)=VR(I)
6     CONTINUE
      Y(IS)=YR
      ELSE IF (YR.LT.Y(IM)) THEN
      DO 7 I=1,2,1
      VE(I)=2.0*VR(I)-VCENT(I)
7     CONTINUE
      CALL FUNCT(VE,YE)
      IF (YE.LT.YR) THEN
      DO 8 I=1,2,1
      V(IS,I)=VE(I)
8     CONTINUE
      Y(IS)=YE
      ELSE
      DO 9 I=1,2,1
      V(IS,I)=VR(I)
9     CONTINUE
      Y(IS)=YR
      END IF
      ELSE
      IF (YR.GT.Y(IS)) THEN
      DO 10 I=1,2,1
      VDUM(I)=V(IS,I)
10    CONTINUE
      YDUM=Y(IS)
      ELSE
      DO 11 I=1,2,1
      VDUM(I)=VR(I)
11    CONTINUE
      YDUM=YR
      END IF
      DO 12 I=1,2,1
      VC(I)=0.5*VDUM(I)+0.5*VCENT(I)
12    CONTINUE
      CALL FUNCT(VC,YC)
      IF (YC.LT.YDUM) THEN
      DO 13 I=1,2,1
      V(IS,I)=VC(I)
13    CONTINUE
      Y(IS)=YC
```

```
      ELSE
      DO 14 I=1,2,1
      IF (I.NE.IM) THEN
      DO 15 J=1,2,1
      V(I,J)=0.5*(V(I,J)+V(IM,J))
      VDUM(J)=V(I,J)
15    CONTINUE
      CALL FUNCT(VDUM,Y(I))
      END IF
14    CONTINUE
      END IF
      END IF
      YTEST=0.0
C     CHECK FOR CONVERGENCE TO REQIRED TOLERENCE
      DO 16 I=1,3,1
      YTEST=YTEST+ABS(Y(I))
16    CONTINUE
      UNTIL (YTEST.LT.TOL)
      VRES(1)=0.0
      VRES(2)=0.0
      DO 19 I=1,3,1
      DO 20 J=1,2,1
      VRES(J)=V(I,J)/3.0+VRES(J)
20    CONTINUE
19    CONTINUE
      CALL FUNCT(VRES,YRES)
      RETURN
      END
C
C
C
      SUBROUTINE GPHASE(TIN,T,FLOW,LABEL)
C     SUBROUTINE TO ITERATE THE GAS PHASE CASE
C
C     SET UPPER AND LOWER BOUNDS TO GASIFYING AGENT FLOWRATE
      FLOWU=20.0
      FLOWL=0.0
C
      LABEL=0
      CALL STAGE2(TIN,T,FLOWL,ERRORL,LABEL)
      IF (LABEL.NE.0) THEN
      LABEL=LABEL+10
      RETURN
      END IF
      J=0
9520  LABEL=0
      CALL STAGE2(TIN,T,FLOWU,ERRORU,LABEL)
      J=J+1
      IF (J.GE.20) THEN
      LABEL=20
      RETURN
      END IF
C     CHECK THAT "STAGE2" HAS WORKED FOR UPPER BOUND, ELSE
C     ADJUST UPPER BOUND
      IF (LABEL.NE.0) THEN
```

```
      FLOWU=FLOWU/2.0
      GOTO 9520
      END IF
C     CHECK UPPER & LOWER BOUNDS ARE ON EITHER SIDE OF THE
C     SOLUTION
      IF (FLOWU*FLOWL.GT.0.0) THEN
      FLOWU=FLOWU*1.9
      GOTO 9520
      END IF
      J=0
C
C     MAIN ITERATIVE LOOP
      DO
      J=J+1
C     CHECK NUMBER OF ITERATIONS
      IF (J.GE.50) THEN
      LABEL=30
      RETURN
      END IF
C     SPLIT THE RANGE OF ITERATION
      FLOW=(FLOWU+FLOWL)/2.0
      LABEL=0
      CALL STAGE2(TIN,T,FLOW,ERROR,LABEL)
      IF (LABEL.NE.0) THEN
      LABEL=40+LABEL
      RETURN
      END IF
C     SET NEW,SMALLER RANGE OF ITERATION
      IF (ERRORU*ERROR.GT.0.0) THEN
      ERRORU=ERROR
      FLOWU=FLOW
      ELSE
      ERRORL=ERROR
      FLOWL=FLOW
      END IF
      ERROR=ABS(ERROR)
C     TEST FOR CONVERGENCE
      UNTIL (ERROR.LT.5.0)
      RETURN
      END
```

Example Printout:

```
******************
RESULTS FROM MODEL
RUN NUMBER  EG M 4
******************


INPUT CONDITIONS

PRESSURE IS   1.0 atm

HEAT LOSS IS  2.5 %

COMPOSITION OF DRY BIOMASS
C 1.000 H 1.500 O 0.700

MOISTURE CONTENT OF BIOMASS     10.00% dry ash free weight
ASH CONTENT OF BIOMASS           1.00% dry ash free weight

HEAT OF COMBUSTION IS    19.03 MJ/kg


FIRST STAGE CONDITIONS

GASIFYING AGENT TEMPERATURE   298.00 K
GASIFYING STEAM ENTHALPY     2780.00 kJ/kg

GASIFYING AGENT COMPOSITION
 H2      CO      CO2     CH4     N2      H2O     O2
0.0000  0.0000  0.0000  0.0000  0.0000  0.1000  0.9000


CARBON BOUNDARY CALCULATION

OUTLET TEMPERATURE            962.44 K

GASIFYING AGENT FLOWRATE   0.3216 mol/mol Biomass
OUTLET GAS FLOWRATE        1.8477 mol/mol Biomass
OUTLET FLOW (DRY BASIS)    1.6673 mol/mol Biomass

OUTLET GAS COMPOSITION
 H2      CO      CO2     CH4     N2      H2O
0.3612  0.3575  0.1643  0.0194  0.0000  0.0976

OUTLET GAS COMPOSITION, DRY BASIS
 H2      CO      CO2     CH4     N2
0.4002  0.3962  0.1821  0.0215  0.0000

OUTLET GAS FLOWRATES, mol/mol Biomass
 H2      CO      CO2     CH4     N2      H2O
0.6673  0.6605  0.3036  0.0358  0.0000  0.1804

OUTLET GAS CALORIFIC VALUE (IGU standard) 10.3946 MJ/m**3

CHEMICAL POTENTIAL ENERGY CONVERSION EFFICIENCY  80.2%
```

continued on next page...

```
RESULTS FROM MODEL
RUN NUMBER  EG M 4 contd.


FIXED TEMPERATURE CALCULATION

OUTLET TEMPERATURE            1073.00 K

GASIFYING AGENT FLOWRATE    0.3600 mol/mol Biomass
OUTLET GAS FLOWRATE         1.9202 mol/mol Biomass
OUTLET FLOW (DRY BASIS)     1.6776 mol/mol Biomass

OUTLET GAS COMPOSITION
 H2       CO       CO2      CH4      N2       H2O
0.3529  0.3741  0.1459  0.0008  0.0000  0.1263

OUTLET GAS COMPOSITION, DRY BASIS
 H2       CO       CO2      CH4      N2
0.4039  0.4282  0.1670  0.0009  0.0000

OUTLET GAS FLOWRATES, mol/mol Biomass
 H2       CO       CO2      CH4      N2       H2O
0.6776  0.7183  0.2802  0.0015  0.0000  0.2425

OUTLET GAS CALORIFIC VALUE (IGU standard) 10.0467 MJ/m**3

CHEMICAL POTENTIAL ENERGY CONVERSION EFFICIENCY  78.4%


GAS PHASE RESULTS

GASIFYING AGENT COMPOSITION
 H2       CO       CO2      CH4      N2       H2O      O2
0.0000  0.0000  0.0000  0.0000  0.0000  0.0000  1.0000

GASIFYING AGENT TEMPERATURE    298.00 K
GASIFYING STEAM ENTHALPY      2780.00 kJ/kg

OUTLET TEMPERATURE            1273.00 K

GASIFYING AGENT FLOWRATE    0.0332 mol/mol Biomass
OUTLET GAS FLOWRATE         1.9232 mol/mol Biomass
OUTLET FLOW (DRY BASIS)     1.5909 mol/mol Biomass

OUTLET GAS COMPOSITION
 H2       CO       CO2      CH4      N2       H2O
0.3072  0.3873  0.1327  0.0000  0.0000  0.1728


OUTLET GAS COMPOSITION, DRY BASIS
 H2       CO       CO2      CH4      N2
0.4039  0.4282  0.1670  0.0009  0.0000

OUTLET GAS FLOWRATES, mol/mol Biomass
 H2       CO       CO2      CH4      N2       H2O
0.5909  0.7448  0.2552  0.0000  0.0000  0.3323

OUTLET GAS CALORIFIC VALUE (IGU standard) 10.0981 MJ/m**3

CHEMICAL POTENTIAL ENERGY CONVERSION EFFICIENCY  75.3%
```

Program FLMOD

The basic flow diagram of this program is shown as Figure B2. A complete list of subroutines and functions is given below:

Subroutines

DUMP	Subroutine to dump the values of all arrays to an output file, to help with debugging.
HTBAL	Main heat balance subroutine. Calls POX to perform the gasification mass balances.
INPUT	Subroutine to perform the interactive data entry.
INIT	Subroutine to initialise all of the arrays used in the program.
ITER	Subroutine to perform iteration. Used in the iteration of the heat balance.
NORM	A general purpose subroutine. Takes an array, and normalises it so that the sum of the elements equals 1.
OPINP	Subroutine to open the results files.
POX	Subroutine to calculate the mass balances of the gasification step. Calls WGSEQ to calculate the equilibrium gas composition.
PYROL	Subroutine which reads pyrolysis data from the pyrolysis data file and loads the pyrolysis products into the product gas array.
RESPRI	Subroutine to print the results.
TERM	Subroutine to terminate the running of the program.
WGSEQ	Subroutine to perform the calculation of gas composition, using equilibrium of the homogeneous water-gas shift reaction.

Functions

HHVCHO	Function to calculate the HHV of a biomass, using the IGT equation.
SHEAT	Function which returns the sensible heat of a given component at a given temperature. Used in heat balance calculations.
WGSKEQ	Function to calculate the equilibrium constant of the homogeneous water gas shift reaction at a given temperature.

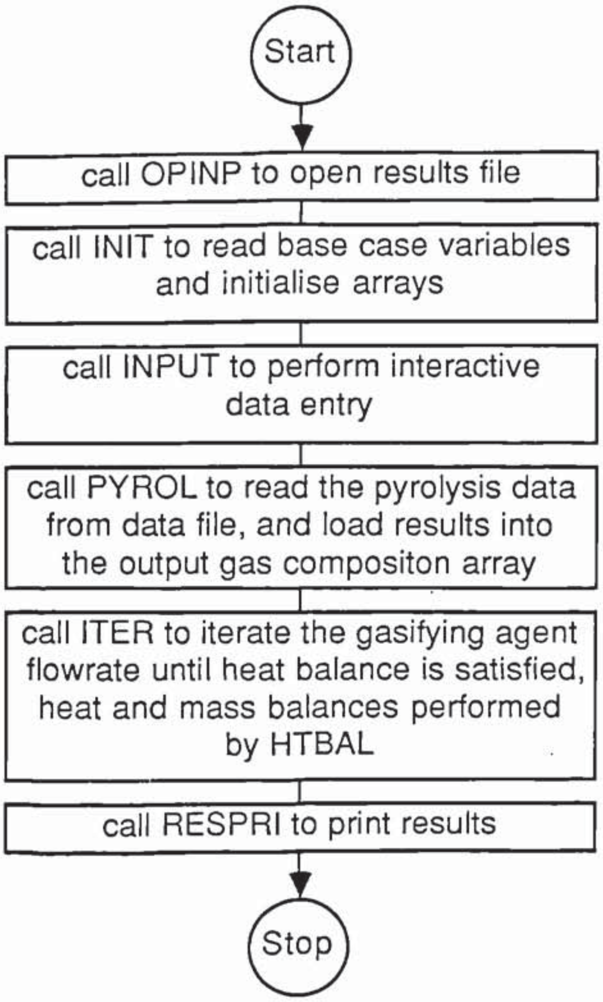


Figure B2 Flow Diagram of Program FLMOD

```
      PROGRAM FLMOD
C
C
C ***************************************************************:
C
C       FLUIDISED BED GASIFIER MODEL
C        BASED ON MODIFIED EQUILIBRIUM METHOD
C        TAKING PYROLYSIS, AND TAR CRACKING INTO ACCOUNT
C
C       WRITTEN BY J M DOUBLE
C                 ASTON UNIVERSITY
C
C ****************************************************************
C
C
C
C   NO ASSIGNMENTS ARE NEEDED WHEN RUNNING THE PROGRAM -
C    ALL FILES ARE OPENED FROM WITHIN THE PROGRAM
C
C
      REAL GAFLOW
      CHARACTER*12 RUN*12,GASIFIER*12,BIOFIL*8
      EXTERNAL HTBAL
      DO 100 I=1,13
        WRITE(3,*) ' '
100   CONTINUE
      WRITE(3,*) '            PROGRAM FLMOD'
      WRITE(3,*) ' '
      WRITE(3,*)'Fluidised bed gasifier model by J.M.Double'
      DO 101 I=1,13
        WRITE(3,*) ' '
101   CONTINUE
      CALL OPINP
      CALL INIT(RUN,GASIFIER,BIOFIL)
1000  CALL INPUT('FLMOD ',RUN,GASIFIER,BIOFIL)
      CALL DUMP('BEFORE PYROL                 ',25)
      CALL PYROL(BIOFIL)
      CALL DUMP('AFTER PYROL                  ',25)
      CALL ITER(GAFLOW,0.01,0.1,HTBAL)
      CALL DUMP('AFTER ITER                   ',25)
      CALL RESPRI(GAFLOW)
      GOTO 1000
      END
C
C
C
C
      SUBROUTINE NORM(X,N)
      REAL X(N),TOTAL
C
C     SUBROUTINE TO NORMALISE COMPOSITIONS
C
      TOTAL=0.0
      DO 1000 I=1,N,1
      TOTAL=TOTAL+X(I)
```

```
1000  CONTINUE
      DO 1001 I=1,N,1
      X(I)=X(I)/TOTAL
1001  CONTINUE
      RETURN
      END
C
C
C
C
C
C
      FUNCTION HHVCHO(CMOL,HMOL,OMOL,NMOL,ASH)
      REAL CMOL,HMOL,OMOL,NMOL,ASH,C,H,O,N,A
      REAL TOTMAS,HHV
C
C     FUNCTION TO CALCULATE HIGHER HEATING VALUE OF BIOMASS
C     USING IGT EQUATION
C
C     CMOL,HMOL,OMOL,NMOL ARE MOLE COMPOSITION OF C,H,O,N
C     ASH IS MASS OF ASH PER KG DRY ASH FREE BIOMASS
C
C
C CALCULATE MASS FRACTIONS
      C=CMOL*12.0
      H=HMOL
      O=OMOL*16.0
      N=NMOL*14.0
      A=ASH
      TOTMAS=C+H+O+N+A
C CALCULATE DRY WEIGHT COMPOSITION IN PERCENT
      C=100.0*C/TOTMAS
      H=100.0*H/TOTMAS
      O=100.0*O/TOTMAS
      N=100.0*N/TOTMAS
      A=100.0*A/TOTMAS
C CALCULATE HEAT OF COMBUSTION OF BIOMASS, USING IGT EQUATION.
      HHV=146.58*C+568.78*H+29.45-6.58*A-51.53*(O+N)
C CONVERT FROM BTU/LB TO MJ/KG
      HHV=HHV*2.326E-3
C CONVERT TO ASH FREE BASIS, IE HHV FOR 1 KG OF DRY BIOMASS +
C   ASSOCIATED ASH
      HHV=HHV/(1.0-A/100.0)
C LOAD RESULT AS RETURN VALUE OF FUNCTION
      HHVCHO=HHV
      RETURN
      END
C
C
C
C
```

```
      SUBROUTINE OPINP
C
******************************************************************
*
C        SUBROUTINE TO OPEN OUTPUT FILES
C
******************************************************************
*
C
      OPEN(UNIT=20,FILE='        B1',STATUS='UNKNOWN')
      OPEN(UNIT=25,FILE='        B9',STATUS='UNKNOWN')
      RETURN
      END
C
C
C
C
C
C
C
      FUNCTION SHEAT(T,I)
C
C FUNCTION TO CALCULATE GAS PHASE SENSIBLE HEAT FOR EACH COMPONENT
C AT TEMPERATURE T
C
      REAL T
      REAL A(22),B(22),C(22),D(22),E(22),F(22)
      DATA (A(J),J=1,22) /6.88,6.92,5.14,5.04,7.07,8.10,2.46,3*0.,
     *7.20,4*0.,6.22,2.673,0.5,0.22275,0.45,0.,0.194088/
      DATA (B(J),J=1,22) /-0.022,-0.65,15.4,9.32,-1.32,-0.72,36.1,
     *3*0.,3.60,4*0.,2.71,2.617,0.,0.21803,2*0.,0.04108/
      DATA (C(J),J=1,22) /0.21,2.8,-9.94,8.87,3.31,3.63,-7.0,8*0.,
     *-0.37,6*0./
      DATA (D(J),J=1,22) /0.13,-1.14,2.42,-5.37,-1.26,1.16,-0.46,
     *8*0.,-0.22,6*0./
      DATA (E(J),J=1,22) /16*0.,116900.,0.,9741.7,2*0.,1933.37/
      DATA (F(J),J=1,22) /8582.2,8609.98,8906.13,8339.83,8691.22,
     *10100.83,9518.41,3*0.,9646.57,4*0.,8245.36,5459.00,623.43,
     *454.92,561.09,0.,276.76/
      SHEAT=A(I)*T+1.0E-3*B(I)*T*T/2.+1.0E-6*C(I)*T*T*T/3.+
     *1.0E-10*D(I)*T*T*T*T/4.+ E(I)/T
      SHEAT=SHEAT*4.182-F(I)
      RETURN
      END
C
C
C
C
      SUBROUTINE ITER(X,SCALE,TOL,DUMMY)
      REAL ERROR,SCALE,TOL,X
      EXTERNAL DUMMY
      REAL XLOW,XHIGH,ERRORL,ERRORH
      TOL=ABS(TOL)
C INITIALISE LOWER BOUND
      DO 100 XLOW=0.0,SCALE*10.,SCALE/100.
      CALL DUMMY(XLOW,ERRORL,LAB)
```

```
      IF (ABS(ERRORL).LT.TOL) THEN
         X=XLOW
         RETURN
      END IF
      IF (LAB.EQ.0) GOTO 200
 100  CONTINUE
      CALL TERM('CANNOT GET A LOWER BOUND IN ITER')
C INITIALISE UPPER BOUND
 200  DO 300 XHIGH=XLOW,XLOW+SCALE*10.,SCALE/5.
      CALL DUMMY(XHIGH,ERRORH,LAB)
      IF (ABS(ERRORH).LT.TOL) THEN
         X=XHIGH
         RETURN
      END IF
      IF (LAB.NE.0) CALL TERM('ERROR WITH UPPER BOUND IN ITER')
      IF (ERRORL*ERRORH.LT.0) GOTO 400
 300  CONTINUE
      CALL TERM('CANNOT GET UPPER BOUND IN ITER')
C MAIN ITERATION LOOP
 400  DO 500 I=1,100
       X=(XHIGH+XLOW)/2.0
       CALL DUMMY(X,ERROR,LAB)
       IF (ABS(ERROR).LT.TOL) RETURN
       IF (ERROR*ERRORL.LT.0.) THEN
         ERRORH=ERROR
         XHIGH=X
       ELSE
         ERRORL=ERROR
         XLOW=X
       END IF
 500  CONTINUE
      CALL TERM('TOO MANY ITERATIONS IN ITER   ')
      END
C
C
C
C
      SUBROUTINE DUMP(POSITION,LFN)
      CHARACTER*30 POSITION
      INTEGER LFN
C ****************************************************************

C      SUBROUTINE TO PRINT OUT CONTENTS OF COMMON BLOCKS
C       INTO A FILE ASSIGNED TO CHANNEL 'LFN'
C ****************************************************************
C
C COMMON BLOCK OPCOND CONTAINS OPERATING CONDITIONS OF THE
GASIFIER
      REAL P,T,HLOSS,GASIN(22),GAST,ENTHS,BIOT
      COMMON /OPCOND/ P,T,HLOSS,GASIN,GAST,ENTHS,BIOT
C COMMON BLOCK FLOWS CONTAINS MOLAR FLOWRATES OF EACH COMPONENT IN
C  A NUMBER OF MATERIAL FLOWS IN THE GASIFIER
      REAL BIOFLO(22),POXIN(22),OUT1(22),OUT2(22),ASHFLO(22)
      COMMON /FLOWS/ BIOFLO,POXIN,OUT1,OUT2,ASHFLO
C COMMON BLOCK PROPS CONTAINS DATA ON EACH COMPONENT
```

```
      REAL DELHC(22),COMPC(22),COMPH(22),COMPO(22),COMPMW(22)
      REAL COMPN(22)
      CHARACTER COMPONENT(22)*5
      COMMON /PROPS/ DELHC,COMPC,COMPH,COMPO,COMPN,
     *COMPMW,COMPONENT
C
C TITLE THE TABLE OF VALUES
      WRITE(LFN,99999) POSITION
C
C PRINT OUT THE VALUE OF GASIFIER TEMPERATURE
      WRITE(LFN,90000) T
C
C PRINT OUT OPERATING CONDITIONS
      WRITE(LFN,99998) P,HLOSS,GAST,ENTHS,BIOT
C
C PRINT THE HEADINGS FOR THE TABLE
      WRITE(LFN,99997)
C
C PRINT THE TABLE OF VALUES
      WRITE(LFN,99996) (COMPONENT(J),COMPMW(J),DELHC(J),COMPC(J),
     *COMPH(J),COMPO(J),COMPN(J),GASIN(J),BIOFLO(J),
     *POXIN(J),OUT1(J),OUT2(J),ASHFLO(J),J=1,22)
C
      RETURN
99999 FORMAT('1  DUMP OF VALUES OF VARIABLES AT POSITION: ',A30)
99998 FORMAT('0  P = ',F6.2,'  HEAT LOSS = ',F6.2,'  G.A.Temp = ',
     *F7.2,'  H for steam = ',F7.1,'  Biomass Temp = ',F7.2)
99997 FORMAT('0',8X,'MW',5X,'HHV',5X,'C',6X,'H',6X,'O',6X,'N',
     *4X,'GASIN  BIOFLO  POXIN   OUT1    OUT2    ASHFLO',/,
     *' ',101('='))
99996 FORMAT(22('0',A5,2X,F5.2,1X,F6.1,5(1X,F6.4),5(1X,F7.4),/))
90000 FORMAT('0 GASIFIER TEMPERATURE = ',F7.2)
      END
C
C
C
C
C
C
C
C
      SUBROUTINE RESPRI(GAFLOW)
      REAL GAFLOW
C COMMON BLOCK OPCOND CONTAINS OPERATING CONDITIONS OF THE
C  GASIFIER
      REAL P,T,HLOSS,GASIN(22),GAST,ENTHS,BIOT
      COMMON /OPCOND/ P,T,HLOSS,GASIN,GAST,ENTHS,BIOT
C COMMON BLOCK FLOWS CONTAINS MOLAR FLOWRATES OF EACH COMPONENT IN
C  A NUMBER OF MATERIAL FLOWS IN THE GASIFIER
      REAL BIOFLO(22),POXIN(22),OUT1(22),OUT2(22),ASHFLO(22)
      COMMON /FLOWS/ BIOFLO,POXIN,OUT1,OUT2,ASHFLO
C COMMON BLOCK PROPS CONTAINS DATA ON EACH COMPONENT
      REAL DELHC(22),COMPC(22),COMPH(22),COMPO(22),COMPMW(22)
      REAL COMPN(22)
      CHARACTER COMPONENT(22)*5
```

```
      COMMON /PROPS/ DELHC,COMPC,COMPH,COMPO,COMPN,
     *COMPMW,COMPONENT
C
      REAL OUTFLO,OUTMF(22),OUTDTF(22),HHVGAS
      INTEGER IGC(8)
C
C COMPONENTS IN GAS STREAM
      DATA (IGC(J),J=1,8) /1,2,3,4,5,6,7,20/
C
C
C CALCULATION OF DRY TAR FREE GAS ETC.
      DO 100 I=1,22
        OUTDTF(I)=OUT2(I)
        OUTMF(I)=OUT2(I)
100   CONTINUE
      OUTDTF(6)=0.
      OUTDTF(20)=0.
      OUTFLO=0.
      DO 150 I=1,22
      OUTFLO=OUTFLO+OUTDTF(I)
150   CONTINUE
      CALL NORM(OUTDTF,22)
      CALL NORM(OUTMF,22)
      HHVGAS=0.0
      DO 200 I=1,22
        HHVGAS=HHVGAS+OUTDTF(I)*DELHC(I)
200   CONTINUE
      HHVGAS=HHVGAS/22.4
C
C
C PRINT HEADING
      WRITE(20,90000)
      WRITE(20,90001)
C PRINT GAS FLOWS
      WRITE(20,90004) GAFLOW,GAFLOW*22.4,OUTFLO,OUTFLO*22.4
C PRINT GAS HEATING VALUE
      WRITE(20,90005) HHVGAS
C PRINT GAS COMPOSITION ETC.
      WRITE(20,90006) (COMPONENT(IGC(J)),J=1,8),
     *(OUT2(IGC(J)),J=1,8),(OUTMF(IGC(J)),J=1,8),
     *(OUTDTF(IGC(J)),J=1,8)
      WRITE(20,90010) ASHFLO(22),ASHFLO(17)*COMPMW(17)
      WRITE(20,90000)
      WRITE(3,*) 'RESULTS HAVE BEEN PRINTED'
      RETURN
90000 FORMAT(5X,67('*'))
90001 FORMAT(/,5X,'RESULTS',/,'+',4X,7('_'),/)
90004 FORMAT(5X,'GAS FLOWS',/,5X,'Gasifying agent:',4X,F9.4,
     *' mol/kg biomass',/,25X,F9.4,' nm3/kg biomass',/,5X,
     *'Outlet gas:',9X,F9.4,' mol/kg biomass',/,5X,
     *'(dry tar free)',6X,F9.4,' nm3/kg biomass',/)
90005 FORMAT(5X,'Outlet gas HHV:      ',F7.2,'    MJ/nm3',/)
90006 FORMAT(5X,'OUTLET GAS DATA',/,7X,'-composition by ',
     *'mole fraction',/,7X,'-component flows in mol/kg biomass',
     */,/,19X,8(A5,1X),/,/,5X,'gas flows      ',8(F5.3,1X),/,
     *5X,'Compositions',/,5X,'raw gas',7X,8(F5.3,1X),/,5X,
```

```
     *'dry tar free  ',8(F5.3,1X))
90010 FORMAT(/,5X,'ASH FLOW - kg Ash/kg Biomass',/,5X,'Ash : ',
     *F6.4,10X,'Carbon contained in ash: ',F6.4,/)
      END
C
C
C
C
C
      SUBROUTINE HTBAL(GAFLOW,ERROR,LABEL)
C ************************************************************
C        SUBROUTINE TO PERFORM HEAT BALANCE
C ************************************************************
C
C
C
C BASIS FOR THE HEAT BALANCE IS:
C      BASE TEMPERATURE = 25 DEG C
C      BASE STATE OF EACH COMPOUND IS ITS USUAL STATE AT 25 DEG C
C      (E.G. WATER - LIQUID, TAR - LIQUID, H2 - GAS ETC.)
C
C COMMON BLOCK OPCOND CONTAINS OPERATING CONDITIONS OF THE
C  GASIFIER
      REAL P,T,HLOSS,GASIN(22),GAST,ENTHS,BIOT
      COMMON /OPCOND/ P,T,HLOSS,GASIN,GAST,ENTHS,BIOT
C COMMON BLOCK FLOWS CONTAINS MOLAR FLOWRATES OF EACH COMPONENT IN
C  A NUMBER OF MATERIAL FLOWS IN THE GASIFIER
      REAL BIOFLO(22),POXIN(22),OUT1(22),OUT2(22),ASHFLO(22)
      COMMON /FLOWS/ BIOFLO,POXIN,OUT1,OUT2,ASHFLO
C COMMON BLOCK PROPS CONTAINS DATA ON EACH COMPONENT
      REAL DELHC(22),COMPC(22),COMPH(22),COMPO(22),COMPMW(22)
      REAL COMPN(22)
      CHARACTER COMPONENT(22)*5
      COMMON /PROPS/ DELHC,COMPC,COMPH,COMPO,COMPN,
     *COMPMW,COMPONENT
C LATENT HEATS OF VAPOURISATION
      REAL HVAP(22)
      DATA (HVAP(I),I=1,22) /5*0.,43952.4,13*0.,4182.,2*0./
C
C PERFORM EQUILIBRIUM CALCULATION ON PYROLYSIS PRODUCTS
C
      CALL POX(GAFLOW,LABEL)
      IF (LABEL.NE.0) THEN
        ERROR=1.E32
        WRITE(3,*) 'SUBROUTINE HTBAL, LABEL = ',LABEL
        RETURN
      END IF
C
C INITIALISE HEAT FLOWS TO ZERO
      CEIN=0.
      CEOUT=0.
      SHIN=0.
      SHOUT=0.
C
C EVALUATE CHEMICAL HEAT OF GASIFYING AGENT
```

```
      DO 100 I=1,22
        CEIN=CEIN+GAFLOW*GASIN(I)*DELHC(I)
100   CONTINUE
C
C ADD CHEMICAL HEAT OF BIOMASS TO THIS
      DO 200 I=1,22
        CEIN=CEIN+BIOFLO(I)*DELHC(I)
200   CONTINUE
C
C EVALUATE CHEMICAL HEAT OF OUTLET GAS
      DO 300 I=1,22
        CEOUT=CEOUT+OUT2(I)*DELHC(I)
300   CONTINUE
C
C ADD CHEMICAL HEAT OF ASH TO THIS
      DO 350 I=1,22
        CEOUT=CEOUT+ASHFLO(I)*DELHC(I)
350   CONTINUE
C
C CONVERT CHEMICAL HEATS FROM MJ TO kJ
      CEIN=CEIN*1000.
      CEOUT=CEOUT*1000.
      HEATLOSS=10.*HLOSS*DELHC(18)
C
C CALCULATE SENSIBLE HEAT OF GASIFYING AGENT
      DO 400 I=1,22
        IF (I.EQ.6) GOTO 399
        SHIN=SHIN+GAFLOW*GASIN(I)*SHEAT(GAST,I)
399     CONTINUE
400   CONTINUE
C
C ADD STEAM ENTHALPY TO THIS
      SHIN=SHIN+GAFLOW*GASIN(6)*(ENTHS-104.8)*COMPMW(6)
C
C ADD BIOMASS ENTHALPY TO THIS
      DO 450 I=1,22
        SHIN=SHIN+BIOFLO(I)*SHEAT(BIOT,I)
450   CONTINUE
C
C CALCULATE SENSIBLE HEAT OF OUTLET GAS
      DO 500 I=1,22
        SHOUT=SHOUT+OUT2(I)*(SHEAT(T,I)+HVAP(I))
500   CONTINUE
C
C ADD SENSIBLE HEAT OF ASH
      DO 600 I=1,22
        SHOUT=SHOUT+ASHFLO(I)*SHEAT(T,I)
600   CONTINUE
C
C
C PERFORM HEAT BALANCE
      ERROR=CEIN+SHIN-HEATLOSS-CEOUT-SHOUT
      WRITE(3,*) 'SUBROUTINE HTBAL, GAFLOW = ',GAFLOW,
     * '    ERROR = ',ERROR
      RETURN
```

```
      END
C
C
C
C
      SUBROUTINE INPUT(PROG,RUN,GASIFIER,BIOFIL)
      CHARACTER*6 PROG
      CHARACTER BIOFIL*8,GASIFIER*12,RUN*12
C
******************************************************************
C       SUBROUTINE TO READ INPUT DATA
C
******************************************************************
C
C COMMON BLOCK OPCOND CONTAINS OPERATING CONDITIONS OF THE
C  GASIFIER
      REAL P,HLOSS,GASIN(22),GAST,ENTHS,BIOT
      COMMON /OPCOND/ P,T,HLOSS,GASIN,GAST,ENTHS,BIOT
C COMMON BLOCK FLOWS CONTAINS MOLAR FLOWRATES OF EACH COMPONENT IN
C  A NUMBER OF MATERIAL FLOWS IN THE GASIFIER
      REAL BIOFLO(22),POXIN(22),OUT1(22),OUT2(22),ASHFLO(22)
      COMMON /FLOWS/ BIOFLO,POXIN,OUT1,OUT2,ASHFLO
C COMMON BLOCK PROPS CONTAINS DATA ON EACH COMPONENT
      REAL DELHC(22),COMPC(22),COMPH(22),COMPO(22),COMPMW(22)
      REAL COMPN(22)
      CHARACTER COMPONENT(22)*5
      COMMON /PROPS/ DELHC,COMPC,COMPH,COMPO,COMPN,
     *COMPMW,COMPONENT
C COMMON BLOCK KWGS CONTAINS THE EQUILIBRIUM CONSTANT FOR
C  THE WATER GAS SHIFT REACTION
      REAL KWGS
      COMMON /EQUK/ KWGS
C
      GAFLOW=0.
      DO 1 I=1,22
        POXIN(I)=0.
        OUT1(I)=0.
        OUT2(I)=0.
        ASHFLO(I)=0.
1     CONTINUE
      WRITE(3,*) 'Enter name or number of run'
      READ(0,99999) RUN
10000 AIR=GASIN(5)/0.79
      OXYGEN=GASIN(16)-AIR*0.21
      STEAM=GASIN(6)
      WRITE(3,*) 'THE VALUES OF THE OPERATING VARIABLES ARE',
     *  ' CURRENTLY:'
      WRITE(3,90004) GASIFIER,RUN
      WRITE(3,90006) BIOFIL
      WRITE(3,90007) AIR,OXYGEN,STEAM,GAST-273.15,ENTHS
      WRITE(3,90008) P,T-273.15,HLOSS
      WRITE(3,90001)
      WRITE(3,*) '  0.  Stop the program'
      WRITE(3,*) '  1.  Change the gasifier name'
      WRITE(3,*) '  2.  Change the biomass'
      WRITE(3,*) '  3.  Change the gasifying agent composition'
```

```
      WRITE(3,*) '  4.  Change the gasifying agent temperature &',
     *'steam enthalpy'
      WRITE(3,*) '  5.  Change the gasifier pressure'
      WRITE(3,*) '  6.  Change the gasifier temperature'
      WRITE(3,*) '  7.  Change the gasifier heat loss'
      WRITE(3,*) '  8.  Perform the calculation for this data set'
      WRITE(3,*) 'ENTER THE NUMBER OF THE ACTION REQUIRED'
      READ(0,*) ICHANG
      GOTO (10001,10002,10003,10004,10005,10006,10007,10008)
     * ICHANG
      IF (ICHANG.EQ.0) CALL TERM('NORMAL END OF PROGRAM         ')
      CALL TERM('INCORRECT SELECTION OF ITEM   ')
10001 WRITE(3,*) 'Enter name of gasifier'
      READ(0,99999) GASIFIER
      GOTO 10000
10002 WRITE(3,*) 'Enter Biomass type'
      READ(0,99998) BIOFIL
      GOTO 10000
C
C
C GASIFYING AGENT
10003 WRITE(3,*) 'Enter type of gasifying agent'
      WRITE(3,*) ' '
      WRITE(3,*) '   1. Air'
      WRITE(3,*) '   2. Oxygen'
      WRITE(3,*) '   3. Steam'
      WRITE(3,*) '   4. Mixture of Air, Oxygen and Steam'
      WRITE(3,*) ' '
      WRITE(3,*) 'Enter the number of the option required'
      READ(0,*) IGATYP
      AIR=0.
      OXYGEN=0.
      STEAM=0.
      IF (IGATYP.EQ.1) THEN
        AIR=1.0
      ELSE IF (IGATYP.EQ.2)
        OXYGEN=1.0
      ELSE IF (IGATYP.EQ.3)
        STEAM=1.0
      ELSE IF (IGATYP.EQ.4)
        WRITE(3,*) 'Enter relative quantities of air, ',
     *   'oxygen and steam'
        WRITE(3,*) 'Enter fraction of air'
        READ(0,*) AIR
        WRITE(3,*) 'Enter fraction of oxygen'
        READ(0,*) OXYGEN
        WRITE(3,*) 'Enter fraction of steam'
        READ(0,*) STEAM
      ELSE
        CALL TERM('INVALID GASIFYING AGENT       ')
      END IF
        TOT=AIR+STEAM+OXYGEN
        AIR=AIR/TOT
        OXYGEN=OXYGEN/TOT
        STEAM=STEAM/TOT
```

```
         GASIN(5)=0.79*AIR
         GASIN(16)=0.21*AIR+OXYGEN
         GASIN(6)=STEAM
         GOTO 10000
C
C
10004 WRITE(3,*) 'Enter gasifying agent temperature'
      READ(0,*) GAST
      IF (GASIN(6).GT.0.00001) THEN
        WRITE(3,*) 'Enter gasifying steam enthalpy (kJ/kg)'
        READ(0,*) ENTHS
      ELSE
        ENTHS=0.
      END IF
      GAST=GAST+273.15
      GOTO 10000
C
10005 WRITE(3,*) 'Enter gasifier pressure (Bar)'
      READ(0,*) P
      GOTO 10000
C
10006 WRITE(3,*) 'Enter gasifier temperature, deg C'
      READ(0,*) T
      T=T+273.15
      GOTO 10000
C
10007 WRITE(3,*) 'Enter gasifier heat loss, expressed as % of'
     *,' biomass HHV'
      READ(0,*) HLOSS
      GOTO 10000
C
10008 KWGS=WGSKEQ(T)
C
C
C
C ECHO DATA TO RESULTS FILE
      WRITE(20,90000)
      WRITE(20,90002)
      WRITE(20,90003) PROG
      WRITE(20,90001)
      WRITE(20,90004) GASIFIER,RUN
      WRITE(20,90002)
      WRITE(20,90005)
      WRITE(20,90006) BIOFIL
      WRITE(20,90001)
      WRITE(20,90007) AIR,OXYGEN,STEAM,GAST-273.15,ENTHS
      WRITE(20,90001)
      WRITE(20,90008) P,T-273.15,HLOSS
      WRITE(20,90001)
      RETURN
90000 FORMAT('1',/)
90001 FORMAT(1X)
90002 FORMAT(5X,67('*'))
90003 FORMAT(5X,'PROGRAM ',A6,5X,'GASIFIER MODEL RESULTS',
     *6X,'J.M.DOUBLE',5X,'ASTON')
```

```
90004 FORMAT(5X,'GASIFIER: ',A12,28X,'RUN: ',A12)
90005 FORMAT(/,5X,'INPUT DATA',/,'+',4X,10('_'),/)
90006 FORMAT(5X,'BIOMASS DATA FROM FILE ',A8)
90007 FORMAT(5X,'GASIFYING AGENT DATA - Mole fractions',
     */,6X,'Air',7X,'O2',7X,'Steam',5X,'T degC',
     *4X,'Enths',/,5X,3(F5.3,5X),2(F6.1,4X))
90008 FORMAT(5X,'OPERATING PARAMETERS',/,5X,'Pressure = ',F4.1,
     *' Bar',5X,'Temperature = ',F6.1,5X,'Heat Loss = ',F4.1,' %')
99998 FORMAT(A8)
99999 FORMAT(A12)
      END
C
C
C
C
      SUBROUTINE INIT(RUN,GASIFIER,BIOFIL)
C
******************************************************************
C       SUBROUTINE TO INITIALISE OPERATING VARIABLES
C
******************************************************************
      CHARACTER RUN*12,GASIFIER*12,BIOFIL*8
C COMMON BLOCK OPCOND CONTAINS OPERATING CONDITIONS OF THE
C  GASIFIER
      REAL P,T,HLOSS,GASIN(22),GAST,ENTHS,BIOT
      COMMON /OPCOND/ P,T,HLOSS,GASIN,GAST,ENTHS,BIOT
C COMMON BLOCK FLOWS CONTAINS MOLAR FLOWRATES OF EACH COMPONENT IN
C  A NUMBER OF MATERIAL FLOWS IN THE GASIFIER
      REAL BIOFLO(22),POXIN(22),OUT1(22),OUT2(22),ASHFLO(22)
      COMMON /FLOWS/ BIOFLO,POXIN,OUT1,OUT2,ASHFLO
C COMMON BLOCK PROPS CONTAINS DATA ON EACH COMPONENT
      REAL DELHC(22),COMPC(22),COMPH(22),COMPO(22),COMPMW(22)
      REAL COMPN(22)
      CHARACTER COMPONENT(22)*5
      COMMON /PROPS/ DELHC,COMPC,COMPH,COMPO,COMPN,
     *COMPMW,COMPONENT
      DATA (COMPONENT(I),I=1,22) /' H2  ',' CO  ',' CO2 ',' CH4 ',
     *' N2  ',' H2O ',' C2+ ',8*'     ',' O2  ',
     *'  C  ','Biom.','Char ',' Tar ','Resi.',' Ash '/
      DATA (COMPMW(I),I=1,22) /2.,28.,44.,16.,28.,18.,30.,
     *8*10000.,32.,12.,5*1.0/
      DATA (DELHC(I),I=1,22) /285.8,283.0,0.,890.4,0.,0.,1411.,
     *9*0.,393.5,5*0./
      DATA (COMPC(I),I=1,22) /0.,3*1.,0.,0.,2.,9*0.,1.,5*0./
      DATA (COMPH(I),I=1,22) /2.,0.,0.,4.,0.,2.,4.,15*0./
      DATA (COMPO(I),I=1,22) /0.,1.,2.,0.,0.,1.,9*0.,2.,6*0./
      DATA (COMPN(I),I=1,22) /4*0.,2.,17*0./
      RUN=    '            '
      GASIFIER = 'FLUID BED   '
      BIOFIL='WOODBASE'
      DO 100 I=1,22
        BIOFLO(I)=0.
        GASIN(I)=0.
100   CONTINUE
C
C
```

```
C GASIFYING AGENT - OXYGEN
      GASIN(16)=1.0
C
C INITIAL INLET TEMPERATURES ARE 25 DEG C
      BIOT=298.15
      GAST=298.15
C
C GASIFYING STEAM ENTHALPY IS THAT OF SATURATED STEAM AT 10 BAR
      ENTHS=2778.
C
C HEAT LOSS IS 5% OF BIOMASS CALORIFIC VALUE (DRY BASIS)
      HLOSS=5.00
C
C PRESSURE IS 1 BAR
      P=1.0
C
C TEMPERATURE IS 780 DEG C
      T=1053.15
      RETURN
      END
C
C
C
C
C
      SUBROUTINE TERM(MESSAGE)
      CHARACTER*30 MESSAGE
      WRITE(20,90001)
      CALL DUMP(MESSAGE,25)
      CLOSE(20)
      CLOSE(25)
      WRITE(3,90000) MESSAGE
      STOP
      RETURN
90000 FORMAT(' PROGRAM STOP BECAUSE ',A30)
90001 FORMAT('1')
      END
C
C
C
C
      SUBROUTINE WGSEQ(C,H,O,N,X,LABEL)
      REAL C,H,O,N,X(22)
      INTEGER LABEL
      REAL KWGS
      COMMON /EQUK/ KWGS
C ****************************************************************
C
C     SUBROUTINE WGSEQ
C
C ****************************************************************
C
C     SUBROUTINE TO CALCULATE A GAS COMPOSITION,
C      GIVEN:
C            NUMBER OF MOLES OF CARBON                        C
```

```
C                                HYDROGEN                  H
C                                OXYGEN                    O
C                                NITROGEN                  N
C           EQUILIBRIUM CONSTANT OF WATER-GAS SHIFT        KWGS
C      LABEL IS ERROR INDICATOR
C      X(22) IS CALCULATED GAS COMPOSITION
C      REACTION CONSIDERED IS:
C        CO + H2O = CO2 + H2
C
C
      REAL AA,BB,CC,BRCKT,XX(2,6),TOTAL(2)
      LOGICAL ROOT(2)
C
C INITIALISE GAS COMPOSITION TO ZERO
C
      DO 10 J=1,22
        X(J)=0.
10    CONTINUE
C
C CALCULATE QUADRATIC CONSTANTS OF QUADRATIC EQUATION IN
C   NUMBER OF MOLES H2
C
      AA=1.-KWGS
      BB=O-C-(H/2.)+KWGS*(H+2.*C-O)
      CC=-KWGS*(H/2.)*(2.*C-O+H/2.)
C
C SOLVE QUADRATIC EQUATION FOR BOTH ROOTS,
C  FIRST TESTING THAT THERE ARE REAL SOLUTIONS
C
      BRCKT=BB*BB-4.*AA*CC
      IF (BRCKT.LT.0.) THEN
        LABEL=1
        RETURN
      END IF
      BRCKT=SQRT(BRCKT)
      XX(1,1)=(-BB+BRCKT)/(2.*AA)
      XX(2,1)=(-BB-BRCKT)/(2.*AA)
C
C CALCULATE NUMBER OF MOLES OF OTHER SPECIES FOR EACH ROOT
C
      DO 100 I=1,2
        XX(I,3)=O-C-H/2.+XX(I,1)
        XX(I,2)=C-XX(I,3)
        XX(I,6)=H/2.-XX(I,1)
        XX(I,4)=0.
        XX(I,5)=N/2.
100   CONTINUE
C
C CONVERT NUMBER OF MOLES OF EACH SPECIES TO MOLE FRACTION
C
      DO 200 I=1,2
        TOTAL(I)=0.
        DO 198 J=1,6
          TOTAL(I)=TOTAL(I)+XX(I,J)
198     CONTINUE
```

```
        DO 199 J=1,6
          XX(I,J)=XX(I,J)/TOTAL(I)
199     CONTINUE
200   CONTINUE
C
C TEST BOTH POSSIBLE SETS OF ANSWERS FOR THE CORRECT ROOT
C
      DO 300 I=1,2
        ROOT(I)=.TRUE.
        DO 299 J=1,6

ROOT(I)=((XX(I,J).GE.0.).AND.(XX(I,J).LE.1.)).AND.ROOT(I)
299     CONTINUE
300   CONTINUE
      IF ((.NOT.ROOT(1)).AND.(.NOT.ROOT(2))) THEN
        LABEL=2
        RETURN
      END IF
      IF (ROOT(1).AND.ROOT(2)) THEN
        LABEL=3
        RETURN
      END IF
       DO 400 I=1,2
        IF (ROOT(I)) THEN
          DO 399 J=1,6
            X(J)=XX(I,J)
399       CONTINUE
        END IF
400   CONTINUE
      LABEL=0
      RETURN
      END
C
C
C
C
C
C
      FUNCTION WGSKEQ(T)
C
C ***************************************************************
C
C  FUNCTION TO CALCULATE EQUILIBRIUM CONSTANT FOR THE REACTION
C
C              CO + H2O = CO2 + H2
C
C ***************************************************************
C
C
C
C    FREE ENERGY DATA FROM BARON, PORTER & HAMMOND
C
      REAL T,B1,B2,B3,B4,B5,G
      DATA B1,B2,B3,B4,B5 /4888,-1.00473,
     *3.24645E-3,-0.95173E-6,1.24721E-10/
      G=B1/T+B2*LOG(T)+B3*T+B4*T*T+B5*T*T*T
```

```
      WGSKEQ=EXP(G)
      RETURN
      END
C
C
C
      SUBROUTINE PYROL(BIOFIL)
      CHARACTER*8 BIOFIL
      CHARACTER*16 FILEN
      REAL CMOL,HMOL,OMOL,NMOL,MOIST,ASH
      REAL TOTC,TOTH,TOTO,TOTN
      REAL BIOC,BIOH,BIOO,BION
C
C
C SUBROUTINE TO READ IN BIOMASS DATA AND PYROLYSIS
C   AND TAR CRACKING OUTPUT GASES, AND TO LOAD THE DATA
C   INTO THE RELEVANT VARIABLES
C
C
C COMMON BLOCK FLOWS CONTAINS MOLAR FLOWRATES OF EACH COMPONENT IN
C  A NUMBER OF MATERIAL FLOWS IN THE GASIFIER
      REAL BIOFLO(22),POXIN(22),OUT1(22),OUT2(22),ASHFLO(22)
      COMMON /FLOWS/ BIOFLO,POXIN,OUT1,OUT2,ASHFLO
C COMMON BLOCK PROPS CONTAINS DATA ON EACH COMPONENT
      REAL DELHC(22),COMPC(22),COMPH(22),COMPO(22),COMPMW(22)
      REAL COMPN(22)
      CHARACTER COMPONENT(22)*5
      COMMON /PROPS/ DELHC,COMPC,COMPH,COMPO,COMPN,
     *COMPMW,COMPONENT
C
C
C OPEN THE BIOMASS DATAFILE
      FILEN='        '//BIOFIL
      OPEN(UNIT=30,FILE=FILEN,STATUS='OLD')
C
C READ THE BIOMASS DATA
C   CMOL,HMOL ETC ARE THE MOLES OF C, H ETC PER KG DRY
C     ASH FREE BIOMASS
      READ(30,*) CMOL,HMOL,OMOL,NMOL
C   MOIST, ASH ARE THE KG OF MOISTURE AND ASH PER
C     KG DRY ASH FREE BIOMASS
      READ(30,*) MOIST,ASH
C
C LOAD BIOMASS DATA INTO PROPERTIES OF COMPONENT 18
      COMPC(18)=CMOL
      COMPH(18)=HMOL
      COMPO(18)=OMOL
      COMPN(18)=NMOL
C
C SET UP BIOFLO
      BIOFLO(18)=1.0
      BIOFLO(22)=ASH
      BIOFLO(6)=MOIST/18.0
C CALCULATE HEATING VALUE OF BIOMASS
      DELHC(18)=HHVCHO(CMOL,HMOL,OMOL,NMOL,ASH)
```

```
C CALCULATE TOTAL C,H,O,N IN BIOMASS, INCLUDING MOISTURE ETC.
      BIOC=0.0
      BIOH=0.0
      BIOO=0.0
      BION=0.0
      DO 10 I=1,22
        BIOC=BIOC+BIOFLO(I)*COMPC(I)
        BIOH=BIOH+BIOFLO(I)*COMPH(I)
        BIOO=BIOO+BIOFLO(I)*COMPO(I)
        BION=BION+BIOFLO(I)*COMPN(I)
10    CONTINUE
C
C READ TAR COMPOSITION
C   CMOL,HMOL ETC ARE THE MOLES OF C, H ETC PER KG TAR
      READ(30,*) CMOL,HMOL,OMOL,NMOL
C
C LOAD TAR DATA INTO PROPERTIES OF COMPONENT 20
      COMPC(20)=CMOL
      COMPH(20)=HMOL
      COMPO(20)=OMOL
      COMPN(20)=NMOL
C CALCULATE HEATING VALUE OF TAR
      DELHC(20)=HHVCHO(CMOL,HMOL,OMOL,NMOL,0.0)
C
C READ THE YIELDS OF THE VARIOUS COMPONENTS WHICH
C  PASS STRAIGHT TO THE EXIT STREAM
C  YIELD IS YIELD IN MOLES PER KG DRY ASH FREE BIOMASS
      DO 100 I=1,22
        READ(30,*) OUT1(I)
100   CONTINUE
C
C
C CLOSE THE INPUT DATA FILE
      CLOSE 30
C
C
C DO MASS BALANCE TO CALCULATE AMOUNT PASSING TO PARTIAL OXIDATION
C
C  CALCULATE TOTAL C,H,O,N PASSING STRAIGHT TO EXIT GAS
      TOTC=0.0
      TOTH=0.0
      TOTO=0.0
      TOTN=0.0
      DO 200 I=1,22
        TOTC=TOTC+OUT1(I)*COMPC(I)
        TOTH=TOTH+OUT1(I)*COMPH(I)
        TOTO=TOTO+OUT1(I)*COMPO(I)
        TOTN=TOTN+OUT1(I)*COMPN(I)
200   CONTINUE
C
C TEST MASS BALANCE
      IF ((TOTC.GT.BIOC).OR.(TOTH.GT.BIOH).OR.
     *    (TOTO.GT.BIOO).OR.(TOTN.GT.BION)) THEN
        CALL TERM('BAD MASS BALANCE IN PYROL      ')
      END IF
```

```
C
C
C LOAD REMAINDER OF MATERIAL INTO RESIDUE IN POXIN
C  INITIALISE POXIN
      DO 300 I=1,22
        POXIN(I)=0.0
300   CONTINUE
      POXIN(21)=1.0
      POXIN(22)=BIOFLO(22)
C
C  LOAD RESIDUE FLOWS OF C,H,O,N INTO PROPERTIES OF
C .  COMPONENT 21
      COMPC(21)=BIOC-TOTC
      COMPH(21)=BIOH-TOTH
      COMPO(21)=BIOO-TOTO
      COMPN(21)=BION-TOTN
      COMPMW(21)=COMPC(21)*12.0+COMPH(21)+
     *           COMPO(21)*16.0+COMPN(21)*14.0
C
C
      RETURN
      END
C
C
C
      SUBROUTINE POX(GAFLOW,LABEL)
C*****************************************************************
C  SUBROUTINE TO PERFORM PARTIAL OXIDATION MASS BALANCE,
C   USING EQUILIBRIUM TO CALCULATE GAS COMPOSITIONS
C*****************************************************************
C
      REAL GAFLOW
      INTEGER LABEL
      REAL TOTC,TOTH,TOTO,TOTN,COMPMOLES,EQH
      REAL EQMF(22)
C COMMON BLOCK OPCOND CONTAINS OPERATING CONDITIONS OF THE
C  GASIFIER
      REAL P,T,HLOSS,GASIN(22),GAST,ENTHS,BIOT
      COMMON /OPCOND/ P,T,HLOSS,GASIN,GAST,ENTHS,BIOT
C COMMON BLOCK FLOWS CONTAINS MOLAR FLOWRATES OF EACH COMPONENT IN
C  A NUMBER OF MATERIAL FLOWS IN THE GASIFIER
      REAL BIOFLO(22),POXIN(22),OUT1(22),OUT2(22),ASHFLO(22)
      COMMON /FLOWS/ BIOFLO,POXIN,OUT1,OUT2,ASHFLO
C COMMON BLOCK PROPS CONTAINS DATA ON EACH COMPONENT
      REAL DELHC(22),COMPC(22),COMPH(22),COMPO(22),COMPMW(22)
      REAL COMPN(22)
      CHARACTER COMPONENT(22)*5
      COMMON /PROPS/ DELHC,COMPC,COMPH,COMPO,COMPN,
     *COMPMW,COMPONENT
C
C INITIALISE NEW OUTLET MOLAR FLOWS WITH THOSE FROM PYROLYSIS ETC.
      DO 100 I=1,22
        OUT2(I)=OUT1(I)
        ASHFLO(I)=0.0
100   CONTINUE
C
```

```
C INITIALISE TOTAL NUMBER OF MOLES OF C,H,O,N IN CALCULATION
      TOTC=0.
      TOTH=0.
      TOTO=0.
      TOTN=0.
      DO 200 I=1,22
        COMPMOLES=GASIN(I)*GAFLOW+POXIN(I)
        TOTC=TOTC+COMPC(I)*COMPMOLES
        TOTH=TOTH+COMPH(I)*COMPMOLES
        TOTO=TOTO+COMPO(I)*COMPMOLES
        TOTN=TOTN+COMPN(I)*COMPMOLES
200   CONTINUE
C
C CALL WGSEQ TO CALCULATE EQUILIBRIUM GAS COMP
      CALL WGSEQ(TOTC,TOTH,TOTO,TOTN,EQMF,LABEL)
      IF (LABEL.NE.0) RETURN
C
C HYDROGEN BALANCE TO CALCULATE OUTLET GAS FLOW
      EQH=2.0*(EQMF(1)+EQMF(6)+2.0*EQMF(4))
      GASOUT=TOTH/EQH
C
C LOAD RESULTING GAS COMPOSITION INTO OUT2
      DO 400 I=1,6
        OUT2(I)=OUT2(I)+GASOUT*EQMF(I)
400   CONTINUE
C
C LOAD ASH INTO ASHFLO
      ASHFLO(22)=POXIN(22)
      RETURN
      END
```

Example Printout:

```
*******************************************************************
PROGRAM FLMOD      GASIFIER MODEL RESULTS       J.M.DOUBLE      ASTON

GASIFIER: FLUID BED                                RUN: EXAMPLE RUN
*******************************************************************

INPUT DATA

BIOMASS DATA FROM FILE WOOD3

GASIFYING AGENT DATA - Mole fractions
 Air       O2        Steam      T degC    Enths
0.000     1.000     0.000        25.0    2778.0

OPERATING PARAMETERS
Pressure =  1.0 Bar    Temperature =  780.0     Heat Loss =  5.0 %

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

RESULTS

GAS FLOWS
Gasifying agent:        0.0125 mol/kg biomass
                        0.2802 nm3/kg biomass
Outlet gas:             0.0610 mol/kg biomass
(dry tar free)          1.3668 nm3/kg biomass

Outlet gas HHV:        11.83   MJ/nm3

OUTLET GAS DATA
  -composition by mole fraction
  -component flows in mol/kg biomass

              H2     CO     CO2    CH4    N2     H2O    C2+    Tar

gas flows     0.020 0.021 0.015 0.004 0.000 0.012 0.001 0.000
Compositions
raw gas       0.275 0.292 0.206 0.050 0.000 0.166 0.011 0.000
dry tar free  0.330 0.350 0.247 0.060 0.000 0.000 0.013 0.000

ASH FLOW - kg Ash/kg Biomass
Ash : 0.0080  ·       Carbon contained in ash: 0.0000

*******************************************************************
```

Program EVAL1

The basic flow diagram of this program is shown as Figure B3. A complete list of subroutines is given below:

Subroutines

CAPPBT	Subroutine to calculate the capital cost required to result in a given payback time for the project.
DCFRR	Subroutine to calculate the Discounted Cash Flow rate of return of the project, given a cash flow table. Uses NPV in its iterations.
FINISH	Subroutine to close the output files and terminate the program.
FIN1	Subroutine to spool the results file to the printer, if required.
INITAL	Subroutine to read the base case data from the data file.
INPUT	Subroutine for interactive data entry.
NPV	Subroutine to calculate the Net Present Value of the project, given a cash flow table.
PAYBAK	Calculates the payback time of the project.
REQPBT	Calculates the annual cash flow required to be added to the cash flow table of the project to result in a given payback time for the project.
REQPB2	Subroutine used by REQPBT.

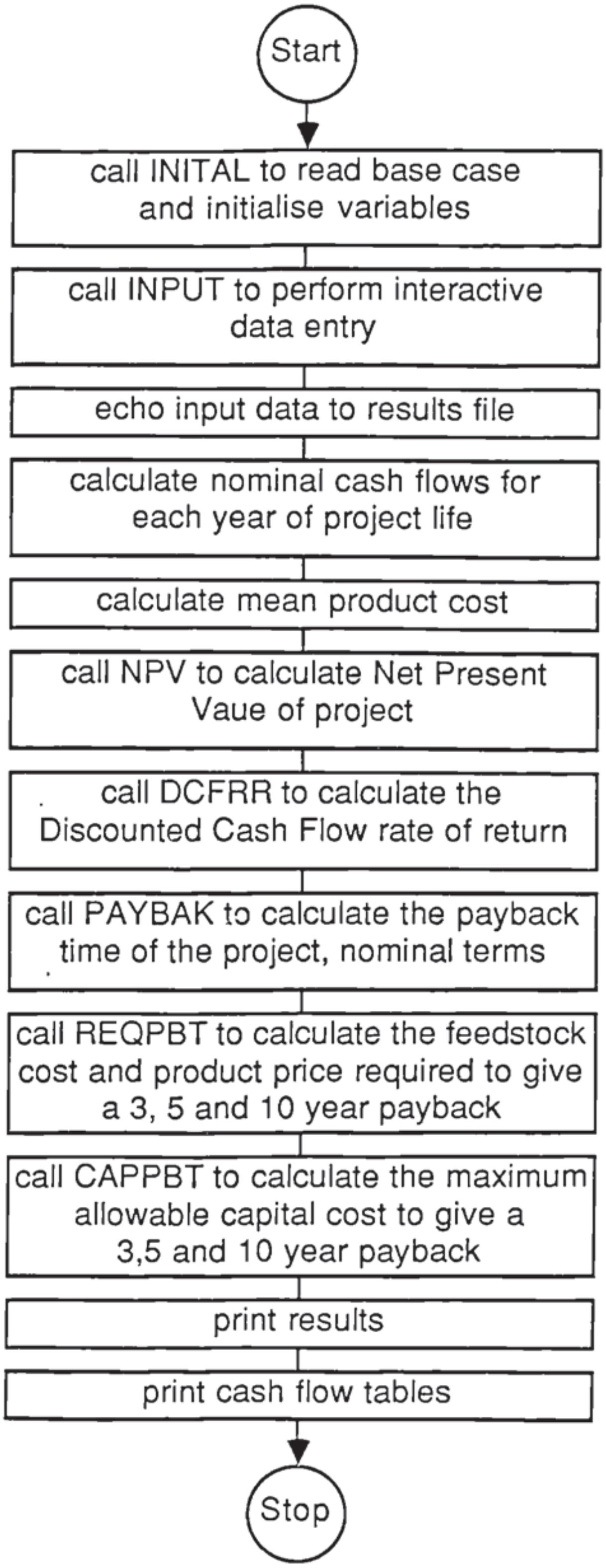


Figure B3 Flow Diagram of Program EVAL1

```
      PROGRAM EVAL1
      REAL UTILYR,LABYR,OVHDYR,MAINYR,RAWYR,PRYR,GJOUT,
     +PRESV,DCF,PBAK
      REAL RAWFL(0:25),PRFL(0:25),UTILFL(0:25),MAINFL(0:25),
     +RCASFL(0:25),CAPFL(0:25),OVHDFL(0:25),LABFL(0:25),
     +CASHFL(0:25)
      REAL INFLAT
      REAL QRAWY3,QRAWY5,QPRY3,QPRY5,QRAWC3,QRAWC5,QPRC3,QPRC5,
     +QRWY10,QPRY10,QRWC10,QPRC10,QCAP3,QCAP5,QCAP10
      REAL FLXRAW(0:25),FLXPR(0:25)
      REAL TEMP,AMORT,RELTOT,AVPRC
      REAL OTHCOS(25),TOTCOS(25),NOMPRC(25),RELPRC(25)
      CHARACTER*80 AGAIN,RUNNO
      REAL CAPCOS,CAPCTY,RAWHHV,EFF,RAWCOS,PRCOS,UTCOS,OVERHD,
     +NUMMEN,MAINTC,MANCOS,RDISC,NOMINT,
     +INFL,ESCRAW,ESCLAB,ESCPR,ESCUT,ESCOVH,ESCMTC
      INTEGER PRJLIF,HRPYR
      COMMON /BDATA/ CAPCOS,CAPCTY,RAWHHV,EFF,RAWCOS,PRCOS,UTCOS,
     +MAINTC,OVERHD,NUMMEN,MANCOS,PRJLIF,HRPYR,RDISC,NOMINT,
     +INFL,ESCRAW,ESCLAB,ESCPR,ESCUT,ESCOVH,ESCMTC
C
C
C A PROGRAM TO PERFORM FINANCIAL CALCULATIONS FOR GASIFICATION
C  SYSTEMS
C ALL CALCULATIONS ARE BEFORE TAX, IN REAL TERMS
C
C GLOBAL VARIABLES ARE AS FOLLOWS:
C     CAPCTY              CAPACITY OF GASIFIER T/H
C     RAWHHV              HHV OF RAW MATERIAL, MJ/KG
C     EFF                 EFFICIENCY OF GASIFIER %
C                         (DEFINED AS HEATING VALUE OF TOTAL PRODUCT
C                         DIVIDED BY HEATING VALUE OF FEEDSTOCK)
C     NUMMEN              NUMBER OF SHIFTS
C     PRJLIF              PROJECT LIFE, YEARS
C     HRPYR               NUMBER OF HOURS OF OPERATION EACH YEAR
C     CAPCOS              CAPITAL COST OF GASIFIER
C     RAWCOS              COST OF RAW MATERIAL, £/T
C     PRCOS               PRODUCT PRICE, £/GJ
C     UTCOS               UTILITIES COST, £/GJ PRODUCT GAS
C     MAINTC              MAINTENANCE COST/YEAR, £/£CAPITAL COST
C     OVERHD              OVERHEADS/YEAR, £/£CAPITAL COST
C     MANCOS              TOTAL COST OF EACH SHIFT, £/YEAR
C     RDISC               REAL TARGET DISCOUNT RATE
C     NOMINT              NOMINAL COST OF CAPITAL
C     INFL                RATE OF INFLATION
C     ESCRAW              REAL ESCALATION RATE OF FEEDSTOCK
C     ESCPR               REAL ESCALATION RATE OF PRODUCT PRICE
C     ESCUT               REAL ESCALATION RATE OF UTILITIES
C     ESCMTC              REAL ESCALATION RATE OF MAINTENANCE
C     ESCOVH              REAL ESCALATION RATE OF OVERHEADS
C     ESCLAB              REAL ESCALATION RATE OF LABOUR COST
C
C
C LOCAL VARIABLES ARE AS FOLLOWS:
C     CASHFL(0:25)        CASHFLOWS FOR 25 YEARS OF PROJECT, NOMINAL
```

```
C      RCASFL(0:25)        YEARLY REAL CASH FLOWS
C      RAWFL(0:25)         FEEDSTOCK YEARLY CASH FLOWS, NOMINAL
C      PRFL(0:25)          PRODUCT YEARLY CASH FLOWS, NOMINAL
C      UTILFL(0:25)        UTILITIES YEARLY CASH FLOWS, NOMINAL
C      MAINFL(0:25)        MAINTENANCE YEARLY CASH FLOWS, NOMINAL
C      OVHDFL(0:25)        OVERHEADS YEARLY CASH FLOWS, NOMINAL
C      LABFL(0:25)         LABOUR YEARLY CASH FLOWS, NOMINAL
C      CAPFL(0:25)         YEARLY CAPITAL CASH FLOWS, NOMINAL
C      UTILYR              UTILITY COST PER YEAR, REAL TERMS, NO ESC.
C      LABYR               LABOUR COST PER YEAR, AS UTILYR
C      OVHDYR              OVERHEADS PER YEAR, AS UTILYR
C      MAINYR              MAINTENANCE PER YEAR, AS UTILYR
C      RAWYR               RAW MATERIAL COST PER YEAR, AS UTILYR
C      PRYR                REVENUE FROM PROD. SALE PER YR, AS UTILYR
C      GJOUT               CAPACITY OF GASIFIER, GJ PRODUCT GAS/HOUR
C      PRESV               NET PRESENT VALUE OF PROJECT
C      DCF                 DCF OF PROJECT, REAL TERMS
C      DCFNOM              DCF OF PROJECT, NOMINAL TERMS
C      PBAK                PAYBACK TIME,YEARS
C      INFLAT              MULTIPLYING FACTOR FOR INFLATION, CASH
C                            FLOW CALC.
C      TEMP                TEMPORARY VARIABLE IN AMORTISATION CALC.
C      AMORT               YEARLY AMORTISATION
C      OTHCOS(25)          NOMINAL YEARLY COSTS, EXCLUDING
C                            AMORTISATION
C      TOTCOS(25)          NOMINAL YEARLY COSTS, INCLUDING
C                            AMORTISATION
C      NOMPRC(25)          NOMINAL PRODUCT COST
C      RELPRC(25)          REAL PRODUCT COST
C      RELTOT              RUNNING TOTAL OF REAL PRODUCT COSTS,
C                            TO CALC MEAN
C      AVPRC               MEAN REAL PRODUCT COST, £/GJ
C      QRAWY3              VALUE OF RAWYR TO GIVE 3 YEAR PAYBACK
C      QRAWY5              VALUE OF RAWYR TO GIVE 5 YEAR PAYBACK
C      QRWY10              VALUE OF RAWYR TO GIVE 10 YEAR PAYBACK
C      QPRY3               VALUE OF PRYR TO GIV 3 YEAR PAYBACK
C      QPRY5               VALUE OF PRYR TO GIVE 5 YEAR PAYBACK
C      QPRY10              VALUE OF PRYR TO GIVE 10 YEAR PAYBACK
C      QRAWC3              VALUE OF RAWCOS TO GIVE 3 YEAR PAYBACK
C      QRAWC5              VALUE OF RAWCOS TO GIVE 5 YEAR PAYBACK
C      QRWC10              VALUE OF RAWCOS TO GIVE 10 YEAR PAYBACK
C      QPRC3               VALUE OF PRCOS TO GIVE 3 YEAR PAYBACK
C      QPRC5               VALUE OF PRCOS TO GIVE 5 YEAR PAYBACK
C      QPRC10              VALUE OF PRCOS TO GIVE 10 YEAR PAYBACK
C      QCAP3               VALUE OF CAPITAL COST TO GIVE 3 YEAR
C                            PAYBACK
C      QCAP5               VALUE OF CAPITAL COST TO GIVE 5 YEAR
C                            PAYBACK
C      QCAP10              VALUE OF CAPITAL COST TO GIVE 10 YEAR
C                            PAYBACK
C      FLXRAW(0:25)        YEARLY CASH FLOWS, EXCL. FEEDSTOCK COST
C      FLXPR(0:25)         YEARLY CASH FLOWS, EXCL. PRODUCT SALES
C
       OPEN(UNIT=20,FILE='2299JMD*R:EVAL',STATUS='OLD')
       CALL INITAL
```

```
10    CALL INPUT
      WRITE(3,*) ' '
      WRITE(3,*) 'ENTER A RUN NUMBER TO IDENTIFY THIS RUN'
      READ(0,89999) RUNNO
      GJOUT=EFF*RAWHHV*CAPCTY/100.0
      UTILYR=GJOUT*HRPYR*UTCOS
      MAINYR=MAINTC*CAPCOS
      OVHDYR=OVERHD*CAPCOS
      PRYR=PRCOS*GJOUT*HRPYR
      RAWYR=RAWCOS*CAPCTY*HRPYR
      LABYR=NUMMEN*MANCOS
C ECHO DATA TO RESULTS FILE BEFORE PERFORMING CALCULATION
      WRITE(20,90000) RUNNO
      WRITE(20,90001) CAPCTY
      WRITE(20,90002) RAWHHV
      WRITE(20,90003) EFF
      WRITE(20,90004) NUMMEN
      WRITE(20,90005) PRJLIF
      WRITE(20,90006) HRPYR
      WRITE(20,90009) CAPCOS
      WRITE(20,90010) RAWCOS
      WRITE(20,90011) PRCOS
      WRITE(20,90012) UTCOS
      WRITE(20,90013) MAINTC
      WRITE(20,90014) OVERHD
      WRITE(20,90015) MANCOS
      WRITE(20,90016) RDISC
      WRITE(20,90017) NOMINT
      WRITE(20,90018) INFL
      WRITE(20,90019) ESCRAW
      WRITE(20,90020) ESCPR
      WRITE(20,90021) ESCUT
      WRITE(20,90022) ESCMTC
      WRITE(20,90023) ESCOVH
      WRITE(20,90024) ESCLAB
C
C CALCULATE YEARLY CASH FLOWS, NOMINAL TERMS
      RAWFL(0)=0.0
      PRFL(0)=0.0
      UTILFL(0)=0.0
      MAINFL(0)=0.0
      OVHDFL(0)=0.0
      LABFL(0)=0.0
      CAPFL(0)=-CAPCOS
      CASHFL(0)=CAPFL(0)
      RCASFL(0)=CAPFL(0)
      FLXRAW(0)=CAPFL(0)
      FLXPR(0)=CAPFL(0)
      DO 100 IYEAR=1,PRJLIF
        INFLAT=(1.0+INFL/100.0)**IYEAR
        RAWFL(IYEAR)=-RAWYR*INFLAT*((1.0+ESCRAW/100.0)**IYEAR)
        PRFL(IYEAR)=PRYR*INFLAT*((1.0+ESCPR/100.0)**IYEAR)
        UTILFL(IYEAR)=-UTILYR*INFLAT*((1.0+ESCUT/100.0)**IYEAR)
        MAINFL(IYEAR)=-MAINYR*INFLAT*((1.0+ESCMTC/100.0)**IYEAR)
        OVHDFL(IYEAR)=-OVHDYR*INFLAT*((1.0+ESCOVH/100.0)**IYEAR)
        LABFL(IYEAR)=-LABYR*INFLAT*((1.0+ESCLAB/100.0)**IYEAR)
```

```
        CAPFL(IYEAR)=0.0
        CASHFL(IYEAR)=CAPFL(IYEAR)+RAWFL(IYEAR)+PRFL(IYEAR)+
     +    UTILFL(IYEAR)+MAINFL(IYEAR)+OVHDFL(IYEAR)+LABFL(IYEAR)
        FLXRAW(IYEAR)=CASHFL(IYEAR)-RAWFL(IYEAR)
        FLXPR(IYEAR)=CASHFL(IYEAR)-PRFL(IYEAR)
        RCASFL(IYEAR)=CASHFL(IYEAR)/INFLAT
100   CONTINUE
C
C PERFORM PRODUCT COST CALCULATIONS
C
C  CALCULATE YEARLY AMORTISATION
      TEMP=(1.+NOMINT/100.)**PRJLIF
      AMORT=CAPCOS*((NOMINT/100.)*TEMP)/(TEMP-1.)
C  CALCULATE YEARLY COSTS, AND HENCE PRODUCT COSTS FOR EACH YEAR
      RELTOT=0.0
      DO 200 IYEAR=1,PRJLIF
        OTHCOS(IYEAR)=-(RAWFL(IYEAR)+UTILFL(IYEAR)+MAINFL(IYEAR)+
     +     OVHDFL(IYEAR)+LABFL(IYEAR))
        TOTCOS(IYEAR)=AMORT+OTHCOS(IYEAR)
        NOMPRC(IYEAR)=TOTCOS(IYEAR)/(GJOUT*HRPYR)
        RELPRC(IYEAR)=NOMPRC(IYEAR)/((1.+INFL/100.)**IYEAR)
        RELTOT=RELTOT+RELPRC(IYEAR)
200   CONTINUE
C  CALCULATE MEAN PRODUCT COST
      AVPRC=RELTOT/(PRJLIF*1.0)
C
C
C CALL NET PRESENT VALUE SUBROUTINE
      CALL NPV(RCASFL,PRJLIF,RDISC,PRESV)
C CALL DISCOUNTED CASH FLOW RATE OF RETURN SUBROUTINE
      LAB1=0
      CALL DCFRR(RCASFL,PRJLIF,DCF,LAB1)
C CALL PAYBACK TIME SUBROUTINE
      LAB2=0
      CALL PAYBAK(CASHFL,PRJLIF,PBAK,LAB2)
C PERFORM CALCULATIONS OF FEEDSTOCK COST OR PRODUCT PRICE
C  OR CAPITAL COST FOR 3, 5, & 10 YEAR PAYBACK PERIODS
C
C 3 YEAR PAYBACK
      IF (PRJLIF.GE.3) THEN
        CALL REQPBT(3.0,FLXRAW,PRJLIF,INFL,ESCRAW,QRAWY3,LAB31)
        IF (LAB31.EQ.0) QRAWC3=-QRAWY3/(CAPCTY*HRPYR)
        CALL REQPBT(3.0,FLXPR,PRJLIF,INFL,ESCPR,QPRY3,LAB32)
        IF (LAB32.EQ.0) QPRC3=QPRY3/(GJOUT*HRPYR)
        CALL CAPPBT(3.0,CASHFL,PRJLIF,QCAP3)
      END IF
C 5 YEAR PAYBACK
      IF (PRJLIF.GE.5) THEN
        CALL REQPBT(5.0,FLXRAW,PRJLIF,INFL,ESCRAW,QRAWY5,LAB51)
        IF (LAB51.EQ.0) QRAWC5=-QRAWY5/(CAPCTY*HRPYR)
        CALL REQPBT(5.0,FLXPR,PRJLIF,INFL,ESCPR,QPRY5,LAB52)
        IF (LAB52.EQ.0) QPRC5=QPRY5/(GJOUT*HRPYR)
        CALL CAPPBT(5.0,CASHFL,PRJLIF,QCAP5)
      END IF
C 10 YEAR PAYBACK
```

```
      IF (PRJLIF.GE.10) THEN
        CALL REQPBT(10.0,FLXRAW,PRJLIF,INFL,ESCRAW,QRWY10,LAB101)
        IF (LAB101.EQ.0) QRWC10=-QRWY10/(CAPCTY*HRPYR)
        CALL REQPBT(10.0,FLXPR,PRJLIF,INFL,ESCPR,QPRY10,LAB102)
        IF (LAB102.EQ.0) QPRC10=QPRY10/(GJOUT*HRPYR)
        CALL CAPPBT(10.0,CASHFL,PRJLIF,QCAP10)
      END IF
C
C PRINT RESULTS
C
C
      DO 10000 IOUT=3,20,17
        WRITE(IOUT,99000) RDISC,PRESV
        IF (LAB1.EQ.0) THEN
          DCFNOM=(1+DCF/100.)*(1+INFL/100.)-1.0
          DCFNOM=DCFNOM*100.
          WRITE(IOUT,99001) DCF,DCFNOM
        ELSE IF (LAB1.LT.0)
          WRITE(IOUT,99999)
        ELSE
          WRITE(IOUT,99998)
        END IF
        IF (LAB2.EQ.0) THEN
          WRITE(IOUT,99002) PBAK
        ELSE
          WRITE(IOUT,99997)
        END IF
        WRITE(IOUT,99003) AVPRC
        IF (PRJLIF.GE.3) THEN
          WRITE(IOUT,99004) 3
          IF (LAB31.EQ.0) THEN
            WRITE(IOUT,99005) QRAWC3
          ELSE
            WRITE(IOUT,99996) LAB31
          END IF
          IF (LAB32.EQ.0) THEN
            WRITE(IOUT,99006) QPRC3
          ELSE
            WRITE(IOUT,99996) LAB32
          END IF
          WRITE(IOUT,99007) QCAP3
        END IF
C
        IF (PRJLIF.GE.5) THEN
          WRITE(IOUT,99004) 5
          IF (LAB51.EQ.0) THEN
            WRITE(IOUT,99005) QRAWC5
          ELSE
            WRITE(IOUT,99996) LAB51
          END IF
          IF (LAB52.EQ.0) THEN
            WRITE(IOUT,99006) QPRC5
          ELSE
            WRITE(IOUT,99996) LAB52
          END IF
```

```
          WRITE(IOUT,99007) QCAP5
        END IF
C
        IF (PRJLIF.GE.10) THEN
          WRITE(IOUT,99004) 10
          IF (LAB101.EQ.0) THEN
            WRITE(IOUT,99005) QRWC10
          ELSE
            WRITE(IOUT,99996) LAB101
          END IF
          IF (LAB102.EQ.0) THEN
            WRITE(IOUT,99006) QPRC10
          ELSE
            WRITE(IOUT,99996) LAB102
          END IF
          WRITE(IOUT,99007) QCAP10
        END IF
10000  CONTINUE
C
C  PRINT CASH FLOW TABLE FOR NPV CALCULATIONS
      WRITE(20,99900) RUNNO
      DO 500 IYEAR=0,PRJLIF
        WRITE(20,99901)
IYEAR,CAPFL(IYEAR),PRFL(IYEAR),RAWFL(IYEAR),
     +   UTILFL(IYEAR),MAINFL(IYEAR),OVHDFL(IYEAR),LABFL(IYEAR),
     +   CASHFL(IYEAR),RCASFL(IYEAR)
500   CONTINUE
C
C
C PRINT CASH FLOW TABLE FOR PRODUCT COSTS
      WRITE(20,99902)
      DO 600 IYEAR=1,PRJLIF
        WRITE(20,99903) IYEAR,AMORT,OTHCOS(IYEAR),TOTCOS(IYEAR),
     +    NOMPRC(IYEAR),RELPRC(IYEAR)
600   CONTINUE
      WRITE(3,*) 'PRESS <RETURN> TO CONTINUE'
      READ(0,89999) AGAIN
      GOTO 10
C
C
89999 FORMAT(A80)
90000 FORMAT('1',14X,45('*'),/,15X,'PROGRAM  EVAL   -  ',
     +'GASIFIER EVALUATION PROGRAM',/,15X,17('*'),
     +' J M Double ',16('*'),/,'0',9X,'RUN NUMBER: ',A80,/,
     +'0',9X,'DATA USED FOR CALCULATION ',/,15X,
     +'(for £ read Pounds Sterling, all tonnes are d.a.f. basis)')
90001 FORMAT('0',14X,'Gasifier throughput',23X,F5.2,'  tonnes/h')
90002 FORMAT(15X,'Feedstock heating value',19X,F5.2,'  GJ/tonne')
90003 FORMAT(15X,'Gasifier efficiency',23X,F5.2,'  %')
90004 FORMAT(15X,'Number of shifts                     ',5X,F5.1)
90005 FORMAT(15X,'Project life',30X,I2,5X,'years')
90006 FORMAT(15X,'Number of operating hours per year',6X,I4)
90009 FORMAT('0',14X,'Capital cost (inc working capital) ',
     +F12.2,'  £')
90010 FORMAT(15X,'Feedstock cost',28X,F5.2,'  £/tonne')
90011 FORMAT(15X,'Product selling price',21X,F6.3,' £/GJ')
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90012 FORMAT(15X,'Utilities cost',28X,F6.3,' £/GJ')
90013 FORMAT(15X,'Yearly maintenance cost,',
     +' fraction capcost  ',F5.3)
90014 FORMAT(15X,'Yearly overheads, fraction capcost',9X,F5.3)
90015 FORMAT(15X,'Total cost of labour per shift',8X,F9.2
     +,'  £/year')
90016 FORMAT('0',14X,'Target real discount rate',17X,F5.2,'  %')
90017 FORMAT(15X,'Nominal cost of capital',19X,F5.2,'  %')
90018 FORMAT(15X,'Inflation rate',28X,F5.2,'  %')
90019 FORMAT(15X,'Feedstock cost real escalation rate',7X,
     +F5.2,'  %')
90020 FORMAT(15X,'Product price real escalation rate',8X,
     +F5.2,'  %')
90021 FORMAT(15X,'Utilities cost real escalation rate',7X,
     +F5.2,'  %')
90022 FORMAT(15X,'Maintenance cost real escalation rate',5X,
     +F5.2,'  %')
90023 FORMAT(15X,'Overheads real escalation rate',12X,F5.2,'  %')
90024 FORMAT(15X,'Labour cost real escalation rate',10X,
     +F5.2,'  %')
99000 FORMAT(/,'0',9X,'RESULTS',/,
     +'0',14X,'NPV at ',F6.2,' % real discount rate is £',
     +F14.2)
99001 FORMAT(15X,'DCF rate of return is ',F6.2,
     +' % real, ',F6.2,' % nominal')
99002 FORMAT(15X,'Payback time is ',F5.2,' years from start-up,',
     +' nominal basis')
99003 FORMAT(15X,'Mean real product cost over project life is ',
     +F6.3,' £/GJ')
99004 FORMAT('0',14X,'FOR A ',I2,' YEAR PAYBACK TIME:')
99005 FORMAT(15X,'Either FEEDSTOCK COST must be ',F7.2,' £/tonne')
99006 FORMAT(19X,'or PRODUCT SELLING PRICE must be ',F5.2,' £/GJ')
99007 FORMAT(19X,'or CAPITAL COST must be £',F14.2)
99900 FORMAT('1RUN NUMBER: ',A80,/,
     +'0NOMINAL CASH FLOW TABLE FOR N.P.V. CALCULATIONS',
     +' (all cash flows expressed in Pounds Sterling)',/,
     +'     Year',6X,'Capital     Product   Feedstock   ',
     +'Utilities Maintenance   Overheads      Labour      ',
     +'TOTAL NOMINAL',6X,'TOTAL REAL')
99901 FORMAT(6X,I2,2X,F12.2,6(1X,F11.2),2X,2(2X,F14.2))
99902 FORMAT('0NOMINAL CASH FLOW TABLE FOR PRODUCT COST ',
     +'CALCULATIONS ',
     +'(all cash flows expressed in Pounds Sterling)',
     +/,5X,'Year   Amortisation',5X,'Other costs',5X,
     +'Total costs   Product cost £/GJ',/,29X,'(see above)',
     +19X,'Nominal',6X,'Real')
99903 FORMAT(6X,I2,3(2X,F14.2),2(4X,F6.3))
99996 FORMAT(20X,'ERROR NUMBER ',I2,' IN CALLING REQPBT')
99997 FORMAT(15X,'Payback time is greater than project life')
99998 FORMAT(15X,'DCF rate of return is greater than ',
     +'500.00 % real')
99999 FORMAT(15X,'DCF rate of return is less than -90.00 % real')
      END
C
C
```

```
      SUBROUTINE INITAL
      REAL CAPCOS,CAPCTY,RAWHHV,EFF,RAWCOS,PRCOS,UTCOS,OVERHD,
     +NUMMEN,MAINTC,MANCOS,RDISC,NOMINT,
     +INFL,ESCRAW,ESCLAB,ESCPR,ESCUT,ESCOVH,ESCMTC
      INTEGER PRJLIF,HRPYR
      COMMON /BDATA/ CAPCOS,CAPCTY,RAWHHV,EFF,RAWCOS,PRCOS,UTCOS,
     +MAINTC,OVERHD,NUMMEN,MANCOS,PRJLIF,HRPYR,RDISC,NOMINT,
     +INFL,ESCRAW,ESCLAB,ESCPR,ESCUT,ESCOVH,ESCMTC
      OPEN(UNIT=21,FILE='        D:EVAL1',STATUS='OLD')
      READ(21,*) CAPCTY
      READ(21,*) RAWHHV
      READ(21,*) EFF
      READ(21,*) NUMMEN
      READ(21,*) PRJLIF
      READ(21,*) HRPYR
      READ(21,*) CAPCOS
      READ(21,*) RAWCOS
      READ(21,*) PRCOS
      READ(21,*) UTCOS
      READ(21,*) MAINTC
      READ(21,*) OVERHD
      READ(21,*) MANCOS
      READ(21,*) RDISC
      READ(21,*) NOMINT
      READ(21,*) INFL
      READ(21,*) ESCRAW
      READ(21,*) ESCPR
      READ(21,*) ESCUT
      READ(21,*) ESCMTC
      READ(21,*) ESCOVH
      READ(21,*) ESCLAB
      CLOSE 21
      RETURN
      END
C
C
C
      SUBROUTINE FINISH
      WRITE(20,90000)
      CLOSE 20
      CALL FIN1
      STOP
90000 FORMAT('1')
      END
C
C
C
C
      SUBROUTINE INPUT
      REAL CAPCOS,CAPCTY,RAWHHV,EFF,RAWCOS,PRCOS,UTCOS,OVERHD,
     +NUMMEN,MAINTC,MANCOS,RDISC,NOMINT,
     +INFL,ESCRAW,ESCLAB,ESCPR,ESCUT,ESCOVH,ESCMTC
      INTEGER PRJLIF,HRPYR
      COMMON /BDATA/ CAPCOS,CAPCTY,RAWHHV,EFF,RAWCOS,PRCOS,UTCOS,
     +MAINTC,OVERHD,NUMMEN,MANCOS,PRJLIF,HRPYR,RDISC,NOMINT,
```

```
     +INFL,ESCRAW,ESCLAB,ESCPR,ESCUT,ESCOVH,ESCMTC
10    DO 11 I=1,3
        WRITE(3,*) ' '
11    CONTINUE
      WRITE(3,*) '*** PROGRAM EVAL ',
     +'- GASIFIER ECONOMIC CALCULATIONS ***'
      WRITE(3,90000)
      WRITE(3,90001) CAPCTY,PRJLIF
      WRITE(3,90002) RAWHHV,HRPYR
      WRITE(3,90003) EFF
      WRITE(3,90004) NUMMEN
      WRITE(3,90005)
      WRITE(3,90006) CAPCOS
      WRITE(3,90007) RAWCOS,ESCRAW
      WRITE(3,90008) PRCOS,ESCPR
      WRITE(3,90009) UTCOS,ESCUT
      WRITE(3,90010) MAINTC,ESCMTC
      WRITE(3,90011) OVERHD,ESCOVH
      WRITE(3,90012) MANCOS,ESCLAB
      WRITE(3,90013) RDISC
      WRITE(3,90014) NOMINT
      WRITE(3,90015) INFL
      WRITE(3,90016)
      WRITE(3,90017)
      WRITE(3,90018)
      WRITE(3,90019)
      WRITE(3,90020)
      READ(0,*) ICHANG
      IF (ICHANG.EQ.0) CALL FINISH
      IF (ICHANG.EQ.98) GOTO 9800
      IF (ICHANG.EQ.99) GOTO 9900
      IF (ICHANG.EQ.100) RETURN
      GOTO(100,200,300,400,500,600,700,800,900,1000,1100,
     +1200,1300,1400,1500,1600,1700,1800,1900,2000,
     +2100,2200,2300,2400) ICHANG
      WRITE(3,*) 'INCORRECT INPUT, TRY AGAIN'
      GOTO 10
100   WRITE(3,*) 'THE CAPACITY OF THE GASIFIER IS ',CAPCTY,' t/h'
      WRITE(3,*) 'ENTER THE NEW CAPACITY'
      READ(0,*) CAPCTY
      GOTO 900
200   WRITE(3,*) 'THE FEEDSTOCK HEATING VALUE IS ',RAWHHV,' MJ/kg'
      WRITE(3,*) 'ENTER THE NEW HEATING VALUE'
      READ(0,*) RAWHHV
      GOTO 10
300   WRITE(3,*) 'THE GASIFIER EFFICIENCY IS ',EFF,' %'
      WRITE(3,*) 'ENTER THE NEW EFFICIENCY'
      READ(0,*) EFF
      GOTO 10
400   WRITE(3,*) 'THERE ARE ',NUMMEN,' EMPLOYED ON THE GASIFIER'
      WRITE(3,*) 'ENTER THE NEW NUMBER OF MEN'
      READ(0,*) NUMMEN
      GOTO 10
500   WRITE(3,*) 'THE PROJECT LIFE IS ',PRJLIF,' years'
      WRITE(3,*) 'ENTER THE NEW PROJECT LIFE'
      READ(0,*) PRJLIF
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      GOTO 10
600   WRITE(3,*) HRPYR,' HOURS ARE WORKED EACH YEAR'
      WRITE(3,*) 'ENTER THE NEW HOURS/YEAR'
      READ(0,*) HRPYR
      GOTO 10
700   WRITE(3,*) 'INCORRECT INPUT, TRY AGAIN'
      GOTO 10
800   WRITE(3,*) 'INCORRECT INPUT, TRY AGAIN'
      GOTO 10
900   WRITE(3,*) 'CAPITAL COST IS CURRENTLY £',CAPCOS
      WRITE(3,*) 'ENTER THE NEW CAPITAL COST'
      READ(0,*) CAPCOS
      GOTO 10
1000  WRITE(3,*) 'THE FEEDSTOCK COST IS ',RAWCOS,' £/tonne'
      WRITE(3,*) 'ENTER THE NEW FEEDSTOCK COST'
      READ(0,*) RAWCOS
      GOTO 10
1100  WRITE(3,*) 'THE PRODUCT COST IS ',PRCOS,' £/GJ'
      WRITE(3,*) 'ENTER THE NEW PRODUCT COST'
      READ(0,*) PRCOS
      GOTO 10
1200  WRITE(3,*) 'THE UTILITIES COST IS ',UTCOS,' £/GJ PRODUCT'
      WRITE(3,*) 'ENTER THE NEW UTILITY COST'
      READ(0,*) UTCOS
      GOTO 10
1300  WRITE(3,*) 'THE MAINTENANCE COST IS ',MAINTC,' FRACTION',
     +' OF CAPITAL COST'
      WRITE(3,*) 'ENTER THE NEW MAINTENANCE COST'
      READ(0,*) MAINTC
      GOTO 10
1400  WRITE(3,*) 'THE OVERHEADS ARE ',OVERHD,' FRACTION OF ',
     +'CAPITAL COST'
      WRITE(3,*) 'ENTER THE NEW OVERHEADS'
      READ(0,*) OVERHD
      GOTO 10
1500  WRITE(3,*) 'THE COST PER MAN IS ',MANCOS,' £/year'
      WRITE(3,*) 'ENTER THE NEW COST PER MAN'
      READ(0,*) MANCOS
      GOTO 10
1600  WRITE(3,*) 'THE TARGET REAL DISCOUNT RATE IS ',RDISC,' %'
      WRITE(3,*) 'ENTER THE NEW TARGET DISCOUNT RATE'
      READ(0,*) RDISC
      GOTO 10
1700  WRITE(3,*) 'THE NOMINAL COST OF CAPITAL IS ',NOMINT,' %'
      WRITE(3,*) 'ENTER THE NEW COST OF CAPITAL'
      READ(0,*) NOMINT
      GOTO 10
1800  WRITE(3,*) 'THE INFLATION RATE IS ',INFL,' %'
      WRITE(3,*) 'ENTER THE NEW INFLATION RATE'
      READ(0,*) INFL
      GOTO 10
1900  WRITE(3,*) 'THE FEEDSTOCK REAL ESCALATION IS ',ESCRAW,' %'
      WRITE(3,*) 'ENTER THE NEW FEEDSTOCK ESCALATION'
      READ(0,*) ESCRAW
      GOTO 10
```

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2000  WRITE(3,*) 'THE PRODUCT PRICE REAL ESCALATION IS ',ESCPR,
     +' %'
      WRITE(3,*) 'ENTER THE NEW PRODUCT PRICE ESCALATION'
      READ(0,*) ESCPR
      GOTO 10
2100  WRITE(3,*) 'THE UTILITIES REAL ESCALATION IS ',ESCUT,' %'
      WRITE(3,*) 'ENTER THE NEW UTILITIES ESCALATION'
      READ(0,*) ESCUT
      GOTO 10
2200  WRITE(3,*) 'THE MAINTENANCE REAL ESCALATION IS ',ESCMTC,' %'
      WRITE(3,*) 'ENTER THE NEW MAINTENANCE ESCALATION'
      READ(0,*) ESCMTC
      GOTO 10
2300  WRITE(3,*) 'THE OVERHEADS REAL ESCALATION IS ',ESCOVH,' %'
      WRITE(3,*) 'ENTER THE NEW OVERHEADS ESCALATION'
      READ(0,*) ESCOVH
      GOTO 10
2400  WRITE(3,*) 'THE LABOUR COST REAL ESCALATION IS ',ESCLAB,' %'
      WRITE(3,*) 'ENTER THE NEW LABOUR ESCALATION'
      READ(0,*) ESCLAB
      GOTO 10
9800  CALL INITAL
      GOTO 10
9900  OPEN(UNIT=21,FILE='2299JMD*D:EVAL1',STATUS='OLD')
      WRITE(21,*) CAPCTY
      WRITE(21,*) RAWHHV
      WRITE(21,*) EFF
      WRITE(21,*) NUMMEN
      WRITE(21,*) PRJLIF
      WRITE(21,*) HRPYR
      WRITE(21,*) CAPCOS
      WRITE(21,*) RAWCOS
      WRITE(21,*) PRCOS
      WRITE(21,*) UTCOS
      WRITE(21,*) MAINTC
      WRITE(21,*) OVERHD
      WRITE(21,*) MANCOS
      WRITE(21,*) RDISC
      WRITE(21,*) NOMINT
      WRITE(21,*) INFL
      WRITE(21,*) ESCRAW
      WRITE(21,*) ESCPR
      WRITE(21,*) ESCUT
      WRITE(21,*) ESCMTC
      WRITE(21,*) ESCOVH
      WRITE(21,*) ESCLAB
      CLOSE 21
      GOTO 10
90000 FORMAT('  ENTER THE NUMBER OF THE VARIABLE TO BE CHANGED',
     +' OR ACTION REQUIRED:')
90001 FORMAT('   1. Gasifier throughput',9X,F6.2,' t/h     ',
     +'5. Project life, years',4X,I3)
90002 FORMAT('   2. HHV of feedstock',12X,F6.2,' GJ/t    ',
     +'6. Working hours / year  ',I4)
90003 FORMAT('   3. Gasifier efficiency',9X,F5.1,'  %')
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90004 FORMAT('   4. Total number of shifts        ',F5.1)
90005 FORMAT('  FINANCIAL DATA:')
90006 FORMAT('   9. Capital cost',10X,F12.2,' £')
90007 FORMAT('  10. Feedstock cost',14X,F6.2,' £/t    ',
     +'19. Feedstock escalation',4X,F5.2,'%')
90008 FORMAT('  11. Product cost',17X,F6.3,'£/GJ   ',
     +'20. Product cost escalation ',F5.2,'%')
90009 FORMAT('  12. Utilities',20X,F6.3,'£/GJ   ',
     +'21. Utilities escalation',4X,F5.2,'%')
90010 FORMAT('  13. Maintenance (frac cap cost)    ',F5.3,6X,
     +'22. Maintenance escalation  ',F5.2,'%')
90011 FORMAT('  14. Overheads (frac cap cost)',5X,F5.3,6X,
     +'23. Overheads escalation',4X,F5.2,'%')
90012 FORMAT('  15. Total cost per shift',5X,F9.2,' £/yr  24.',
     +' Labour escalation',7X,F5.2,'%')
90013 FORMAT('  16. Real target discount rate',3X,F6.2,' %')
90014 FORMAT('  17. Nominal cost of capital',5X,F6.2,' %')
90015 FORMAT('  18. Inflation rate',14X,F6.2,' %')
90016 FORMAT('0  0. STOP THE PROGRAM')
90017 FORMAT('  98. GO BACK TO THE BASE CASE')
90018 FORMAT('  99. STORE THESE FIGURES AS THE NEW BASE CASE')
90019 FORMAT(' 100. START THE CALCULATION')
90020 FORMAT(' ** ENTER THE NUMBER OF THE ACTION REQUIRED **')
      END
C
C
C
      SUBROUTINE NPV(CASFLO,PRLIFE,DISRAT,PRVAL)
      REAL CASFLO(0:PRLIFE),PRVAL,DISRAT
      INTEGER PRLIFE
C SUBROUTINE TO PERFORM NET PRESENT VALUE CALCULATION, GIVEN
C CASH FLOW TABLE.
C VARIABLES:
C       CASFLO(0:PRLIFE)        CASH FLOWS FOR EACH YR OF PROJ LIFE
C       PRLIFE                  NUMBER OF YEARS OF PROJECT LIFE
C       DISRAT                  DISCOUNT RATE
C       PRVAL                   NET PRESENT VALUE
C
C
      PRVAL=0
      DO 200 I=0,PRLIFE
        PRVAL=PRVAL+CASFLO(I)/(1+DISRAT/100.0)**I
200   CONTINUE
      RETURN
      END
C
C
      SUBROUTINE DCFRR(CASFLO,PRLIF,DCF,LABEL)
      REAL DCF,CASFLO(0:PRLIFE)
      INTEGER PRLIF,LABEL
C SUBROUTINE TO PERFORM DCF CALCULATION, GIVEN
C CASH FLOW TABLE
C VARIABLES:
C       CASFLO(0:PRLIFE)    CASH FLOWS FOR EACH YEAR OF PROJ LIFE
C       DCF                 FINAL CALCULATED VALUE OF DCF
C       DCFL                LOWER BOUND OF DCF FOR ITERATION
```

```
C       DCFU                    UPPER BOUND OF DCF FOR ITERATION
C       PRLIF                   PROJECT LIFE, YEARS
C       LABEL                   LABEL TO INDICATE FAILURE OF CALC
      DCFL=-90.0
      DCFU=500.0
      CALL NPV(CASFLO,PRLIF,DCFL,PRV)
      IF (PRV.LE.0.0) THEN
        LABEL=-1
        DCF=DCFL
        RETURN
      END IF
      CALL NPV(CASFLO,PRLIF,DCFU,PRV)
      IF (PRV.GT.0.0) THEN
        LABEL=1
        DCF=DCFU
        RETURN
      END IF
      DO 1000 I=1,100
        DCF=(DCFU+DCFL)/2.0
        CALL NPV(CASFLO,PRLIF,DCF,PRV)
        IF (ABS(PRV).LT.0.001) RETURN
        IF (PRV.GT.0.0) THEN
          DCFL=DCF
        ELSE
          DCFU=DCF
        END IF
1000  CONTINUE
      WRITE(3,*) 'DCF CALCULATION HAS NOT CONVERGED'
      CALL FINISH
      END
C
C
C
      SUBROUTINE PAYBAK(CASHFL,PRJLIF,PBACKT,LABEL)
      REAL CASHFL(0:PRJLIF)
      REAL PBACKT
      INTEGER PRJLIF,LABEL
C A SUBROUTINE TO CALCULATE PAYBACK TIME GIVEN A CASH FLOW TABLE
C ASSUMES CUMULATIVE CASHFLOW CURVE IS A STRAIGHT LINE BETWEEN
C YEAR ENDS
C VARIABLES:
C     CASHFL(0:PRJLIF)     ARRAY CONTAINING CASH FLOW TABLE
C     PBACKT               PAYBACK TIME
C     PRJLIF               PROJECT LIFE IN YEARS
C     LABEL                LABEL INDICATING FAILURE OF CALCULATION
C
      CUMCF=0
      DO 20 I=0,PRJLIF
        CUMCF=CUMCF+CASHFL(I)
        IF (CUMCF.GT.0.0) GOTO 30
20    CONTINUE
C IF THIS LOOP HAS BEEN COMPLETED, THEN THE PROJECT NEVER PAYS
C  BACK ITS CAPITAL. SET PAYBACK TIME TO PROJECT LIFE + 1 YEAR,
C  AND SET LABEL(ERROR INDICATOR) TO 1
      PBACKT=PRJLIF+1.0
```

```
      LABEL=1
      RETURN
C INPERPOLATE USING STRAIGHT LINE INTERPOLATION TO
C  GET EXACT PAYBACK TIME
30    XX=CUMCF/CASHFL(I)
      PBACKT=I-XX
      LABEL=0
      RETURN
      END
C
C
C
      SUBROUTINE REQPBT(RPBT,CFLO,PRJLIF,INFL,ESC,ANSW,LAB)
      REAL RPBT,CFLO(0:PRJLIF),INFL,ESC,ANSW
      INTEGER PRJLIF
      REAL ANSL,ANSU,PBTL,PBTU,PBTM
C SUBROUTINE TO CALCULATE YEARLY CASH FLOW NEEDED TO GIVE A
C  REQUIRED PAYBACK TIME, GIVEN A TABLE OF THE SUM OF THE OTHER
C  CASH FLOWS, AND ESCALATION AND INFLATION RATES
C
C THIS SUBROUTINE USES SUBROUTINE REQPB2 DURING THE ITERATIVE
C   PROCEDURE
C
C VARIABLES USED:
C     RPBT                 REQUIRED PAY BACK TIME
C     CFLO(0:PRJLIF)       TABLE OF TOTAL CASH FLOWS, EXCLUDING
C                           THE VARIABLE TO BE FOUND
C     PRJLIF               PROJECT LIFE, YEARS
C     INFL                 OVERALL INFLATION RATE, %
C     ESC                  REAL ESCALATION RATE OF ITERATED
C                           VARIABLE,%
C     ANSW                 RESULT OF THE CALCULATION
C LOCAL VARIABLES
C     ANSL                 LOWER LIMIT OF ITERATED VARIABLE
C     PBTL                 PAYBACK TIME AT ANSL
C     ANSU                 UPPER LIMIT OF ITERATED VARIABLE
C     PBTU                 PAYBACK TIME AT ANSU
C     PBTM                 PAYBACK TIME AT ANSW, DURING ITERATIONS
C
C
C CALCULATE PAYBACK TIME FOR CASHFLOWS IN CFLO, WITHOUT ADDITION
C  OF ITERATED VARIABLE. THIS IS USED TO DETERMINE WHETHER THE
C  SOLUTION WILL BE NEGATIVE OR POSITIVE
C
      ANSL=0.0
      LABELL=0
      CALL REQPB2(CFLO,PRJLIF,INFL,ESC,ANSL,PBTL,LABELL)
C
C CHECK THAT PBTL IS NOT NEAR ENOUGH TO BE A SOLUTION
      TEST=ABS(RPBT-PBTL)
      IF (TEST.LT.0.001) THEN
        ANSW=ANSL
        LAB=0
        RETURN
      END IF
```

```
C TEST FOR PAYBACK TIME GREATER THAN PROJECT LIFE (LABELL.NE.0),
C  OR PAYBACK TIME GREATER THAN RPBT, IN WHICH CASE POSITIVE CASH
C  FLOWS MUST BE ADDED, IN ORDER TO REDUCE THE PAYBACK TIME.
C  HENCE SET ANSU LARGE POSITIVE NUMBER, AND TEST FOR PBTU < RPBT
      IF ((LABELL.NE.0).OR.(PBTL.GT.RPBT)) THEN
        ANSU=1.E10
        DO 200 I=1,10
          LABELU=0
          CALL REQPB2(CFLO,PRJLIF,INFL,ESC,ANSU,PBTU,LABELU)
C TEST IF ANSU IS A SOLUTION
          TEST=ABS(PBTU-RPBT)
          IF (TEST.LT.0.001) THEN
            ANSW=ANSU
            LAB=0
            RETURN
          END IF
          IF ((PBTU.LT.RPBT).AND.(LABELU.EQ.0)) GOTO 1000
          ANSU=ANSU*2.
200     CONTINUE
        LAB=1
        RETURN
C IN THE OTHER CASE, A NEGATIVE CASH FLOW IS REQUIRED TO BE ADDED
C  TO THE EXISTING CASH FLOWS TO OBTAIN THE REQUIRED PAYBACK TIME.
C  HENCE SET ANSU TO A LARGE NEGATIVE NUMBER, AND TEST FOR
C  PBTU > RPBT.
      ELSE
        ANSU=-1.E10
        DO 300 I=1,10
          LABELU=0
          CALL REQPB2(CFLO,PRJLIF,INFL,ESC,ANSU,PBTU,LABELU)
C TEST IF ANSU IS A SOLUTION
          TEST=ABS(PBTU-RPBT)
          IF (TEST.LT.0.001) THEN
            ANSW=ANSU
            LAB=0
            RETURN
          END IF
          IF ((PBTU.GT.RPBT).OR.(LABELU.NE.0)) GOTO 1000
          ANSU=ANSU*2.
300     CONTINUE
        LAB=2
        RETURN
      END IF
C
C PERFORM ITERATIVE CALCULATION
C
1000  DO 1100 I=1,100
        ANSW=(ANSL+ANSU)/2.
        LABEL=0
        CALL REQPB2(CFLO,PRJLIF,INFL,ESC,ANSW,PBTM,LABEL)
C TEST FOR A SOLUTION
        TEST=ABS(RPBT-PBTM)
        IF (TEST.LT.0.001) THEN
          LAB=0
          RETURN
        END IF
```

```
C SET NEW VALUE OF ANSL OR ANSU, AS APPROPRIATE
        IF (LABEL.NE.0) THEN
          IF ((LABELL.NE.0).OR.(PBTL.GT.RPBT)) THEN
            LABELL=LABEL
            ANSL=ANSW
            PBTL=PBTM
          ELSE
            LABELU=LABEL
            ANSU=ANSW
            PBTU=PBTM
          END IF
        ELSE
          TEST=(RPBT-PBTM)*(RPBT-PBTU)
          IF (TEST.GT.0) THEN
            LABELU=LABEL
            ANSU=ANSW
            PBTU=PBTM
          ELSE
            LABELL=LABEL
            ANSL=ANSW
            PBTL=PBTM
          END IF
        END IF
1100  CONTINUE
      LAB=3
      RETURN
      END
C
C
C
C
      SUBROUTINE REQPB2(CFLO,PRJLIF,INFL,ESC,VAR,ANS,LABEL)
      REAL CFLO(0:PRJLIF),INFL,ESC,VAR,ANS
      REAL CASFLO(0:25)
      INTEGER PRJLIF,LABEL
C SUBROUTINE USED BY REQPBT TO FIND THE VALUE OF A VARIABLE
C  REQUIREDTO GIVE A SPECIFIED PAYBACK TIME
C THIS SUBROUTINE PERFORMS THE CASH FLOW CALCULATION, AND
C  CALCULATES THE PAYBACK TIME BY CALLING SUBROUTINE PAYBAK
C
C VARIABLES:
C     CFLO(0:PRJLIF)      TOTAL YEARLY CASH FLOWS EXCLUDING THE
C                            ITERATED VARIABLE
C     PRJLIF              PROJECT LIFE, YEARS
C     INFL                INFLATION RATE, %
C     ESC                 REAL ESCALATION RATE OF ITERATED
C                           VARIABLE
C     VAR                 VALUE OF THE ITERATED VARIABLE FOR THIS
C                           ITERATION
C     ANS                 PAYBACK TIME FOR THIS VALUE OF VAR
C     LABEL               LABEL TO INDICATE WHETHER CALCULATION
C                           HAS BEEN SUCCESSFUL
C
C LOCAL VARIABLES:
C     CASFLO(0:25)        ARRAY USED TO PASS VALUES OF TOTAL CASH
```

```
C                           FLOWS TO PAYBAK
C     TEMP                TEMPORARY STORAGE
C
      CASFLO(0)=CFLO(0)
      DO 100 IYEAR=1,PRJLIF
        TEMP=((1.+INFL/100.)**IYEAR)*((1.+ESC/100.)**IYEAR)
        CASFLO(IYEAR)=CFLO(IYEAR)+VAR*TEMP
100   CONTINUE
      CALL PAYBAK(CASFLO,PRJLIF,ANS,LABEL)
      RETURN
      END
C
C
C
C
      SUBROUTINE CAPPBT(RPBT,CASFLO,PRJLIF,CAPCOS)
      REAL RPBT,CASFLO(0:PRJLIF),CAPCOS
      INTEGER PRJLIF
C SUBROUTINE TO CALCULATE CAPITAL COST, GIVEN A CASH FLOW TABLE
C  AND A REQUIRED PAYBACK TIME
C
C VARIABLES:
C     RPBT                  REQUIRED PAYBACK TIME
C     CASFLO(0:PRJLIF)      CASH FLOW TABLE
C     PRJLIF                PROJECT LIFE
C     CAPCOS                CALCULATED CAPITAL COST
C
      CAPCOS=0.0
      DO 100 IYEAR=1,RPBT
        CAPCOS=CAPCOS+CASFLO(IYEAR)
100   CONTINUE
      RETURN
      END
C
C
C
      SUBROUTINE FIN1
      CHARACTER*80 ANS
      WRITE(3,*) 'DO YOU WANT TO PRINT THE RESULTS'
      READ(0,90001) ANS
      IF (ANS(1:1).EQ.'Y') THEN
        WRITE(3,*) 'ENTER THE PRINTER NUMBER REQUIRED, 6 OR 7'
        READ(0,*) K

        CALL SPOOL(14H2299JMD*R:EVAL,K,IERR)
        IF (IERR.EQ.0) THEN
          WRITE(3,*) 'R:EVAL HAS BEEN PASSED TO THE PRINTER'
        ELSE
          WRITE(3,*) 'ERROR PASSING R:EVAL TO PRINTER,',
     +      ' ERROR NO.',IERR
        END IF
      END IF
      RETURN
90001 FORMAT(A80)
      END
```

Example Printout:

```
     ********************************************
     PROGRAM  EVAL  -  GASIFIER EVALUATION PROGRAM
     **************** J M Double ****************

RUN NUMBER: EXAMPLE

DATA USED FOR CALCULATION
     (for £ read Pounds Sterling, all tonnes are d.a.f. basis)
     Gasifier throughput                        1.00  tonnes/h
     Feedstock heating.value                   20.00  GJ/tonne
     Gasifier efficiency                       70.00  %
     Number of shifts                           1.0
     Project life                              10     years
     Number of operating hours per year      6000

     Capital cost (inc working capital)    528000.00  £
     Feedstock cost                            20.00  £/tonne
     Product selling price                      2.400 £/GJ
     Utilities cost                             0.300 £/GJ
     Yearly maintenance cost, fraction capcost  0.025
     Yearly overheads, fraction capcost         0.080
     Total cost of labour per shift         40000.00  £/year
     Target real discount rate                 10.00  %

     Nominal cost of capital                   12.00  %
     Inflation rate                             5.00  %
     Feedstock cost real escalation rate        0.00  %
     Product price real escalation rate         0.00  %
     Utilities cost real escalation rate        0.00  %
     Maintenance cost real escalation rate      0.00  %
     Overheads real escalation rate             0.00  %
     Labour cost real escalation rate           0.00  %

RESULTS
     NPV at  10.00 % real discount rate is £     -767883.90
     DCF rate of return is less than -90.00 % real
     Payback time is greater than project life
     Mean real product cost over project life is  3.724 £/GJ

     FOR A  3 YEAR PAYBACK TIME:
     Either FEEDSTOCK COST must be  -13.09 £/tonne
         or PRODUCT SELLING PRICE must be  4.76 £/GJ
         or CAPITAL COST must be £    -129227.28

     FOR A  5 YEAR PAYBACK TIME:
     Either FEEDSTOCK COST must be   -1.68 £/tonne
         or PRODUCT SELLING PRICE must be  3.95 £/GJ
         or CAPITAL COST must be £    -226506.68

     FOR A 10 YEAR PAYBACK TIME:
     Either FEEDSTOCK COST must be    6.83 £/tonne
         or PRODUCT SELLING PRICE must be  3.34 £/GJ
         or CAPITAL COST must be £    -515592.97
```

```
RUN NUMBER: EXAMPLE

NOMINAL CASH FLOW TABLE FOR N.P.V. CALCULATIONS (all cash flows expressed in Pounds Sterling)
 Year      Capital     Product   Feedstock   Utilities Maintenance   Overheads      Labour   TOTAL NOMINAL    TOTAL REAL
    0   -528000.00        0.00        0.00        0.00        0.00        0.00        0.00      -528000.00    -528000.00
    1         0.00   211680.00  -126000.00   -26460.00   -13860.00   -44352.00   -42000.00       -40992.00     -39040.00
    2         0.00   222264.00  -132300.00   -27783.00   -14553.00   -46569.60   -44100.00       -43041.60     -39040.00
    3         0.00   233377.20  -138915.00   -29172.15   -15280.65   -48898.08   -46305.00       -45193.68     -39040.00
    4         0.00   245046.06  -145860.75   -30630.76   -16044.68   -51342.98   -48620.25       -47453.36     -39040.00
    5         0.00   257298.36  -153153.79   -32162.30   -16846.92   -53910.13   -51051.26       -49826.03     -39040.00
    6         0.00   270163.28  -160811.48   -33770.41   -17689.26   -56605.64   -53603.83       -52317.33     -39040.00
    7         0.00   283671.45  -168852.05   -35458.93   -18573.73   -59435.92   -56284.02       -54933.20     -39040.00
    8         0.00   297855.02  -177294.65   -37231.88   -19502.41   -62407.72   -59098.22       -57679.86     -39040.00
    9         0.00   312747.77  -186159.39   -39093.47   -20477.53   -65528.10   -62053.13       -60563.85     -39040.00
   10         0.00   328385.16  -195467.36   -41048.14   -21501.41   -68804.51   -65155.79       -63592.05     -39040.00

NOMINAL CASH FLOW TABLE FOR PRODUCT COST CALCULATIONS (all cash flows expressed in Pounds Sterling)
 Year   Amortisation    Other costs    Total costs   Product cost £/GJ
                        (see above)                   Nominal    Real
    1       93447.64      252672.00      346119.64      4.120   3.924
    2       93447.64      265305.60      358753.24      4.271   3.874
    3       93447.64      278570.88      372018.52      4.429   3.826
    4       93447.64      292499.42      385947.06      4.595   3.780
    5       93447.64      307124.40      400572.03      4.769   3.736
    6       93447.64      322480.61      415928.25      4.952   3.695
    7       93447.64      338604.65      432052.28      5.143   3.655
    8       93447.64      355534.88      448982.52      5.345   3.618
    9       93447.64      373311.62      466759.26      5.557   3.582
   10       93447.64      391977.20      485424.84      5.779   3.548
```

Program EVAL2

The basic flow diagram of this program is shown as Figure B4. A complete list of subroutines is given below:

Subroutines

FINISH	Subroutine to close the output files and terminate the program.
FIN1	Subroutine to spool the results file to the printer, if required.
INITAL	Subroutine to read the base case data from the data file.
INPUT	Subroutine for interactive data entry.

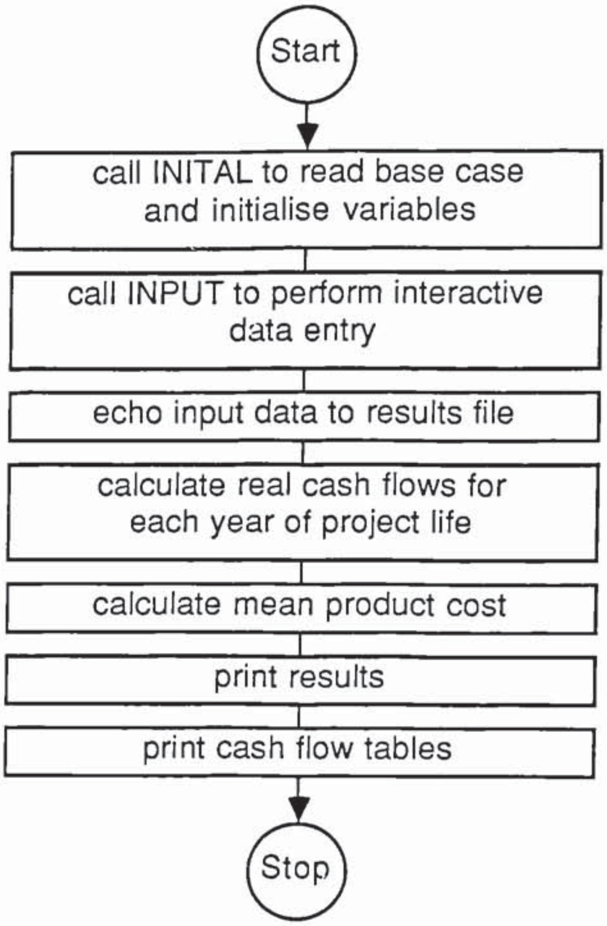

Figure B4 Flow Diagram of Program EVAL2

```
      PROGRAM EVAL2
C A PROGRAM TO PERFORM FINANCIAL CALCULATIONS FOR GASIFICATION
C  SYSTEMS
C ALL CALCULATIONS ARE BEFORE TAX, IN REAL TERMS
C
C DECLARE GLOBAL VARIABLES
      REAL CAPCTY,RAWHHV,EFF
      INTEGER PRJLIF,HRPYR
      REAL CAPCOS,RAWMAT,UTILS,MAINTC,OVERHD,MANCOS,NOMINT
      REAL INFL,ESCRAW,ESCUT,ESCMTC,ESCOVH,ESCLAB,NUMMEN
      COMMON /INDATA/ CAPCTY,RAWHHV,EFF,NUMMEN,PRJLIF,HRPYR,
     +CAPCOS,RAWMAT,UTILS,MAINTC,OVERHD,MANCOS,NOMINT,INFL,ESCRAW,
     +ESCUT,ESCMTC,ESCOVH,ESCLAB
C
C GLOBAL VARIABLES ARE AS FOLLOWS:
C     CAPCTY             CAPACITY OF GASIFIER T/H
C     RAWHHV             HHV OF RAW MATERIAL, MJ/KG
C     EFF                EFFICIENCY OF GASIFIER %
C                        (DEFINED AS HEATING VALUE OF TOTAL PRODUCT
C                        DIVIDED BY HEATING VALUE OF FEEDSTOCK)
C     PRJLIF             PROJECT LIFE, YEARS
C     HRPYR              NUMBER OF HOURS OF OPERATION EACH YEAR
C     NUMMEN             NUMBER OF MEN REQUIRED TO OPERATE PLANT
C     CAPCOS             CAPITAL COST OF GASIFIER
C     RAWMAT             COST OF RAW MATERIAL, £/T
C     UTILS              UTILITIES COST, £/GJ PRODUCT GAS
C     MAINTC             MAINTENANCE COST/YEAR, £/£CAPITAL COST
C     OVERHD             OVERHEADS/YEAR, £/£CAPITAL COST
C     MANCOS             TOTAL COST OF EMPLOYING ONE MAN, £/YEAR
C     NOMINT             NOMINAL COST OF CAPITAL
C     INFL               RATE OF INFLATION
C     ESCRAW             REAL ESCALATION RATE OF FEEDSTOCK
C     ESCUT              REAL ESCALATION RATE OF UTILITIES
C     ESCMTC             REAL ESCALATION RATE OF MAINTENANCE
C     ESCOVH             REAL ESCALATION RATE OF OVERHEADS
C     ESCLAB             REAL ESCALATION RATE OF LABOUR COST
C
C DECLARE LOCAL VARIABLES
      REAL AMTTOT,RAWTOT,UTTOT,OVHTOT,LABTOT,TOTTOT,PRCTOT
      REAL AVPRC,AMTPC,RAWPC,UTPC,OVHPC,LABPC,MTCPC,MTCTOT
      REAL UTILYR,LABYR,OVHDYR,MAINYR,RAWYR,GJOUT,TEMP,AMORT
      REAL AMTCOS(25),RAWCOS(25),UTCOS(25),OVHCOS(25),LABCOS(25)
      REAL TOTCOS(25),RELPRC(25),MTCCOS(25)

      REAL PRCKWE,TOTGJ
      CHARACTER*80 RUNNO,AGAIN
C
C LOCAL VARIABLES ARE AS FOLLOWS:
C     UTILYR             UTILITY COST PER YEAR, REAL TERMS, NO ESC.
C     LABYR              LABOUR COST PER YEAR, AS UTILYR
C     OVHDYR             OVERHEADS PER YEAR, AS UTILYR
C     MAINYR             MAINTENANCE PER YEAR, AS UTILYR
C     RAWYR              RAW MATERIAL COST PER YEAR, AS UTILYR
C     GJOUT              CAPACITY OF GASIFIER, GJ PRODUCT GAS/HOUR
C     TEMP               TEMPORARY VARIABLE IN AMORTISATION CALC.
```

```
C      AMORT              YEARLY AMORTISATION, NOMINAL TERMS
C      AMTCOS(25)         REAL AMORTISATION YEARLY COSTS
C      RAWCOS(25)         REAL FEEDSTOCK YEARLY COSTS
C      UTCOS(25)          REAL UTILITIES YEARLY COSTS
C      MTCCOS(25)         REAL MAINTENANCE YEARLY COSTS
C      OVHCOS(25)         REAL OVERHEADS YEARLY COSTS
C      LABCOS(25)         REAL LABOUR YEARLY COSTS
C      TOTCOS(25)         REAL YEARLY COSTS, INCLUDING AMORTISATION
C      RELPRC(25)         REAL PRODUCT COST
C      AMTTOT             RUNNING TOTAL OF REAL AMORTISATION COSTS,
C                           TO CALC MEAN
C      RAWTOT             RUNNING TOTAL OF REAL FEEDSTOCK COSTS, TO
C                           CALC MEAN
C      UTTOT              RUNNING TOTAL OF REAL UTILITIES COSTS, TO
C                           CALC MEAN
C      MTCTOT             RUNNING TOTAL OF REAL MAINTENANCE COSTS,
C                           TO CALC MEAN
C      OVHTOT             RUNNING TOTAL OF REAL OVERHEADS COSTS TO
C                           CALC MEAN
C      LABTOT             RUNNING TOTAL OF REAL LABOUR COSTS TO
C                           CALC MEAN
C      TOTTOT             RUNNING TOTAL OF REAL TOT COSTS TO
C                           CALC MEAN
C      PRCTOT             RUNNING TOTAL OF REAL PRODUCT COSTS
C                           TO CALC MEAN
C      AVPRC              MEAN REAL PRODUCT COST, £/GJ
C      AMTPC              AMORTISATION £/GJ
C      RAWPC              FEEDSTOCK COST £/GJ
C      UTPC               UTILITIES COST £/GJ
C      MTCPC              MAINTENANCE COST £/GJ
C      OVHPC              OVERHEADS COST £/GJ
C      LABPC              LABOUR COST £/GJ
C      PRCKWE             COST/kWhe ASSUMING COSTS ARE FOR WHOLE
C                          SYSTEM, INCLUDING GENERATOR SET
C      TOTGJ              TOTAL GJ OF PRODUCT PRODUCED IN LIFE OF
C                          PROJECT
C      RUNNO              CHARACTER VARIABLE USED TO IDENTIFY RUN
C      AGAIN              CHARACTER VARIABLE
C
       OPEN(UNIT=20,FILE='2299JMD*R:EVAL',STATUS='OLD')
       CALL INITAL
10     CALL INPUT
       WRITE(3,*) ' '
       WRITE(3,*) 'ENTER A RUN NUMBER TO IDENTIFY THIS RUN'
       READ(0,89999) RUNNO
       GJOUT=EFF*RAWHHV*CAPCTY/100.0
       UTILYR=GJOUT*HRPYR*UTILS
       MAINYR=MAINTC*CAPCOS
       OVHYR=OVERHD*CAPCOS
       RAWYR=RAWMAT*CAPCTY*HRPYR
       LABYR=NUMMEN*MANCOS
C ECHO DATA TO RESULTS FILE BEFORE PERFORMING CALCULATION
       WRITE(20,90000) RUNNO
       WRITE(20,90001) CAPCTY
       WRITE(20,90002) RAWHHV
```

```
      WRITE(20,90003) EFF
      WRITE(20,90004) NUMMEN
      WRITE(20,90005) PRJLIF
      WRITE(20,90006) HRPYR
      WRITE(20,90009) CAPCOS
      WRITE(20,90010) RAWMAT
      WRITE(20,90012) UTILS
      WRITE(20,90013) MAINTC
      WRITE(20,90014) OVERHD
      WRITE(20,90015) MANCOS
      WRITE(20,90017) NOMINT
      WRITE(20,90018) INFL
      WRITE(20,90019) ESCRAW
      WRITE(20,90021) ESCUT
      WRITE(20,90022) ESCMTC
      WRITE(20,90023) ESCOVH
      WRITE(20,90024) ESCLAB
C
C PERFORM PRODUCT COST CALCULATIONS
C
C  CALCULATE YEARLY AMORTISATION
      TEMP=(1.+NOMINT/100.)**PRJLIF
      AMORT=CAPCOS*((NOMINT/100.)*TEMP)/(TEMP-1.)
C  CALCULATE YEARLY COSTS, AND HENCE PRODUCT COSTS FOR EACH YEAR
      PRCTOT=0.0
      TOTTOT=0.0
      RAWTOT=0.0
      UTTOT=0.0
      MTCTOT=0.0
      OVHTOT=0.0
      LABTOT=0.0
      AMTTOT=0.0
      DO 100 I=1,PRJLIF
        AMTCOS(I)=AMORT/((1.0+INFL/100.)**I)
        RAWCOS(I)=RAWYR*(1.0+ESCRAW/100.0)**I
        UTCOS(I)=UTILYR*(1.0+ESCUT/100.0)**I
        MTCCOS(I)=MAINYR*(1.0+ESCMTC/100.0)**I
        OVHCOS(I)=OVHYR*(1.0+ESCOVH/100.0)**I
        LABCOS(I)=LABYR*(1.0+ESCLAB/100.0)**I
        TOTCOS(I)=AMTCOS(I)+RAWCOS(I)+UTCOS(I)+MTCCOS(I)+
     +      OVHCOS(I)+LABCOS(I)

        AMTTOT=AMTTOT+AMTCOS(I)
        RAWTOT=RAWTOT+RAWCOS(I)
        UTTOT=UTTOT+UTCOS(I)
        MTCTOT=MTCTOT+MTCCOS(I)
        OVHTOT=OVHTOT+OVHCOS(I)
        LABTOT=LABTOT+LABCOS(I)
        TOTTOT=TOTTOT+TOTCOS(I)
        RELPRC(I)=TOTCOS(I)/(HRPYR*GJOUT)
        PRCTOT=PRCTOT+RELPRC(I)
100   CONTINUE
C  CALCULATE MEAN PRODUCT COST
      AVPRC=PRCTOT/(PRJLIF*1.0)
      TOTGJ=GJOUT*HRPYR*PRJLIF
      LABPC=LABTOT/TOTGJ
```

```
      RAWPC=RAWTOT/TOTGJ
      UTPC=UTTOT/TOTGJ
      MTCPC=MTCTOT/TOTGJ
      OVHPC=OVHTOT/TOTGJ
      AMTPC=AMTTOT/TOTGJ
      PRCKWE=TOTTOT/(1000.*CAPCTY*HRPYR*PRJLIF)
C
C
C PRINT RESULTS
      DO 200 IOUT=3,20,17
        WRITE(IOUT,99000) AVPRC
        WRITE(IOUT,99001)
        WRITE(IOUT,99002) AMTPC
        WRITE(IOUT,99003) RAWPC
        WRITE(IOUT,99004) UTPC
        WRITE(IOUT,99005) MTCPC
        WRITE(IOUT,99006) OVHPC
        WRITE(IOUT,99007) LABPC
        WRITE(IOUT,99008) MTCPC+OVHPC
        WRITE(IOUT,99009) RAWPC+UTPC+MTCPC+OVHPC+LABPC
        WRITE(IOUT,99010) PRCKWE*100.0
200   CONTINUE
      WRITE(20,99900) RUNNO
      DO 300 I=1,PRJLIF
        WRITE(20,99901) I,AMTCOS(I),RAWCOS(I),UTCOS(I),MTCCOS(I),
     +    OVHCOS(I),LABCOS(I),TOTCOS(I),RELPRC(I)
300   CONTINUE
C
      WRITE(3,*) 'PRESS <RETURN> TO CONTINUE'
      READ(0,89999) AGAIN
      GOTO 10
C
C
89999 FORMAT(A80)
90000 FORMAT('1',14X,45('*'),/,15X,'PROGRAM  EVAL2 -  ',
     +'GASIFIER EVALUATION PROGRAM',/,15X,17('*'),
     +' J M Double ',16('*'),/,'0',9X,'RUN NUMBER: ',A80,/,
     +'0',9X,'DATA USED FOR CALCULATION ',
     +/,15X,'(for £ read Pounds Sterling,',
     +' all tonnes are d.a.f. basis)')
90001 FORMAT('0',14X,'Gasifier throughput',23X,F5.2,'  tonnes/h')
90002 FORMAT(15X,'Feedstock heating value',19X,F5.2,'  GJ/tonne')
90003 FORMAT(15X,'Gasifier efficiency',23X,F5.2,'  %')
90004 FORMAT(15X,'Number of shifts                        ',6X,F4.1)
90005 FORMAT(15X,'Project life',30X,I2,5X,'years')
90006 FORMAT(15X,'Hours of operation per year',13X,I4)
90009 FORMAT('0',14X,'Capital cost (inc working capital) ',
     +F12.2,'  £')
90010 FORMAT(15X,'Feedstock cost',28X,F5.2,'  £/tonne')
90011 FORMAT(15X,'Product selling price',21X,F6.3,' £/GJ')
90012 FORMAT(15X,'Utilities cost',28X,F6.3,' £/GJ')
90013 FORMAT(15X,'Yearly maintenance cost,',
     +' fraction capcost  ',F5.3)
90014 FORMAT(15X,'Yearly overheads, fraction capcost',9X,F5.3)
90015 FORMAT(15X,'Total cost of labour per shift',8X,F9.2,
```

```
     +'  £/year')
90017 FORMAT('0',14X,'Nominal cost of capital',19X,F5.2,'  %')
90018 FORMAT(15X,'Inflation rate         ',19X,F5.2,'  %')
90019 FORMAT(15X,'Feedstock cost real escalation rate',7X,
     +F5.2,'  %')
90020 FORMAT(15X,'Product price real escalation rate',8X,
     +F5.2,'  %')
90021 FORMAT(15X,'Utilities cost real escalation rate',7X,
     +F5.2,'  %')
90022 FORMAT(15X,'Maintenance cost real escalation rate',5X,
     +F5.2,'  %')
90023 FORMAT(15X,'Overheads real escalation rate',12X,F5.2,'  %')
90024 FORMAT(15X,'Labour cost real escalation rate',10X,
     +F5.2,'  %')
99000 FORMAT(/,'0',9X,'RESULTS',/,
     +'0',14X,'Mean real product cost is ',F7.3,' £/GJ')
99001 FORMAT('0',14X,'BREAKDOWN OF PRODUCT COST',
     +' (averaged over project life)')
99002 FORMAT(20X,'Capital costs                 ',F6.3,' £/GJ')
99003 FORMAT(20X,'Feedstock                     ',F6.3,' £/GJ')
99004 FORMAT(20X,'Utilities                     ',F6.3,' £/GJ')
99005 FORMAT(20X,'Maintenance                   ',F6.3,' £/GJ')
99006 FORMAT(20X,'Overheads                     ',F6.3,' £/GJ')
99007 FORMAT(20X,'Labour                        ',F6.3,' £/GJ')
99008 FORMAT('0',19X,'Overheads + maintenance       ',F6.3,' £/GJ')
99009 FORMAT('0',19X,'Subtotal, variable costs      ',F6.3,' £/GJ')
99010 FORMAT(/,'0',14X,'Electricity cost ',F6.2,' p/kWh')
99900 FORMAT('1',/,'0    RUN NUMBER: ',A80,/,'0    ',
     +'TABLE OF YEARLY COSTS, AND PRODUCT COST EACH YEAR. ',
     +'All figures are in Pounds Sterling, real terms',/,
     +'0    Year',4X,'Amortisation',5X,'Feedstock',5X,'Utilities',
     +3X,'Maintenance',5X,'Overheads',8X,'Labour',6X,
     +'TOTAL COSTS',5X,'PRODUCT COST',/,121X,'£/GJ')
99901 FORMAT(6X,I2,3X,6(2X,F12.2),4X,F14.2,7X,F6.3)
      END
C
C

      SUBROUTINE INITAL
C SUBROUTINE TO READ IN THE BASE CASE VALUES FROM FILE
C  2299JMD*D:EVAL2
C DECLARE GLOBAL VARIABLES
      REAL CAPCTY,RAWHHV,EFF
      INTEGER PRJLIF,HRPYR
      REAL CAPCOS,RAWMAT,UTILS,MAINTC,OVERHD,MANCOS,NOMINT
      REAL INFL,ESCRAW,ESCUT,ESCMTC,ESCOVH,ESCLAB,NUMMEN
      COMMON /INDATA/ CAPCTY,RAWHHV,EFF,NUMMEN,PRJLIF,HRPYR,
     +CAPCOS,RAWMAT,UTILS,MAINTC,OVERHD,MANCOS,NOMINT,INFL,ESCRAW,
     +ESCUT,ESCMTC,ESCOVH,ESCLAB
      OPEN(UNIT=21,FILE='2299JMD*D:EVAL2',STATUS='OLD')
      READ(21,*) CAPCTY
      READ(21,*) RAWHHV
      READ(21,*) EFF
      READ(21,*) PRJLIF
      READ(21,*) HRPYR
```

```
      READ(21,*) NUMMEN
      READ(21,*) CAPCOS
      READ(21,*) RAWMAT
      READ(21,*) UTILS
      READ(21,*) MAINTC
      READ(21,*) OVERHD
      READ(21,*) MANCOS
      READ(21,*) NOMINT
      READ(21,*) INFL
      READ(21,*) ESCRAW
      READ(21,*) ESCUT
      READ(21,*) ESCMTC
      READ(21,*) ESCOVH
      READ(21,*) ESCLAB
      CLOSE 21
      RETURN
      END
C
C
C
      SUBROUTINE FINISH
      WRITE(20,90000)
      CLOSE 20
      CALL FIN1
      STOP
90000 FORMAT('1')
      END
C
C
C
C
      SUBROUTINE INPUT
C SUBROUTINE TO ENABLE DATA FOR RUN TO BE ENTERED INTERACTIVELY
C  BY CHANGING VALUES OF VARIABLES FROM THE BASE CASE
C
C DECLARE GLOBAL VARIABLES
      REAL CAPCTY,RAWHHV,EFF
      INTEGER PRJLIF,HRPYR
      REAL CAPCOS,RAWMAT,UTILS,MAINTC,OVERHD,MANCOS,NOMINT
      REAL INFL,ESCRAW,ESCUT,ESCMTC,ESCOVH,ESCLAB,NUMMEN
      COMMON /INDATA/
CAPCTY,RAWHHV,EFF,NUMMEN,PRJLIF,HRPYR,CAPCOS,
     +RAWMAT,UTILS,MAINTC,OVERHD,MANCOS,NOMINT,INFL,ESCRAW,
     +ESCUT,ESCMTC,ESCOVH,ESCLAB
C
10      DO 11 I=1,7
          WRITE(3,*) ' '
11      CONTINUE
        WRITE(3,*) '** PROGRAM EVAL2 ',
     +'- GASIFIER ECONOMIC CALCULATIONS **'
        WRITE(3,90000)
        WRITE(3,90001) CAPCTY,PRJLIF
        WRITE(3,90002) RAWHHV,HRPYR
        WRITE(3,90003) EFF,NUMMEN
        WRITE(3,90005)
        WRITE(3,90006) CAPCOS
```

```
      WRITE(3,90007) RAWMAT,ESCRAW
      WRITE(3,90009) UTILS,ESCUT
      WRITE(3,90010) MAINTC,ESCMTC
      WRITE(3,90011) OVERHD,ESCOVH
      WRITE(3,90012) MANCOS,ESCLAB
      WRITE(3,90014) NOMINT
      WRITE(3,90015) INFL
      WRITE(3,90016)
      WRITE(3,90017)
      WRITE(3,90018)
      WRITE(3,90019)
      WRITE(3,90020)
      READ(0,*) ICHANG
      IF (ICHANG.EQ.0) CALL FINISH
      IF (ICHANG.EQ.98) GOTO 9800
      IF (ICHANG.EQ.99) GOTO 9900
      IF (ICHANG.EQ.100) RETURN
      GOTO(100,200,300,400,500,600,700,800,900,1000,1100,
     +1200,1300,1400,1500,1600,1700,1800,1900) ICHANG
      WRITE(3,*) 'INCORRECT INPUT, TRY AGAIN'
      GOTO 10
100   WRITE(3,*) 'THE CAPACITY OF THE GASIFIER IS ',CAPCTY,' t/h'
      WRITE(3,*) 'ENTER THE NEW CAPACITY'
      READ(0,*) CAPCTY
      GOTO 10
200   WRITE(3,*) 'THE FEEDSTOCK HEATING VALUE IS ',RAWHHV,' MJ/kg'
      WRITE(3,*) 'ENTER THE NEW HEATING VALUE'
      READ(0,*) RAWHHV
      GOTO 10
300   WRITE(3,*) 'THE GASIFIER EFFICIENCY IS ',EFF,' %'
      WRITE(3,*) 'ENTER THE NEW EFFICIENCY'
      READ(0,*) EFF
      GOTO 10
400   WRITE(3,*) 'THE PROJECT LIFE IS ',PRJLIF,' years'
      WRITE(3,*) 'ENTER THE NEW PROJECT LIFE'
      READ(0,*) PRJLIF
      GOTO 10
500   WRITE(3,*) HRPYR,' HOURS ARE WORKED EACH YEAR'
      WRITE(3,*) 'ENTER THE NEW HOURS/YEAR'
      READ(0,*) HRPYR
      GOTO 10
600   WRITE(3,*) 'THERE ARE ',NUMMEN,' SHIFTS'
      WRITE(3,*) 'ENTER THE NEW NUMBER OF SHIFTS'
      READ(0,*) NUMMEN
      GOTO 10
700   WRITE(3,*) 'CAPITAL COST IS CURRENTLY £',CAPCOS
      WRITE(3,*) 'ENTER THE NEW CAPITAL COST'
      READ(0,*) CAPCOS
      GOTO 10
800   WRITE(3,*) 'THE FEEDSTOCK COST IS ',RAWMAT,' £/tonne'
      WRITE(3,*) 'ENTER THE NEW FEEDSTOCK COST'
      READ(0,*) RAWMAT
      GOTO 10
900   WRITE(3,*) 'THE UTILITIES COST IS ',UTILS,' £/GJ PRODUCT'
      WRITE(3,*) 'ENTER THE NEW UTILITY COST'
```

```
      READ(0,*) UTILS
      GOTO 10
 1000 WRITE(3,*) 'THE MAINTENANCE COST IS ',MAINTC,' FRACTION',
     +' OF CAPITAL COST'
      WRITE(3,*) 'ENTER THE NEW MAINTENANCE COST'
      READ(0,*) MAINTC
      GOTO 10
 1100 WRITE(3,*) 'THE OVERHEADS ARE ',OVERHD,' FRACTION OF ',
     +'CAPITAL COST'
      WRITE(3,*) 'ENTER THE NEW OVERHEADS'
      READ(0,*) OVERHD
      GOTO 10
 1200 WRITE(3,*) 'THE COST PER SHIFT IS ',MANCOS,' £/year'
      WRITE(3,*) 'ENTER THE NEW COST PER MAN'
      READ(0,*) MANCOS
      GOTO·10
 1300 WRITE(3,*) 'THE NOMINAL COST OF CAPITAL IS ',NOMINT,' %'
      WRITE(3,*) 'ENTER THE NEW COST OF CAPITAL'
      READ(0,*) NOMINT
      GOTO 10
 1400 WRITE(3,*) 'THE INFLATION RATE IS ',INFL,' %'
      WRITE(3,*) 'ENTER THE NEW INFLATION RATE'
      READ(0,*) INFL
      GOTO 10
 1500 WRITE(3,*) 'THE FEEDSTOCK REAL ESCALATION IS ',ESCRAW,' %'
      WRITE(3,*) 'ENTER THE NEW FEEDSTOCK ESCALATION'
      READ(0,*) ESCRAW
      GOTO 10
 1600 WRITE(3,*) 'THE UTILITIES REAL ESCALATION IS ',ESCUT,' %'
      WRITE(3,*) 'ENTER THE NEW UTILITIES ESCALATION'
      READ(0,*) ESCUT
      GOTO 10
 1700 WRITE(3,*) 'THE MAINTENANCE REAL ESCALATION IS ',ESCMTC,' %'
      WRITE(3,*) 'ENTER THE NEW MAINTENANCE ESCALATION'
      READ(0,*) ESCMTC
      GOTO 10
 1800 WRITE(3,*) 'THE OVERHEADS REAL ESCALATION IS ',ESCOVH,' %'
      WRITE(3,*) 'ENTER THE NEW OVERHEADS ESCALATION'
      READ(0,*) ESCOVH
      GOTO 10
 1900 WRITE(3,*) 'THE LABOUR COST REAL ESCALATION IS ',ESCLAB,' %'
      WRITE(3,*) 'ENTER THE NEW LABOUR ESCALATION'
      READ(0,*) ESCLAB
      GOTO 10
 9800 CALL INITAL
      GOTO 10
 9900 OPEN(UNIT=21,FILE='2299JMD*D:EVAL2',STATUS='OLD')
      WRITE(21,*) CAPCTY
      WRITE(21,*) RAWHHV
      WRITE(21,*) EFF
      WRITE(21,*) PRJLIF
      WRITE(21,*) HRPYR
      WRITE(21,*) NUMMEN
      WRITE(21,*) CAPCOS
      WRITE(21,*) RAWMAT
```

```
      WRITE(21,*) UTILS
      WRITE(21,*) MAINTC
      WRITE(21,*) OVERHD
      WRITE(21,*) MANCOS
      WRITE(21,*) NOMINT
      WRITE(21,*) INFL
      WRITE(21,*) ESCRAW
      WRITE(21,*) ESCUT
      WRITE(21,*) ESCMTC
      WRITE(21,*) ESCOVH
      WRITE(21,*) ESCLAB
      CLOSE 21
      GOTO 10
90000 FORMAT('  ENTER THE NUMBER OF THE VARIABLE TO BE CHANGED',
     +' OR ACTION REQUIRED:')
90001 FORMAT('   1. Gasifier throughput',9X,F6.2,' t/h    ',
     +'4. Project life, years',4X,I3)
90002 FORMAT('   2. HHV of feedstock',12X,F6.2,' GJ/t   ',
     +'5. Working hours / year  ',I4)
90003 FORMAT('   3. Gasifier efficiency',9X,F5.1,'  %',6X,
     +'6. Number of shifts        ',F4.1)
90005 FORMAT('  FINANCIAL DATA:')
90006 FORMAT('   7. Capital cost',10X,F12.2,' £')
90007 FORMAT('   8. Feedstock cost',14X,F6.2,' £/t   ',
     +'15. Feedstock escalation',4X,F5.2,'%')
90009 FORMAT('   9. Utilities',20X,F6.3,'£/GJ  ',
     +'16. Utilities escalation',4X,F5.2,'%')
90010 FORMAT('  10. Maintenance (frac cap cost)   ',F5.3,6X,
     +'17. Maintenance escalation  ',F5.2,'%')
90011 FORMAT('  11. Overheads (frac cap cost)',5X,F5.3,6X,
     +'18. Overheads escalation',4X,F5.2,'%')
90012 FORMAT('  12. Cost per shift    ',7X,F9.2,' £/yr  19.',
     +' Labour escalation',7X,F5.2,'%')
90014 FORMAT('  13. Nominal cost of capital',5X,F6.2,' %')
90015 FORMAT('  14. Inflation rate',14X,F6.2,' %')
90016 FORMAT('0  0. STOP THE PROGRAM')
90017 FORMAT('  98. GO BACK TO THE BASE CASE')
90018 FORMAT('  99. STORE THESE FIGURES AS THE NEW BASE CASE')
90019 FORMAT(' 100. START THE CALCULATION')
90020 FORMAT(' ** ENTER THE NUMBER OF THE ACTION REQUIRED **')
      END
C
C
C
      SUBROUTINE FIN1
      CHARACTER*80 ANS
      WRITE(3,*) 'DO YOU WANT TO PRINT THE RESULTS'
      READ(0,90001) ANS
      IF (ANS(1:1).EQ.'Y') THEN
        WRITE(3,*) 'ENTER THE PRINTER NUMBER REQUIRED, 6 OR 7'
        READ(0,*) K
        CALL SPOOL(14H2299JMD*R:EVAL,K,IERR)
        IF (IERR.EQ.0) THEN
          WRITE(3,*) 'R:EVAL HAS BEEN PASSED TO THE PRINTER'
        ELSE
          WRITE(3,*) 'ERROR PASSING R:EVAL TO PRINTER,',
```

```
     +    ' ERROR NO.',IERR
          END IF
        END IF
        RETURN
90001 FORMAT(A80)
        END
```

Example Printout:

```
     *******************************************
     PROGRAM  EVAL2 -  GASIFIER EVALUATION PROGRAM
     **************** J M Double ***************

RUN NUMBER: EXAMPLE

DATA USED FOR CALCULATION
     (for £ read Pounds Sterling, all tonnes are d.a.f. basis)

     Gasifier throughput                           1.00  tonnes/h
     Feedstock heating value                      20.00  GJ/tonne
     Gasifier efficiency                          70.00  %
     Number of shifts                              1.0
     Project life                                 10     years
     Hours of operation per year                6000

     Capital cost (inc working capital)     352000.00  £
     Feedstock cost                               20.00  £/tonne
     Utilities cost                                0.200 £/GJ
     Yearly maintenance cost, fraction capcost     0.025
     Yearly overheads, fraction capcost            0.080
     Total cost of labour per shift           25000.00  £/year

     Nominal cost of capital                      12.00  %
     Inflation rate                                5.00  %
     Feedstock cost real escalation rate           0.00  %
     Utilities cost real escalation rate           0.00  %
     Maintenance cost real escalation rate         0.00  %
     Overheads real escalation rate                0.00  %
     Labour cost real escalation rate              0.00  %


RESULTS

     Mean real product cost is    2.939 £/GJ

     BREAKDOWN OF PRODUCT COST (averaged over project life)
          Capital costs                   0.573 £/GJ
          Feedstock                       1.429 £/GJ
          Utilities                       0.200 £/GJ
          Maintenance                     0.105 £/GJ
          Overheads                       0.335 £/GJ
          Labour                          0.298 £/GJ
          Overheads + maintenance         0.440 £/GJ
          Subtotal, variable costs        2.366 £/GJ

     Electricity cost    4.11 p/kWh
```

```
RUN NUMBER: EXAMPLE.
TABLE OF YEARLY COSTS, AND PRODUCT COST EACH YEAR. All figures are in Pounds Sterling, real terms

Year    Amortisation    Feedstock    Utilities   Maintenance    Overheads     Labour     TOTAL COSTS   PRODUCT COST
                                                                                                           £/GJ

  1         59331.83    120000.00     16800.00       8800.00     28160.00    25000.00     258091.83       3.073
  2         56506.51    120000.00     16800.00       8800.00     28160.00    25000.00     255266.51       3.039
  3         53815.72    120000.00     16800.00       8800.00     28160.00    25000.00     252575.72       3.007
  4         51253.07    120000.00     16800.00       8800.00     28160.00    25000.00     250013.07       2.976
  5         48812.45    120000.00     16800.00       8800.00     28160.00    25000.00     247572.45       2.947
  6         46488.04    120000.00     16800.00       8800.00     28160.00    25000.00     245248.04       2.920
  7         44274.33    120000.00     16800.00       8800.00     28160.00    25000.00     243034.33       2.893
  8         42166.03    120000.00     16800.00       8800.00     28160.00    25000.00     240926.03       2.868
  9         40158.12    120000.00     16800.00       8800.00     28160.00    25000.00     238918.12       2.844
 10         38245.83    120000.00     16800.00       8800.00     28160.00    25000.00     237005.83       2.821
```

APPENDIX C

FURTHER RESULTS OF THE CARBON BOUNDARY MODEL SENSITIVITY ANALYSIS

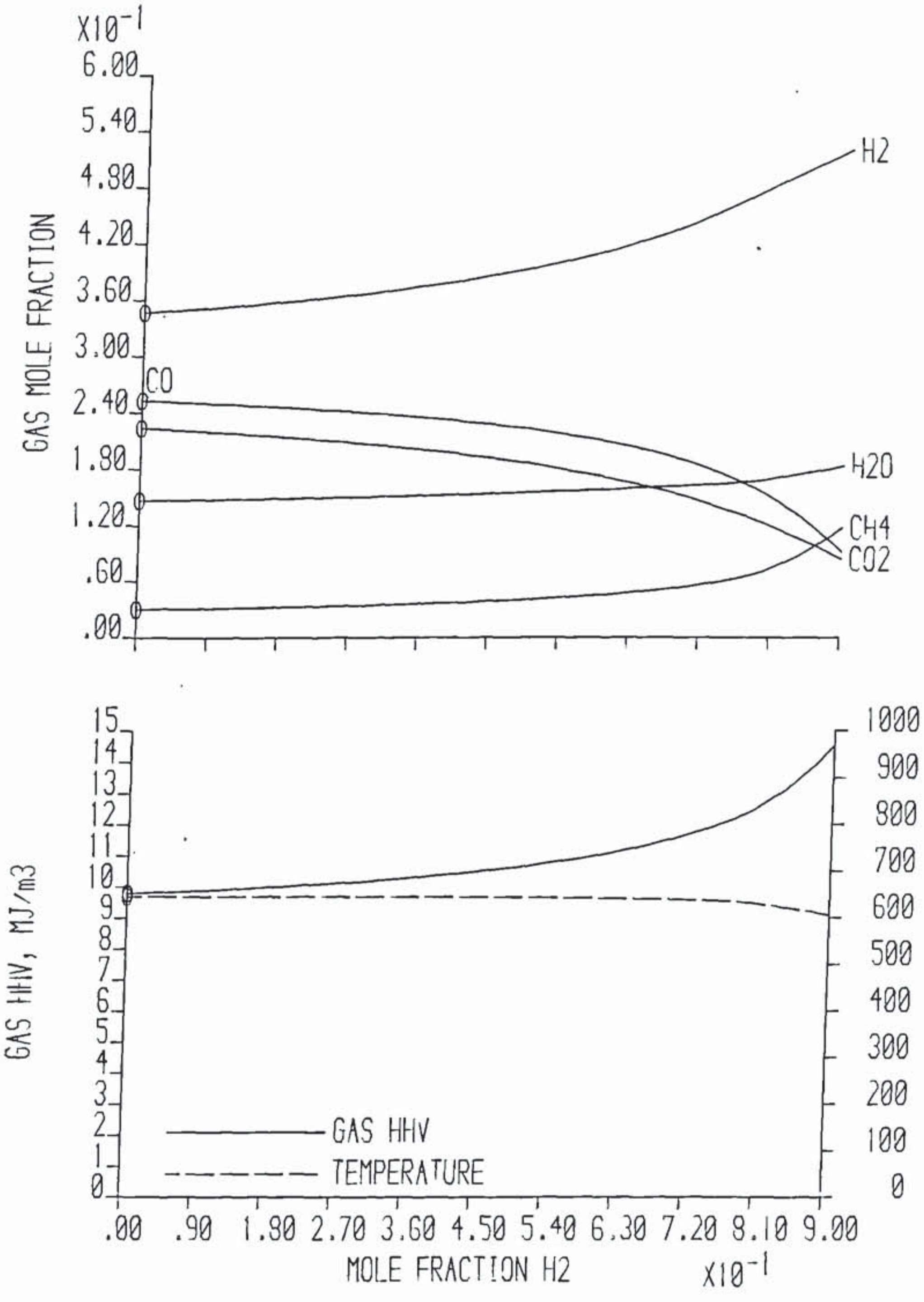

Figure C1 THEORETICAL GASIFIER PERFORMANCE AGAINST GASIFYING AGENT H2 MOLE FRACTION
OXYGEN GASIFER

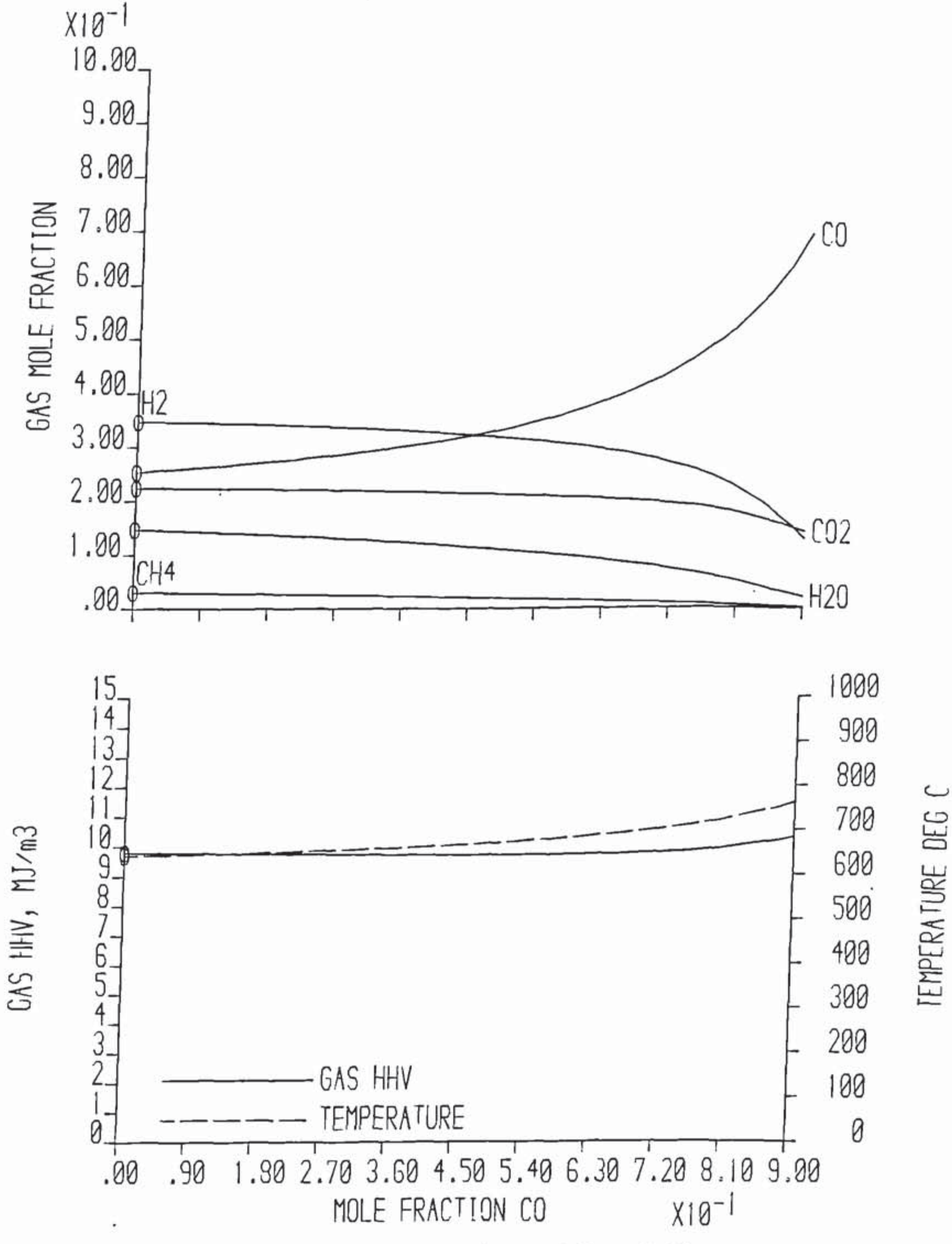

Figure C2 THEORETICAL GASIFIER PERFORMANCE AGAINST GASIFYING AGENT CO MOLE FRACTION OXYGEN GASIFER

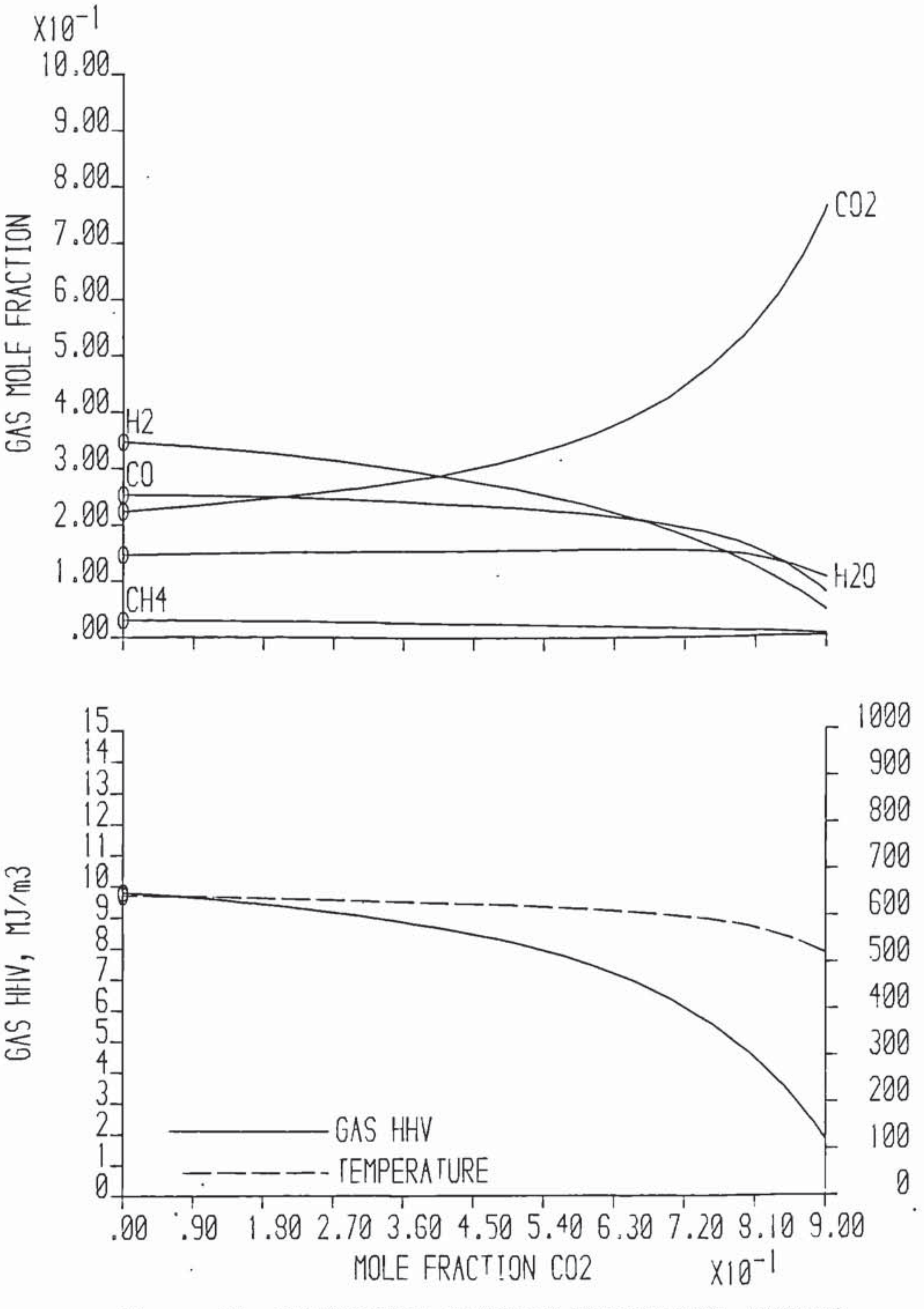

Figure C3 THEORETICAL GASIFIER PERFORMANCE AGAINST GASIFYING AGENT CO2 MOLE FRACTION OXYGEN GASIFER

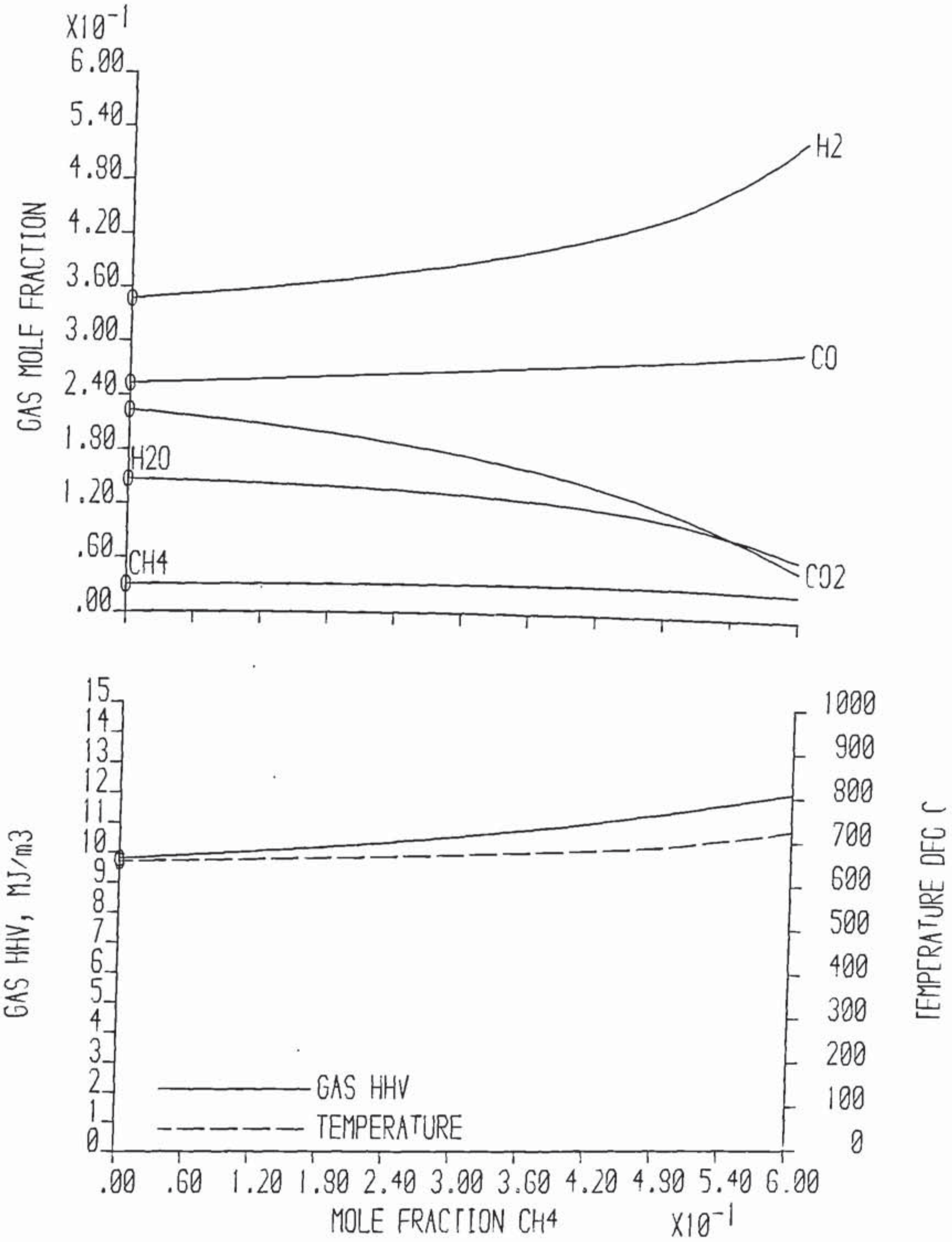

Figure C4 THEORETICAL GASIFIER PERFORMANCE AGAINST GASIFYING AGENT CH4 MOLE FRACTION OXYGEN GASIFER

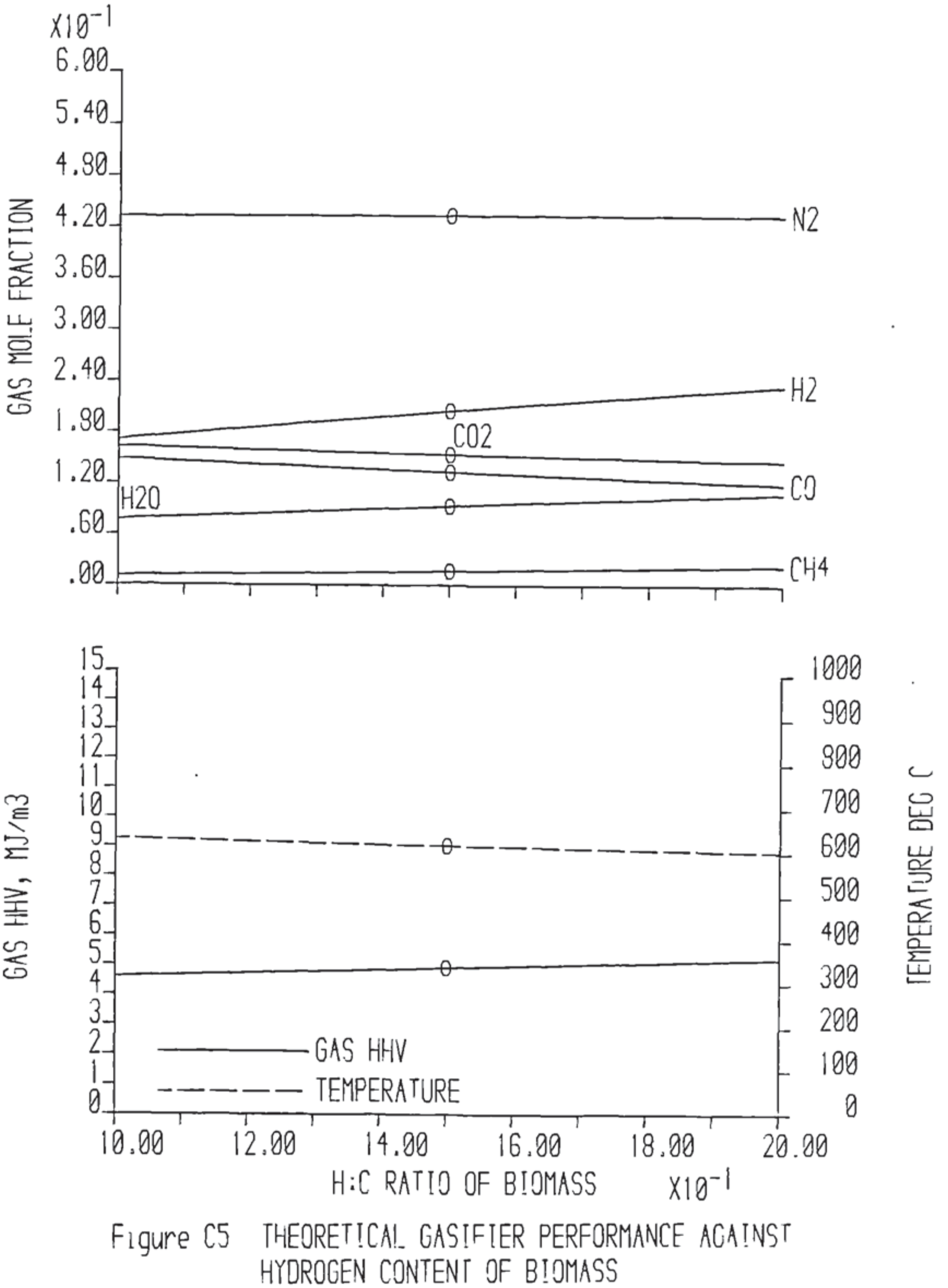

Figure C5 THEORETICAL GASIFIER PERFORMANCE AGAINST HYDROGEN CONTENT OF BIOMASS AIR GASIFIER

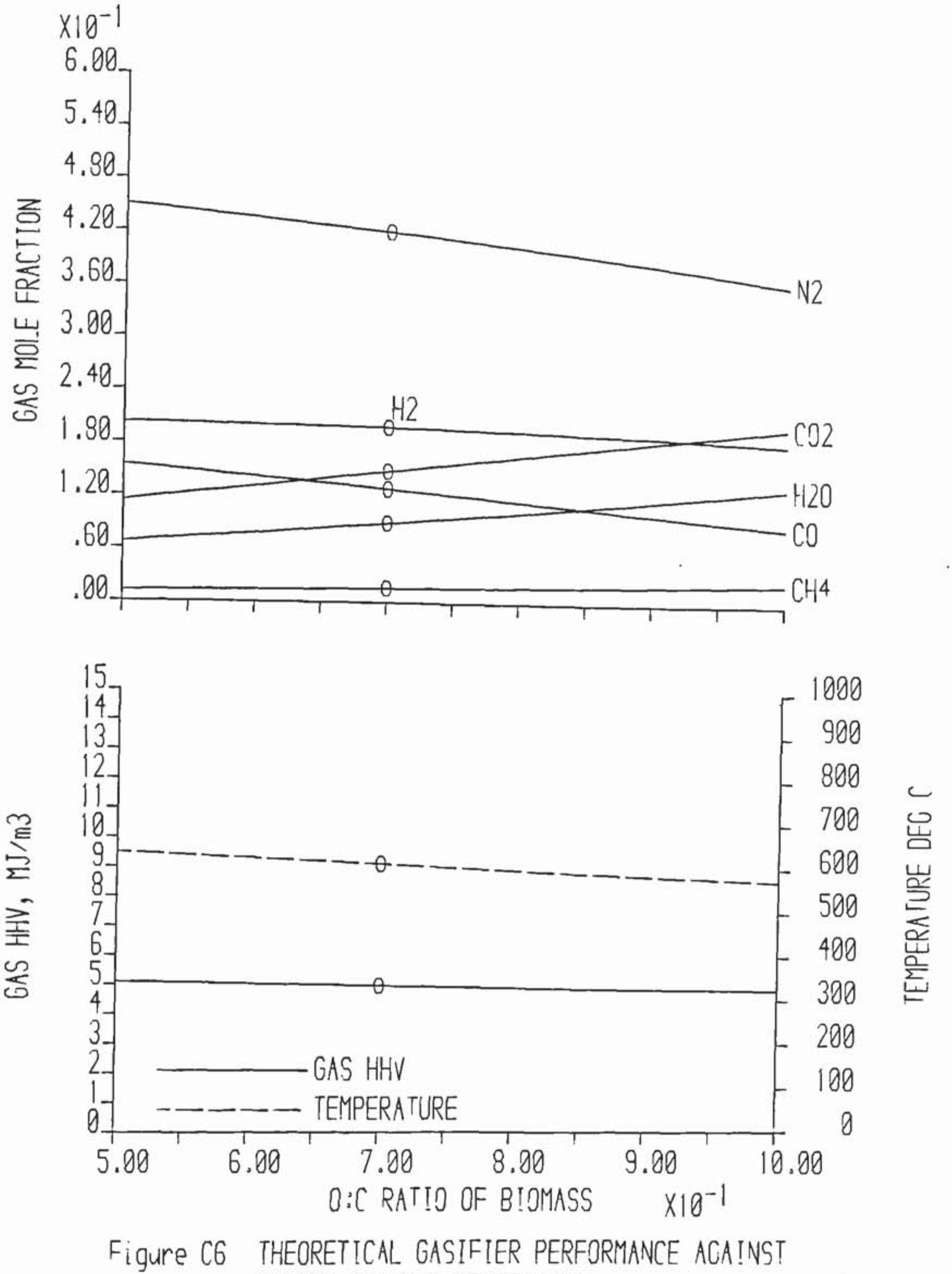

Figure C6 THEORETICAL GASIFIER PERFORMANCE AGAINST
OXYGEN CONTENT OF BIOMASS
AIR GASIFIER

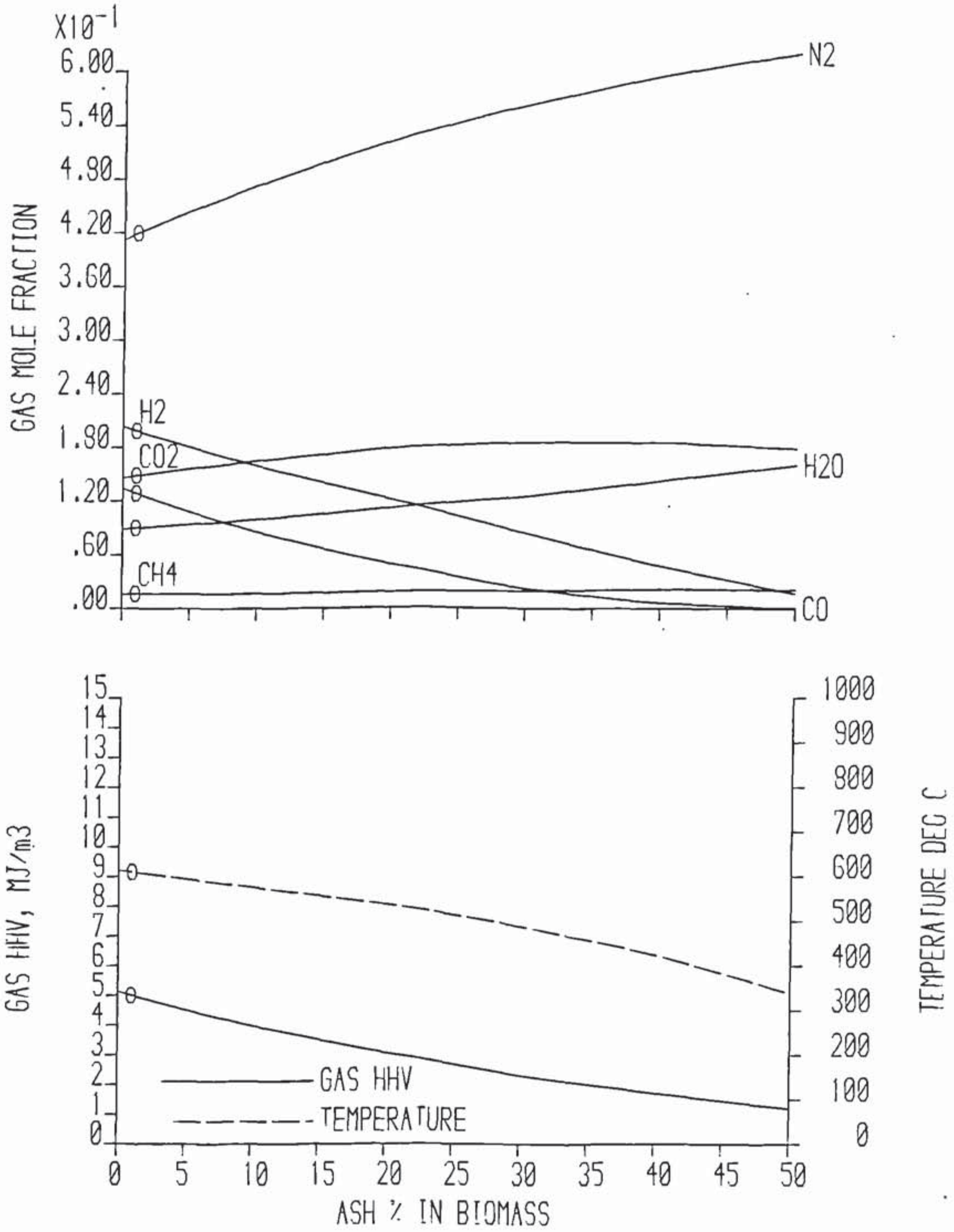

Figure C7 THEORETICAL GASIFIER PERFORMANCE AGAINST
BIOMASS ASH CONTENT
AIR GASIFIER

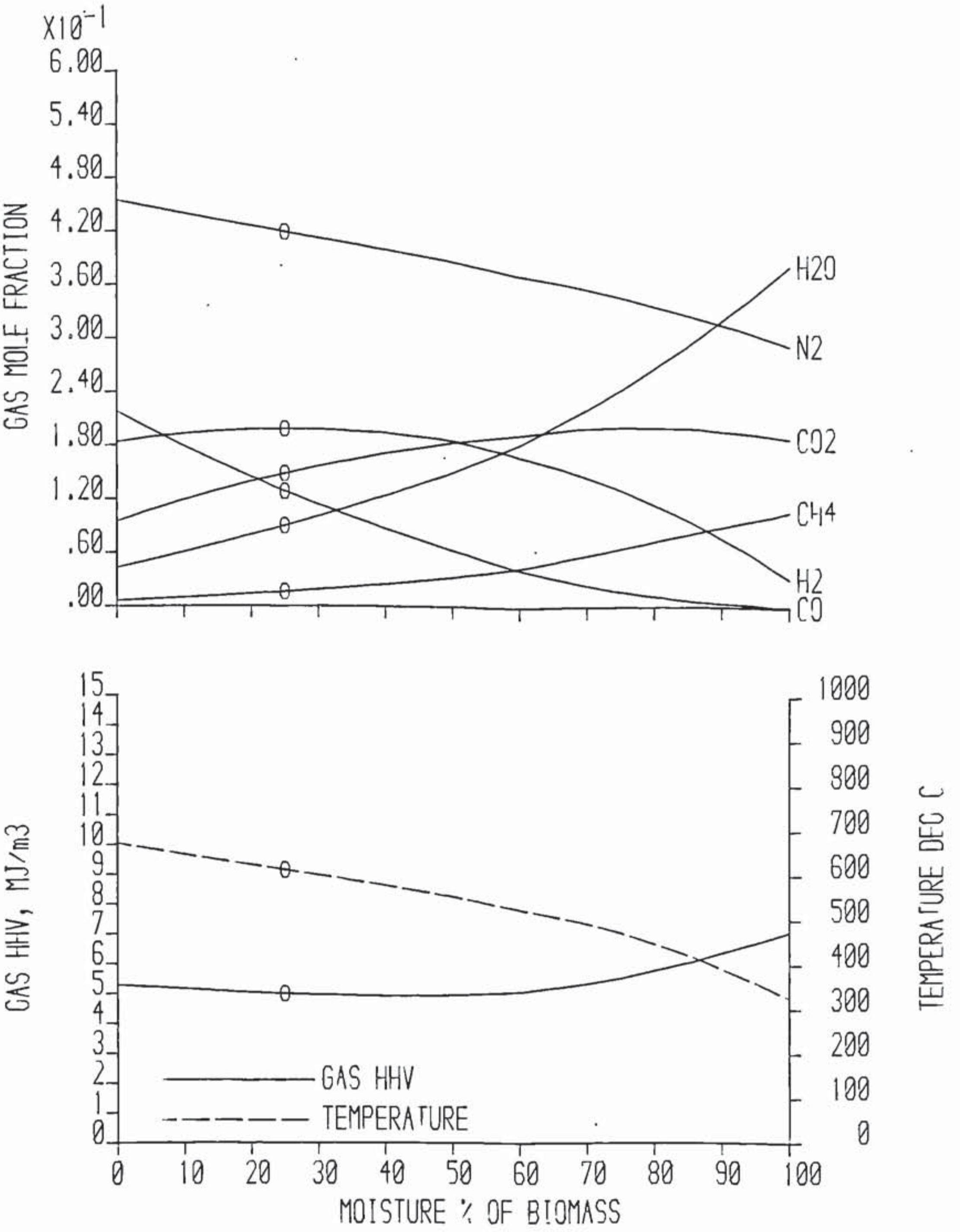

Figure C8 THEORETICAL GASIFIER PERFORMANCE AGAINST MOISTURE CONTENT OF BIOMASS AIR GASIFIER

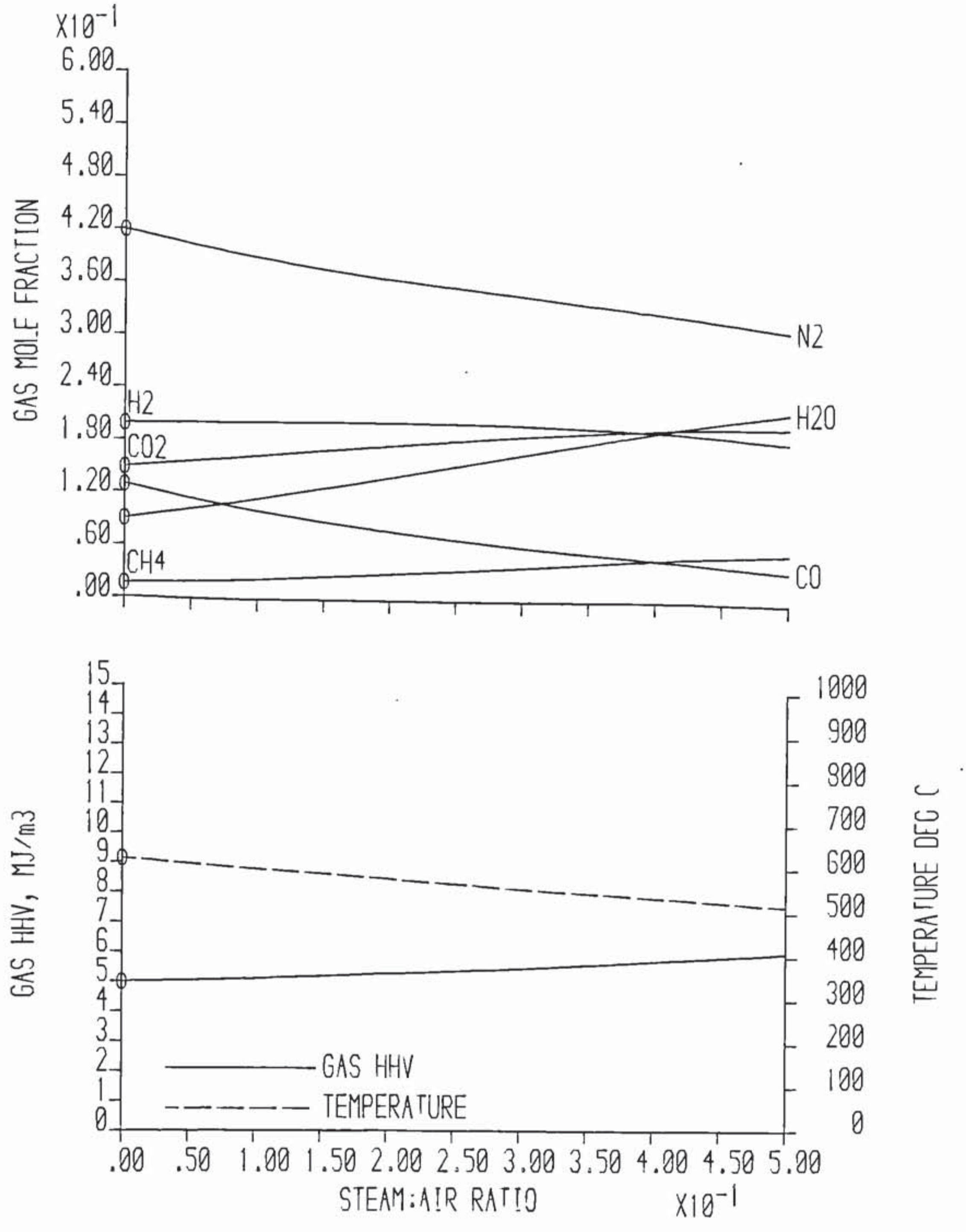

Figure C9 THEORETICAL GASIFIER PERFORMANCE AGAINST
GASIFYING STEAM:AIR RATIO
AIR GASIFIER

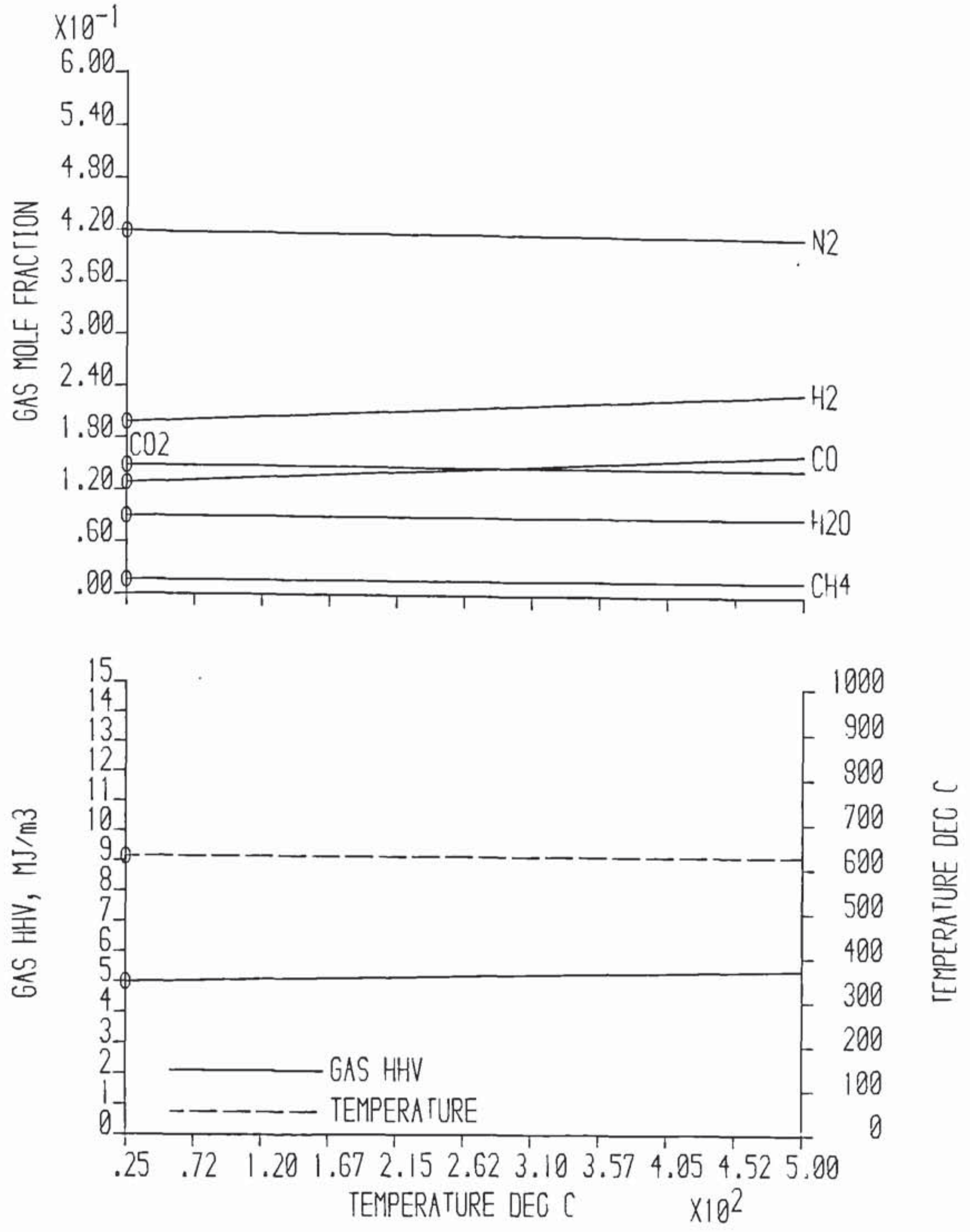

Figure C10 THEORETICAL GASIFIER PERFORMANCE AGAINST GASIFYING AGENT TEMPERATURE AIR GASIFIER

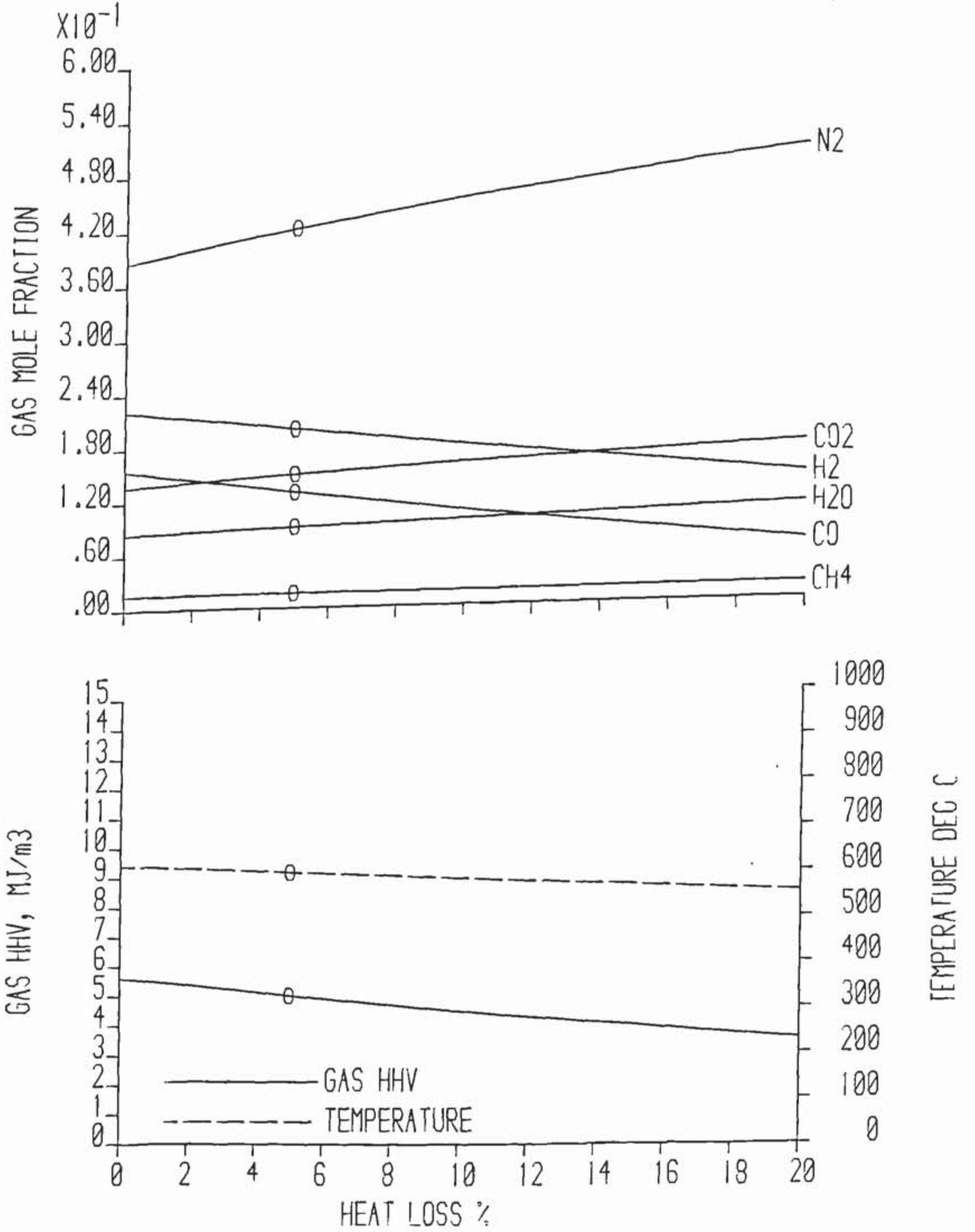

Figure C11 THEORETICAL GASIFIER PERFORMANCE AGAINST
GASIFIER HEAT LOSS
AIR GASIFIER

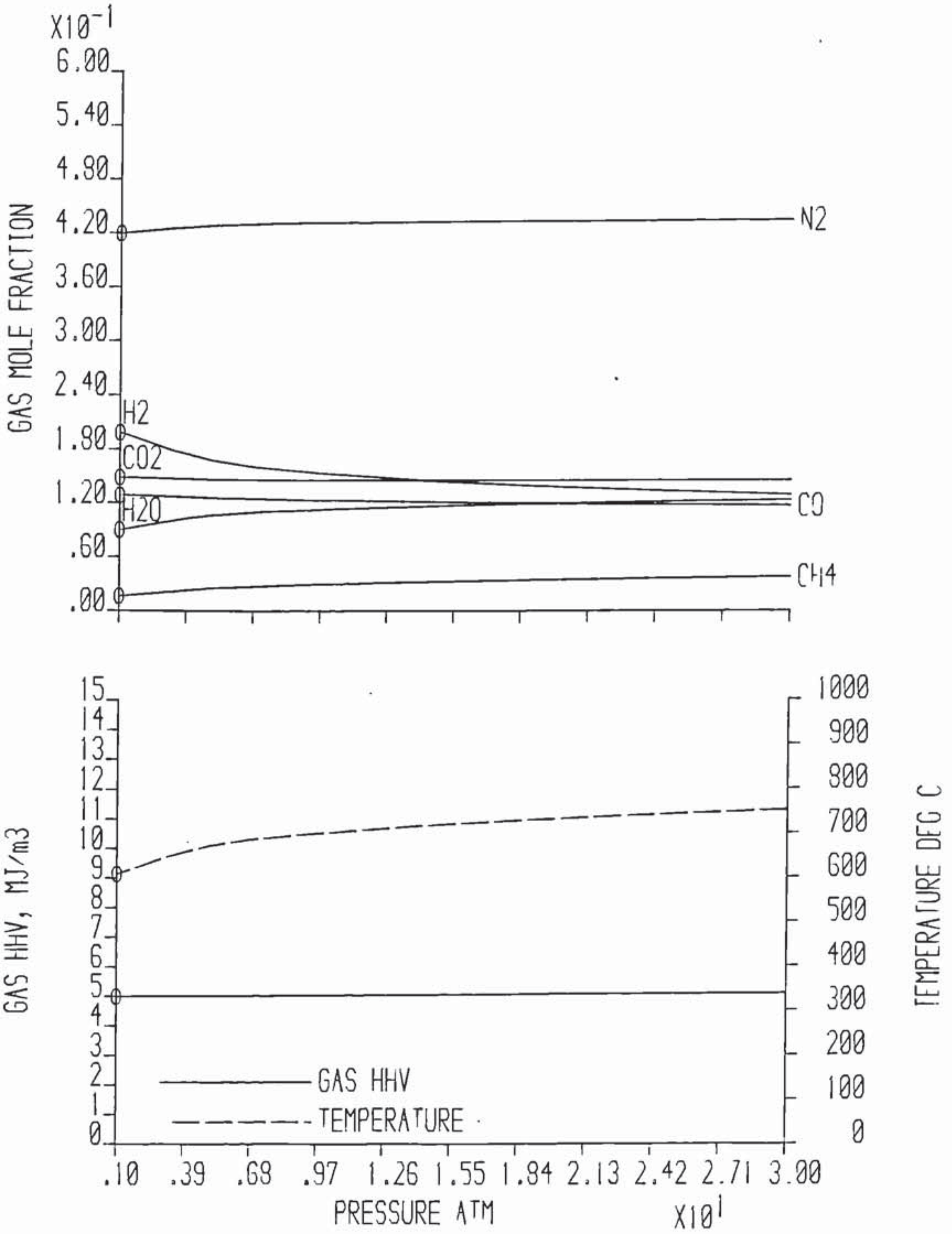

Figure C12 THEORETICAL GASIFIER PERFORMANCE AGAINST
GASIFIER PRESSURE
AIR GASIFIER

APPENDIX D

GASIFIER CAPITAL COST DATA

Table D1 Equipment Costs of Gasifiers

No	Company / Trade Name	Cap. daft/h	Date	Equip. Cost Original Units	Equip. Cost 1985 £	Notes	Ref
1	John Brown/Wellman ODG Twin Fluid Bed	5.0	10/84	£873 800	907 400	Total Plant = £2 037 500	101
2	Applied Engineering Co. Updraft	1.45	1982	Can $ 800 000	550 300	Includes burner for retrofit	43
3	Century Research Inc Updraft	1.9	1982	Can $ 300 000	206 400	Very rough estimate. Installed cost = 1.66 x DEC	43
4	EZ Manufacturing Updraft	0.22	1982	Can $ 65 000	44 700		43
5	Forest Fuels Updraft moving grate	0.35	1980	US $147 000	131 600		43
6	Biomass Corporation Downdraft	0.17	1982	Can $155 000	106 600	Installed cost = 1.15 to 1.2 x DEC	43
7	Biomass Corporation Downdraft	0.43	1982	Can $194 000	133 500	Installed cost = 1.15 to 1.2 x DEC	43
8	Biomass Corporation Downdraft	0.67	1982	Can $ 241 350	166 000	Installed cost= 1.15 to 1.2 x DEC	43
9	Duvant Moteurs Diesel Downdraft Model GA10	0.18	1982	Can $ 84 000	57 800		43
10	Duvant Moteurs Diesel Downdraft Model GA10	0.45	1982	Can $130 000	89 400		43
11	Duvant Moteurs Diesel Downdraft model GA20	0.7	1982	Can $320 000	220 700	Cost fob France	43

Table D1 Equipment Costs of Gasifiers (continued)

No	Company / Trade Name	Cap. daft/h	Date	Equip. Cost Original Units	Equip. Cost 1985 £	Notes	Ref
12	Energy Equipment Engineering Downdraft	0.3	1982	Guilders 180 000	43 600	Cost fob Netherlands	43
13	Fritz Werner Downdraft	0.11	1982	Can $86 000	59 200	Cost fob Canada	43
14	Imbert Energietechnik Downdraft	0.29	1982	Can $285 000	196 100	Cost fob Germany Add 7% delivery	43
15	Pillard Downdraft	0.62	1982	Can $150 000	103 200		43
16	Pillard Downdraft	0.11	1982	Can $190 000	130 700	Includes 110kWe engine/generator	43
17	Pyrenco Downdraft	0.30	1982	US $250 000	223 800		43
18	Pyrenco Downdraft	0.52	1982	US $335 000	299 800		43
19	Vyncke Warmetechnik Downdraft	0.12	1982	Can $110 000	75 700		43
20	National Synfuels Stirred Bed	1.2	1982	US $600 000	537 000	Includes micro-processor control	43
21	National Synfuels Stirred Bed	1.8	1982	US $650 000	581 800	Includes micro-processor control	43
22	National Synfuels Stirred Bed	4.1	1982	US $950 000	850 300	Includes micro-processor control	43

Table D1 Equipment Costs of Gasifiers (continued)

No	Company / Trade Name	Cap. daft/h	Date	Equip. Cost Original Units	Equip. Cost 1985 £	Notes	Ref
23	Alberta Industrial Developments Fluid Bed	2	1982	US $200 000	179 000	Based on data for fluid bed peat pyrolyser (below)	43
24	Energy Products of Idaho; Fluid Bed	1.6	1982	US $300 000	268 500	Slightly adapted fluid bed combustor	43
25	Sur-Lite Fluid Bed	0.59	1982	US $150 000	134 300	Skid mounted system with Cyclone gas cleaning	43
26	Sur-Lite Fluid Bed	2.4	1982	US $400 000	358 000	Skid mounted system with Cyclone gas cleaning	43
27	Sur-Lite Fluid Bed	4.7	1982	US $650 000	581 800	Skid mounted system with cyclone gas cleaning	43
28	Pyro-Sol Kiln Pyrolyser	2.3	1982	US $672 000*	601 400*	*Installed plant cost	43
29	Alberta Industrial Developments Fluid Bed Pyrolyser	1.7	1982	Can $200 000	137 600		43
30	KHD model DUM 506 Downdraft	0.1	7/85	DM674 000	174 800	Package units mounted in iso-container	101
31	KHD model DUM 509 Downdraft	0.2	7/85	DM864 000	224 000	Package units mounted in iso container	101

Table D1 Equipment Costs of Gasifiers (continued)

No	Company / Trade Name	Cap. daft/h	Date	Equip. Cost Original Units	Equip. Cost 1985 £	Notes	Ref
32	KHD model DUM 514 Downdraft	0.36	7/85	DM1 14 000	288 900	Package units mounted in iso container	101
33	Ebara Pyrolyser for RDF	4x10.4	1984	US $24 326 000	20 540 900		101
34	Ebara Pyrolyser for sorted refuse	3x10	1984	US $33 077 000	27 930 200		101
35	TNEE	7	7/83	FF30 000 000*	3 051 000*	* Installed plant cost	101
36	CPC Fluid Bed	2.6 50 mBTU	10/84	US $950 000	802 200		101
37	ADENAP Fluid Bed (Sur-Lite)	0.55t 10 mBTU	1/83	US $160 000	139 600	Skid mounted units feed to Hot Clean Gas. cf 25,26,27	102
38	ADENAP Fluid Bed (Sur-Lite)	1.1t 20 mBTU	1/83	US $255 000	222 500	Skid mounted units feed to hot clean gas, cf 25,26,27	102
39	ADENAP Fluid Bed (Sur-Lite)	1.6t 30 mBTU	1/83	US $355 000	309 700	Skid mounted units feed to Hot Clean Gas, cf 25,26,27	102
40	ADENAP Fluid Bed (Sur-Lite)	2.1t 40 mBTU	1/83	US $410 000	357 800	Skid mounted units feed to Hot Clean Gas, cf 25,26,27	102

Table D1 Equipment Costs of Gasifiers (continued)

No	Company / Trade Name	Cap. daft/h	Date	Equip. Cost Original Units	Equip. Cost 1985 £	Notes	Ref
41	ADENAP Fluid Bed (Sur-Lite)	3.2t 60 mBTU	1/83	US $545 000	475 600	Skid mounted units feed to Hot Clean Gas, cf 25,26,27	102
42	ADENAP Fluid Bed (Sur-Lite)	4.2t 80 mBTU	1/83	US $660 000	575 900	Skid mounted units feed to Hot Clean Gas, cf 25,26,27	102
43	Energy Equipment Engineering Downdraft	0.15	1983	US $50 000	42 900	US $60 000 including storage Skid mounted cf 12	102
44	Energy Products of Idaho Fluid Bed	2.8t 53 mBTU	1983	US $1 580 000*	1 345 000*	* Total plant cost including boiler retrofit, cf 24	102
45	Foster Wheeler Power Fluid Bed	7.0	1985	£1 710 000*	1 710 000*	* Total plant cost	101
46	Alternate Energy Industries Horizontal Moving Bed Pyrolyser	1.0	1985	£225 000	225 000	cf 28	101
47	Alternate Energy Industries Horizontal Moving Bed Pyrolyser	4.5	1985	£425 000	425 000		101
48	Battelle Columbus	20.8	1981	US $2 594 500	2 562 800		101

Table D1 Equipment Costs of Gasifiers (continued)

No	Company / Trade Name	Cap. daft/h	Date	Equip. Cost Original Units	Equip. Cost 1985 £	Notes	Ref
49	Buck Rogers	0.2	1985	US $57 000	45 000	Includes transport at 10%	101

Table D2 Total Plant Costs of Gasifiers

No	Company and/or Trade Name	Cap. daft/h	Total Plant Cost Original Units	Total Plant Cost In 1985 £	Notes	Ref
1	John Brown/Wellman ODG Twin Fluid Bed	5.0	£1 962 000	2 037 500		101
2	Applied Engineering Updraft	1.45	Taken as 1.75 x Equipment	963 000		
3	Century Research Inc Updraft	1.9	1.66 x Equipment	342 624	1.66 x Equipment quoted with costs	43
4	EZ Manufacturing Updraft	0.22	Taken as 1.75 x Equipment	78 225		
5	Forest Fuels Updraft moving grate	0.35	Taken as 1.75 x Equipment	230 300		
6	Biomass Corporation Downdraft	0.17	1.2 x Equipment	127 900	1.15 to 1.2 x Equip. quoted with costs	43
7.	Biomass Coporation Downdraft	0.43	1.2 x Equipment	160 200	1.15 to 1.2 x Equip. quoted with costs	43
8	Biomass Coporation Downdraft	0.67	1.2 x Equipment	199 200	1.15 to 1.2 x Equip. quoted with costs	43
9	Duvant Moteurs Diesel Downdraft Model GA10	0.18	Taken as 1.75 x Equipment	101 000		
10	Duvant Moteurs Diesel Downdraft Model GA10	0.45	Taken as 1.75 x Equipment	156 450		
11	Duvant Moteurs Diesel Downdraft model GA20	0.7	Taken as 1.75 x Equipment	386 200		
12	Energy Equipment Engineering Downdraft	0.3	See No 43			

Table D2 Total Plant Costs of Gasifiers (Continued)

No	Company and/or Trade Name	Cap. daft/h	Total Plant Cost Original Units	Total Plant Cost In 1985 £	Notes	Ref
13	Fritz Werner Downdraft	0.11	Taken as 1.2 x Equipment	71 040		
14	Imbert Energietechnik Downdraft	0.29	Taken as 1.27 x Equipment	249 047	1.2 x Equipment + 7% for delivery	43
15	Pillard Downdraft	0.62	Taken as 1.75 x Equipment	180 600		
16	Pillard Downdraft				Not applicable - includes generator	43
17	Pyrenco Downdraft	0.30	Taken as 1.75 x Equipment	391 650		
18	Pyrenco Downdraft	0.52	Taken as 1.75 x Equipment	524 650		
19	Vyncke Warmetechnik Downdraft	0.12	Taken as 1.2 x Equipment	90 840		
20	National Synfuels Stirred Bed	1.2	Taken as 2.25 x Equipment	1 208 250		
21	National Synfuels Stirred Bed	1.8	Taken as 2.25 x Equipment	1 309 050		
22	National Synfuels Stirred Bed	4.1	Taken as 2.25 x Equipment	1 913 175		
23	Alberta Industrial Developments Fluid Bed	2	Taken as 2.25 x Equipment	402 800		
24	Energy Products of Idaho; Fluid Bed	1.6	Taken as 2.25 x Equipment	604 100		

Table D2 Total Plant Costs of Gasifiers (Continued)

No	Company and/or Trade Name	Cap. daft/h	Total Plant Cost Original Units	Total Plant Cost In 1985 £	Notes	Ref
25	Sur-Lite Fluid Bed	0.59	Taken as 1.2 x Equipment	161 200		
26	Sur-Lite Fluid Bed	2.4	Taken as 1.2 x Equipment	429 600		
27	Sur-Lite Fluid Bed	4.7	Taken as 1.2 x Equipment	698 200		
28	Pyro-Sol Kiln Pyrolyser	2.3	US $672 000	601 400		43
29	Alberta Industrial Developments Fluid Bed Pyrolyser	1.7	Taken as 2.25 x Equipment	309 600		
30	KHD model DUM 506 Downdraft	0.1	Taken as 1.2 x Equipment	209 760		
31	KHD model DUM 509 Downdraft	0.2	Taken as 1.2 x Equipment	268 800		
32	KHD model DUM 514 Downdraft	0.36	Taken as 1.2 x Equipment	346 680		
33	Ebara Twin fluid bed	4x10.4		2 310 000	Cost of 1 system @ 45% total system cost	101
34	Ebara Twin fluid bed	3 x 10		4 190 000	Cost of 1 system @ 45% total system cost	101
35	TNEE	7	FF30 000 000	3 051 000		101

Table D2 Total Plant Costs of Gasifiers (Continued)

No	Company and/or Trade Name	Cap. daft/h	Total Plant Cost Original Units	Total Plant Cost In 1985 £	Notes	Ref
36	CPC Fluid Bed	2.6t	Taken as 2.25 x Equipment	1 805 000		101
37	ADENAP (Sur-Lite) Fluid Bed	0.55t	Taken as 1.2 x Equipment	167 520	cf 25,26,27	
38	ADENAP (Sur-Lite) Fluid Bed	1.1t	Taken as 1.2 x Equipment	267 000	cf 25,26,27	
39	ADENAP (Sur-Lite) Fluid Bed	1.6t	Taken as 1.2 x Equipment	371 600	cf 25,26,27	
40	ADENAP (Sur-Lite) Fluid Bed	2.1t	Taken as 1.2 x Equipment	429 400	cf 25,26,27	
41	ADENAP (Sur-Lite) Fluid Bed	3.2t	Taken as 1.2 x Equipment	570 700	cf 25,26,27	
42	ADENAP (Sur-Lite) Fluid Bed	4.2t	Taken as 1.2 x Equipment	691 100	cf 25,26,27	
43	Energy Equipment Engineering Downdraft	0.15	US $74 000	63 500	Includes boiler retrofit, cf 12	102
44	Energy Products of Idaho Fluid Bed	2.8	US $1 580 000	1 345 000	Includes boiler retrofit, cf 24	102
45	Foster Wheeler Power Fluid Bed	7.0	£1 710 000	1 710 000		101
46	Alternate Energy Ind. Horizontal Moving Bed Pyrolyser	1.0	Taken as 2.25 x Equipment	506 500		

Table D2 Total Plant Costs of Gasifiers (Continued)

No	Company and/or Trade Name	Cap. daft/h	Total Plant Cost Original Units	Total Plant Cost In 1985 £	Notes	Ref
47	Alternate Energy Ind. Horizontal Moving Bed Pyrolyser	4.5	Taken as 2.25 x Equipment	956 200		
48	Battelle Columbus	20.8	US $4 874 400	4 819 932		101
49	Buck Rogers	0.2	Taken as 1.75 x Equipment	79 000		

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