

**EVALUATION OF A DOWNDRAFT WOOD GASIFIER  
FOR TEA MANUFACTURING IN SRI LANKA**

**Tuan Haris Jayah**

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**THE UNIVERSITY OF  
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**Department of Civil and Environmental Engineering  
The University of Melbourne**

*Sincerely dedicated in memory of my late parents*

## ABSTRACT

One of the main reasons for the high cost of production in the tea industry in Sri Lanka is the high energy consumption. It has been found that to produce a kilogram of “made” tea in Sri Lanka, an average 22.4 MJ of thermal energy is used. Eighty five percent of the thermal energy requirement is met by burning fuel wood of which 70% is rubber wood. Rubber wood is now becoming expensive and scarce, as the wood has now been re-discovered for furniture making and acclaimed by experts as the timber of the future. Therefore greater energy efficiency is required by the Sri Lankan tea industry in order to achieve sustainability.

In this thesis, various energy options are investigated in terms of their potential and the quality of energy that can be produced by each of these resources in order to reduce the fuel wood consumption. The development of local technologies to harness those resources for tea processing and a life cycle energy cost analysis of the available technologies are carried out to assess their financial feasibility. From a financial standpoint, gasifiers using wood are identified as a promising choice. Problems associated with the design and operation of a locally fabricated gasifier raised doubts about its applicability and productivity for the tea industry. Therefore, a computer program based on available models has been developed to study the performance of the gasifier. Experimental results from a locally fabricated gasifier are used to calibrate the model, which was then used to investigate the effects of operating parameters such as air fuel ratio, chip size, fuel moisture content, inlet air temperature and design parameters such as insulation level and throat angle on the conversion efficiency of the gasifier.

Based on the simulation study, the required length of gasification zone for maximum char conversion, the desirable range of wood size, moisture content and heat loss for the locally fabricated downdraft gasifier are presented.

## **DECLARATION**

This is to certify that

- (i) the thesis is approximately 30,000 words in length, exclusive of tables, maps, bibliographies, appendices and footnotes,
- (ii) due acknowledgement has been made in the text to all other material used.

Tuan Haris Jayah

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## NOMENCLATURE

$A_{pT}$	Total particle surface area	$m^2$
$C_i$	Concentration of $i^{\text{th}}$ element	kmol
$C_P$	Specific heat capacity	kJ/kmol K
$D_e$	Effective diffusivity	$m^2/s$
$D_{ij}$	Binary diffusivity between gas component i and j	$m^2/s$
$D_{im}$	Binary diffusivity between gas component i and gas mixture	$m^2/s$
$D_j$	Diffusivity of component i	$m^2/s$
$F_i$	Mass fraction of $i^{\text{th}}$ element in wood	%
$F_{i,cha}$	Mass fraction of $i^{\text{th}}$ element in char	%
$r$		
$F_{i,tar}$	Mass fraction of $i^{\text{th}}$ element in tar	%
$G$	Superficial gas mass flux	$kg/m^2 s$
$H_c$	Higher heating value	MJ/kg
$H_{gas}$	Enthalpy flux of gas	kJ/kmol
$H_{g-s}$	Heat flux transfer from gas to solid	kJ/kmol
$H_i$	Specific heat of the $i^{\text{th}}$ component	kJ/kg K
$H_L$	Heat loss	kJ/kmol
$H_s$	Sensible heat	kJ/kmol
$H_{solid}$	Enthalpy flux of solid	kJ/kmol
$h$	Heat transfer coefficient between gas and particle surface	$W/m^2 K$
$J_H$	Colburn factor	-
$K$	Reaction rate constant for char gasification reaction	$m^2/mol^{-0.7}s$
$K_B$	Boltzmann constant	$W/m^2 K^4$
$K_1$	Equilibrium constant of Boudouard reaction	-
$K_2$	Equilibrium constant of water-gas shift reaction	-
$k$	Mass transfer coefficient	m/s
$k_e$	Effective mass transfer coefficient	m/s
$k_m$	Mass transfer coefficient of the mixture	m/s
$k_{mi}$	Mass transfer coefficient of the $i^{\text{th}}$ element in the mixture	m/s
$M_i$	Molecular weight of component i	kg/kmol

$m$	Moles of air per moles of wood	kmol/kmol
$mc_{db}$	Moisture content of wood on dry basis	%
$N_i$	Molar flux of $i^{\text{th}}$ component	kmol
$n_p$	Number of particles	-
$P$	Pressure	Pa
$Pr$	Prandtl number	-
$R$	Ideal gas constant	kJ/kg K
$Re$	Reynolds number	-
$R$	Particle radius at any time	m
$R_0$	Initial particle radius	m
$R_p$	Particle radius	m
$Sc$	Schmidt number	-
$T$	Temperature	K
$T_b$	Gas bulk temperature	K
$T_g$	Gas temperature	K
$T_r$	Reference temperature	K
$T_s$	Solid temperature	K
$T$	Time	s
$v_g$	Gas velocity	m/s
$V_p$	Particle volume	$\text{m}^3$
$w$	Moles of water per mole of wood	kmol/kmol
$w_s$	Shift of water-gas reaction	kmol
$x_i$	Moles of $i^{\text{th}}$ component per mole of wood	kmol
$x_{char}$	Moles of char per mole of wood	kmol
$x_{tar}$	Moles of tar per mole of wood	kmol
$Y_i$	Molar fraction of $i^{\text{th}}$ component	%
$\Delta H_1$	Change of enthalpy of Boudouard reaction	kJ/kmol
$\Delta H_2$	Change of enthalpy of water-gas reaction	kJ/kmol
$\Delta H_3$	Change of enthalpy of water-gas shift reaction	kJ/kmol
$\Delta L$	Distance	m
$\Delta t$	Time step	s
$\varepsilon_b$	Bed void fraction	-
$\varepsilon_{ij}$	Energy of molecular interaction	kJ

$\mu_i$	Gas viscosity of $i^{\text{th}}$ component	kg/m s
$\mu_{mix}$	Gas viscosity of the mixture	kg/m s
$\rho$	Gas density	kg/m <sup>3</sup>
$\sigma_{ij}$	Collision diameter	m
$\Omega_D$	Collision integral	-

# CHAPTER 1

## INTRODUCTION

### 1.1 Background to the Tea Industry in Sri Lanka

Sri Lanka is an agricultural country where the economy is largely dependent upon agricultural products such as tea, rubber and coconut. Tea is one of the major contributors to the Sri Lankan economy and is the top net foreign exchange earner in the country. In 2000, earnings from tea amounted to US\$700 million of foreign exchange and tea production reached  $306 \times 10^3$  tonnes (Central Bank of Sri Lanka 2001). Sri Lanka exports around 90-95 per cent of its annual tea production (de Silva 1994), is the largest tea exporter in the world (Leonard 2001) and the third largest tea producer behind India and China (Haskoning 1989).

The tea industry in Sri Lanka has a long history stretching back more than a century. The first commercial planting of tea in Sri Lanka was undertaken in 1867 on 8 ha of land on Loolecondera Estate, Hewaheta (Haskoning 1989). In 2000, the area under tea cultivation was 180,000 ha, i.e. 2.8% of the total land area (Central Bank of Sri Lanka 2001) and presently there are 594 tea factories operating in Sri Lanka (Ziyad Mohamed 1998). High quality tea is grown in the “up country” regions where the altitude is higher than 1200 m above mean sea level, and partly in the “mid-country” regions where the altitude is between 600 and 1200 m above mean sea level. The low grade tea is produced in the remaining part of the “mid-country” and in the “low country” regions where the altitude is less than 600 m above mean sea level.

Tea is manufactured using “Crush-Tear-Curl” (CTC), “Orthodox or Conventional” and “Rotor Vane” manufacturing processes. Although the Rotor Vane manufacturing process is also used in some factories, Sri Lanka mainly produces tea using conventional technology, and the product is known as “Orthodox” tea. The more efficient “CTC” manufacturing process is not widely used in Sri Lanka but it is very

popular in the Indian subcontinent and East African countries. In 1985, “CTC” tea amounted to only 1% of the total production (Haskoning 1989), although by 1998, this figure had reached 6% (Cassim 1999).

Although changes are in progress from the “Orthodox” to the CTC system because of the low production cost of the latter and the demand in the world market for CTC manufactured tea, the pace at which the change is taking place is slow. This is probably due to the high cost of replacing the existing orthodox rollers, shifters etc., with CTC rollers and the high competition in finding new CTC tea markets at the expense of the well established existing orthodox tea market.

Almost sixty per cent of tea is produced by small-scale growers whose land holdings are less than 20 ha. The remaining forty per cent is produced by estates managed by private companies. Presently there are about 250,000 small-scale growers engaged in the tea industry (Ziyad Mohamed 1998). In addition to these growers, of the island’s 860,000 registered plantation workers, half of them are engaged in the tea plantations (Sivaram & Herath 1996). These factors show that the tea industry is one of the main pillars of the Sri Lankan economy.

## **1.2 Problems Faced by the Tea Industry**

The tea industry in Sri Lanka is faced with several problems that result in a high cost of production relative to other producer countries. Low yield is one of the major problems. The average yield of “made” or finished tea is 1320 kg per hectare tonnes (Central Bank of Sri Lanka 2001) and this is low compared to India and Kenya which produce around 1700 kg of “made” tea per hectare (Haskoning 1989). Higher labour costs (Table 1.1) for picking and low labour productivity are the other problems that the industry is facing. Statistics show that in overall terms, labour cost accounts for approximately 56% of the cost of production (Figure 1.1).

Table 1.1: Wage rates in competing tea regions in 1995

Country	Wage rate/(US\$/day)
Sri Lanka	1.36
Kenya - Kericho (1994)	1.22
India - Tamil Nadu	1.35
- Kerala	1.27
- Assam	1.04

(Source: Sivaram & Herath 1996)

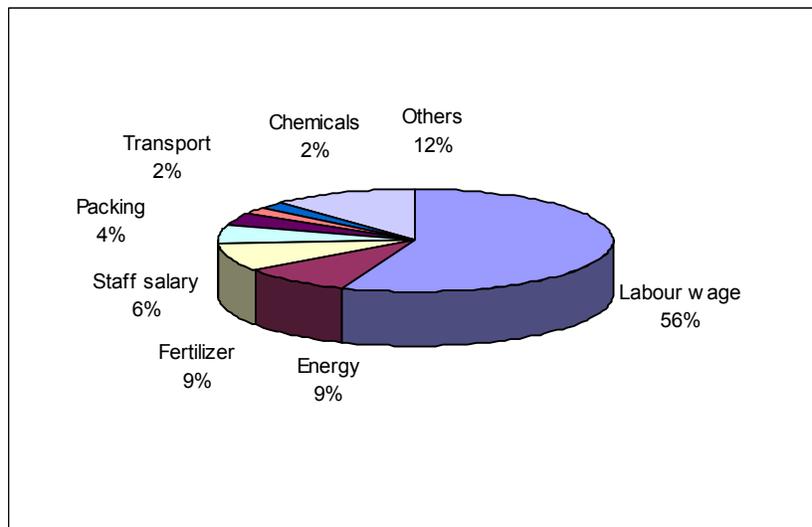


Figure 1.1: Proportion of production cost of tea by expenditure in 1995

(Source: Sivaram & Herath 1996)

The next biggest factor that contributes to the high cost of production is the high energy consumption (Figure 1.1). In Sri Lanka, 1 kg of “made” tea requires approximately 0.75-0.94 kWh of electrical energy and 22.0-22.7 MJ of thermal energy (de Silva 1994), whereas in India, to produce 1 kg of “made” tea 0.55-0.66 kWh of electrical energy and 14.4-18.0 MJ of thermal energy are used (Riva & Palaniappan 1989), which is 40% less than the average energy consumption by the tea industry in Sri Lanka. Gupta (1983) has quoted the figures of 0.7 kWh of electrical energy and 10.8 MJ of thermal energy to produce 1 kg of “made” tea in India.

In tea processing, 90% of the energy requirement is related to thermal energy in the form of hot air and most of this energy is obtained from fuel wood. Wijesinghe (1988) has quoted from Nanayakkara (1986) that the tea industry in Sri Lanka is the largest consumer of fuel wood (Figure 1.2), electricity (Figure 1.3) and the second largest consumer of fuel oil (Figure 1.4) compared to other local industries.

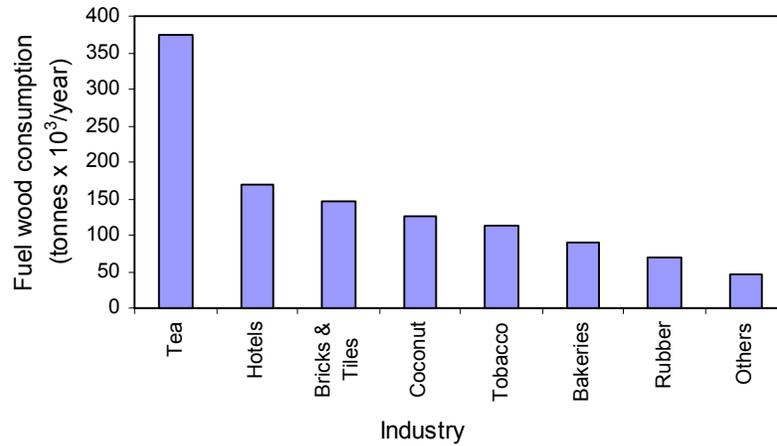


Figure 1.2: Fuel wood consumption in Sri Lanka by industries  
(Source: Haskoning 1989)

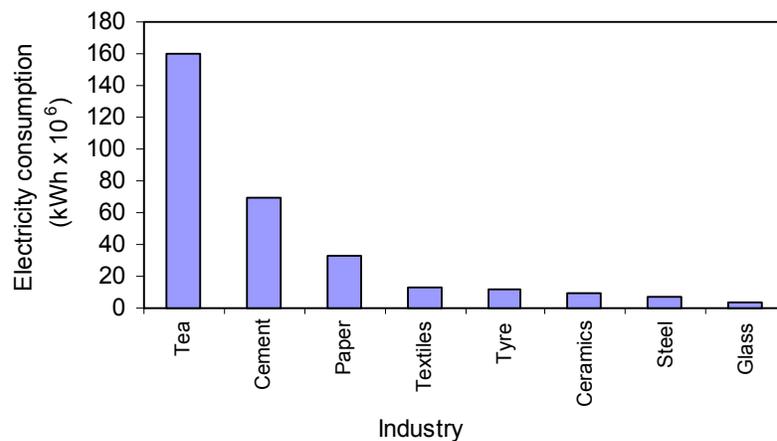


Figure 1.3: The annual electricity consumption of largest industries in Sri Lanka  
(Source: Haskoning 1989)

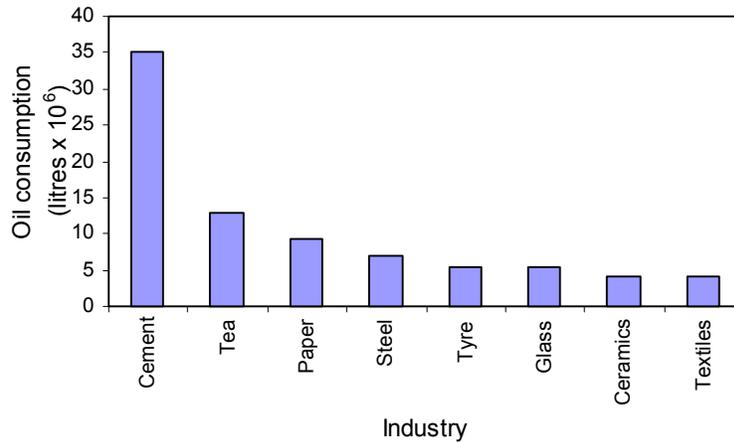


Figure 1.4: The annual oil consumption of largest industries in Sri Lanka  
(Source: Haskoning 1989)

The high energy consumption is mainly due to the huge volume of production and in part due to the low efficiency of equipment such as boilers, heat exchangers, furnaces and the inefficient use of electrical energy in the withering process. Because of the higher prices for fuel oil and electricity, fuel wood has become the main energy source in the tea industry and this dependence is creating concern because local forests are being lost at an alarming rate. Forest cover in Sri Lanka decreased from 44 per cent in 1960 to 24 per cent in 1982 (Navaratna 1985).

### 1.3 Objective of the Study

A major need of the tea industry is to find ways to reduce the cost of production. The cost of production of tea in Sri Lanka is high compared to other countries like India and Kenya. Although labour cost is the largest cost component in tea production, any reduction in labour component would lead to social problems like unemployment. Sri Lanka is a developing country where the unemployment rate is high and half of the registered plantation workers, in addition to the 250,000 small-scale growers, depend on the tea industry. Thus, from the author's perspective, any reduction in the labour component may not be appropriate for Sri Lanka at this stage in the country's development. Therefore, if any cost reduction is to be considered it should focus on the next biggest cost component i.e. energy.

Energy cost contributes approximately nine per cent to the total production cost. Thus, improving the efficiency of energy use will be of utmost importance for any cost reduction. Eighty five percent of the thermal energy is generated by combustion of fuel wood (de Silva 1994). The tea industry, being the largest fuel wood consumer in the country, accounts for 33 per cent of the total industrial fuel wood consumption and 70 per cent of this fuel wood is rubber wood (Haskoning 1989). Rubber wood which is the main source of energy for the tea industry is now becoming expensive and scarce as the wood has now been re-discovered for furniture making and acclaimed by experts as the timber of the future with its eco-friendly properties (Jayatilleke 1997). A shortage of rubber wood for the tea industry is expected in the future.

The decline in forest cover in the country and rising carbon dioxide levels in the atmosphere for which the whole Sri Lankan society is paying “externally” are also important factors that should be taken into account. A reduction in the use of fuel wood would help to reduce the extent of these problems.

Thus, a need has arisen to determine other alternatives to reduce the fuel wood consumption in the Sri Lankan tea industry. As one of the solutions to the problem of the inefficient use of fuel wood in the tea drying process, the Tea Research Institute (TRI) of Sri Lanka with the collaboration of the National Engineering Research and Development (NERD) Centre, Sri Lanka, has been investigating the potential of biomass gasifiers to replace the traditional furnace system. Although initial studies proved the technology to be viable, problems associated with the refractory lining used in the throat and the lack of knowledge of desirable operating parameters raised doubts about its applicability and productivity for the tea industry. Moreover a cost and benefit analysis was required to assess the economic viability of the system.

The overall goal of this study is to contribute to the improved cost competitiveness of the tea industry and to the reduction in the use of natural resources and CO<sub>2</sub> emissions in Sri Lanka. The main objective to achieve this goal is to assist in the improvement of the downdraft gasifier that has been developed by the NERD Centre for tea drying in Sri Lanka.

The main objective is achieved by investigating:

- the specific energy requirements of the tea process,
- the renewable energy resources available to meet those energy needs,
- the cost effectiveness of the gasifier technology to harness those resources and to compare it with other locally available technologies,
- the developments in the gasifier technology that have occurred locally,
- the development of a computer program to carry out a parametric study of the gasifier,
- the effects of operating and design parameters on the thermal performance of the gasifier.

#### **1.4 Outline of the Research Thesis**

Chapter One outlines the background and the current problems faced by the tea industry in Sri Lanka. It is followed by the research objectives and the limitation of this study.

Chapter Two discusses the tea manufacturing processes, the drying characteristics and factors affecting the tea process. The energy requirements for conventional tea drying process in Sri Lanka is then discussed and the summary of the energy consumption is presented in this chapter.

Chapter Three investigates the available energy resources and possible energy options for tea processing in Sri Lanka to meet the energy demand outlined in the previous chapter. Then the development of local technologies available to harness those resources are discussed in detail. A comparison of the cost and benefit analysis of those technologies for the tea industry in Sri Lanka is also investigated in this chapter.

Chapter Four describes the gasification process in detail. Next, the criteria and review of the available models are described. It is followed by model selection and necessary

improvements to the selected models. Finally the objective, development and implementation of such models are described.

Chapter Five presents the layout of the experimental system used to collect data to calibrate the model developed in Chapter Four. It is followed by the details of the test gasifier and other experimental apparatus. Processing and characterisation of rubber wood are then discussed. Instrumentation and calibration of experimental equipment are also described. Finally the experimental procedure and a summary of the results are presented in this chapter.

Chapter Six begins with a description of verification of the flaming pyrolysis zone sub-model using the results of the theoretical models available in the literature. It is followed by a parametric study to investigate the parameters which effect the gas temperature. Next, the energy and material balances are carried out to determine the heat loss. Then verification and calibration of the gasifier zone model are carried out by using the experimental data given in the literature and in Chapter Five. It is followed by a detailed parametric study of the gasifier developed by the NERD Centre to optimize its operating and design parameters. Then a detailed discussion of the model outputs is described.

Chapter Seven outlines the research findings, limitations of present study and conclusions.

### **1.5 Scope of the Study**

There are several technologies available for tea manufacturing in Sri Lanka but this study will focus on the “Orthodox” or conventional tea manufacturing process. In the tea manufacturing process, 95% of the energy is used for the withering and the drying processes. Hence, the study is limited to these two processes because of their high energy consumption compared to other processes in the tea production. The study is restricted to downdraft wood gasifiers with rubber wood as the feed material. The gasifier is of the type used for thermal energy generation (heat gasifiers) as more than 90% of the energy requirement for tea drying is thermal energy. Cleaning and

cracking of tar in the product gas to use in power generating equipment are beyond the scope of this study as the electrical energy requirement is not considered to be as significant as the thermal energy requirement for tea drying process.

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## **CHAPTER 2**

### **ENERGY UTILIZATION IN THE TEA PROCESS**

#### **2.1 Introduction**

Chapter One outlines the reasons for the high cost of production in the tea industry in Sri Lanka. In order to improve its cost competitiveness and to address the national issues of forest cover and CO<sub>2</sub> emissions, improvements in the use of energy in the tea industry are required. To identify the improvements, it is necessary to discuss the details of energy use in the tea process. This chapter describes the energy used in the tea manufacturing process. In order to understand the energy use of these processes, it is first necessary to discuss the tea manufacturing process. Thus the chapter begins with a discussion of the Orthodox or conventional tea manufacturing technology. Then the characteristics and factors affecting the withering and drying processes are discussed. Although the Orthodox manufacturing technology consists of several processes, only withering and drying processes require thermal energy. Thus the energy requirement for these two processes are reviewed in detail to identify the different process requirements. Finally, a summary of energy consumption is presented.

#### **2.2 Tea Manufacturing Process**

Sri Lanka mainly produces tea by the Orthodox technology. The Orthodox tea manufacturing technology consists of several processes starting from the green leaf up to the final product (“made” tea). These processes are withering, rolling, roll breaking or sifting, fermentation, drying, sifting, grading and packing (Figure 2.1). Since this study is focused on the thermal energy requirement for tea processing, only the withering and drying processes are discussed in the following section.

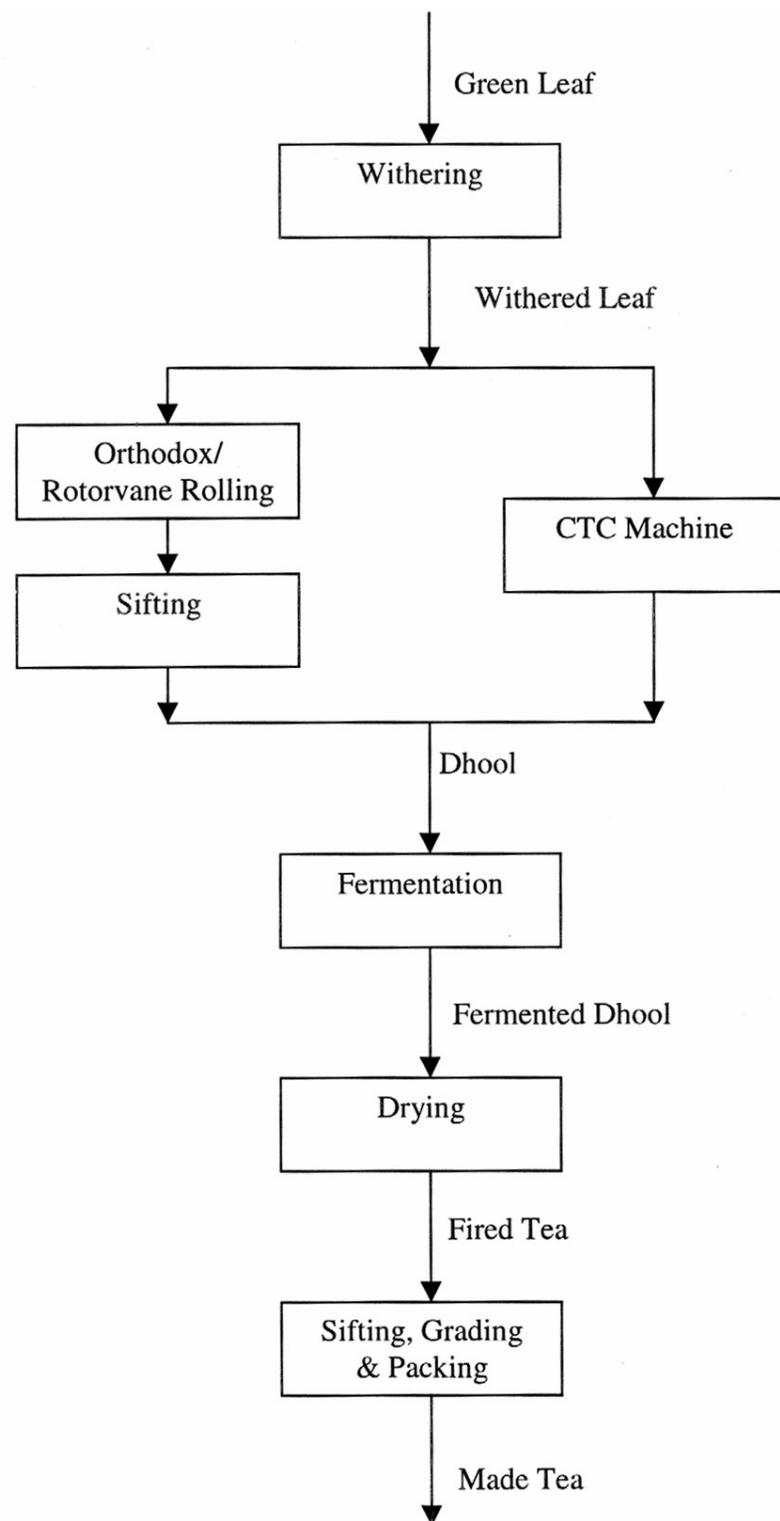


Figure 2.1: Tea manufacturing process

## 2.2.1 Withering

Withering is the extraction of some of the sap of the green leaf under controlled conditions. The green leaves are spread in troughs, measuring from 12 to 36 m in length and 2 m in width, at a rate of approximately 30 kg/m<sup>2</sup>. The moisture content of the green leaf is reduced from 70~80% (w.b.) to 54% ± 1.5% (w.b.) for “up and mid” country manufacture, 59% ± 2% (w.b.) for “low” country manufacture and 70% ± 2% (w.b.) for CTC manufacture (de Silva 1996). A mixture of ambient air and exhaust hot air from the drier is used to meet the hot air requirement for withering. If the weather condition is favourable hot air from the drier is not utilized. The temperature of this air mixture is maintained below 35°C (Riva & Palaniappan 1989). The degree of wither, i.e. the percentage of moisture extracted by weight from the green leaf, varies from 25 to 50% depending on the type of the manufacture, quality of tea and factory elevation (Jayatunge 1986). The moisture content of green leaves depends on the weather conditions, and typical moisture contents of tea during different stages of processing are given in Table 2.1 (Samaraweera 1986). The duration of the withering process depends on the type of manufacture. Withering time for “up and mid” country tea leaves is 12 hrs and for “low” country tea leaves it is 18 hrs (de Silva 1994). When the desired wither has been obtained, the withered leaves are conveyed to the rolling process, followed by the sifting process. The sifted tea leaves, also known as ‘dhool’, are next sent for fermentation. This is followed by the drying process.

Table 2.1: Moisture content of tea at different stages of processing

Type of leaf		Moisture content (% w.b.)
Green leaf	- Wet weather	83
	- Dry weather	70
Withered leaf	- Very soft wither	70
	- Soft wither	58
	- Medium wither	55
Dry leaf	- Made tea	3

(Source: Samaraweera 1986)

### **2.2.2 Drying**

Drying is the extraction of most of the remaining moisture from the dhool. The moisture content of the dhool is reduced from 58% to 3% (w.b.) and this is achieved by using a stream of hot air. Wood or oil is burnt in a furnace and then heat is transferred to the air by a heat exchanger. The temperature of this air depends on the type of the drier in use. Generally two types of driers are used for this process. These are the Endless Chain Pressure (ECP) drier and Fluid Bed (FB) drier of which the former is the most commonly used in Sri Lanka. In ECP driers, perforated trays are used to transport the dhool through a stream of hot air by means of continuous chains. In FB driers, hot air is used to dry and transport the dhool through the chamber. The inlet operating temperatures of ECP drier are 90-120°C and those of FB drier are 120-130°C (Jayatunge 1986). The drying process lasts 15 to 42 minutes (Keegel 1983). Tea emerging from the drier is known as “fired” tea. The “fired” tea is next sent for sifting and packing.

### **2.3 Characteristics and Factors Affecting the Withering and Drying Processes**

Keegel (1983) has shown that the rate of withering is appreciably influenced by withering temperature, air flow rate, period of wither, thickness of spread, condition of the leaf (surface moisture), type of the leaf, atmospheric conditions, etc. Drying is influenced by the inlet temperature of the drier, thickness of the spread of the dhool, airflow rate, period of drying, moisture content of the dhool, etc. Table 2.2 shows the withering and drying characteristics of tea.

Table 2.2: Withering and drying characteristics of tea

Process	Parameter	value
Withering	Initial moisture content of green leaf	70-80% w.b.
	Final moisture content of withered leaf	54-70% w.b.
	Maximum inlet air temperature	35°C
	Period of withering	12-18 hrs
Drying	Initial moisture content of the dhool	66-70% w.b.
	Final moisture content of fired tea	3% w.b.
	Inlet air temperature - ECP Drier	90-120°C
	- FB Drier	120-130°C
	Period of drying	15-42 min

## 2.4 Energy Requirement for Tea Processing

Both withering and drying processes require the input of thermal energy. For the withering process, in practice, the thermal and electrical energy requirements per kilogram of “made” tea are 9.0 MJ and 0.46 kWh respectively, whereas for the drying process, the thermal and electrical energy requirements per kilogram of “made” tea are 13 to 13.7 MJ and 0.07 kWh respectively (de Silva 1994). Table 2.3 provides a summary of the energy consumption in the tea process in Sri Lanka.

Table 2.3: Summary of energy consumption in practice in the tea process

Process	Energy consumption/kg of “made” tea	
	Thermal (MJ)	Electrical (kWh)
Withering	9.0	0.46
Rolling	-	0.10-0.20
Drying	13.0-13.7	0.07
Sorting, grading & packing	-	0.07-0.09
Total	22.0-22.7	0.70-0.82

(Source: de Silva 1994)

## 2.5 Fuel Wood and Oil Consumption

According to de Silva (1996) if only fuel wood is used for tea processing, the wood demand varies from 1.8 to 2.2 kg fuel wood/kg of made tea for fuel wood moisture contents of 25 and 35% respectively. Thus, on average for every kilogram of tea produced two kilograms of fuel wood is used for withering and drying in Sri Lanka, assuming a calorific value of 11.5 MJ/kg at 30% moisture content. Most of tea factories in Sri Lanka depend on fuel wood for thermal energy requirements, i.e. 85% of thermal energy is produced from fuel wood and 15% from oil (de Silva 1996). Thus for a tea production of  $306 \times 10^3$  tonnes, the annual fuel wood consumption for the year 2000 by the tea industry in Sri Lanka will be  $520 \times 10^3$  tonnes. Using the assumptions, the annual fuel wood consumption for 1985, 1992 and 1993 have been calculated and compared with the figures quoted in the literature (Table 2.4). If an average value of 22.4 MJ of thermal energy and 0.76 kWh of electrical energy per kilogram of “made” tea are then assumed, the energy resources consumed by the tea industry in Sri Lanka in 2000 would be as shown in Table 2.5.

Table 2.4: The annual fuel wood consumption by the Sri Lankan tea industry

Year	Fuel wood consumption (tonnes)		Reference
	Estimated	Quoted by others	
1985	363,800	373,890	Haskoning (1989)
1992	304,300	317,200	Tariq & Purvis (1997)
1993	394,400	406,000	Kumaradasa et al. (1999)
2000	520,200	-	-

Table 2.5: The energy consumption by the Sri Lankan tea industry in 2000

Source of Energy	Quantity
Fuel wood (thermal energy)	5982 TJ
Diesel oil (thermal energy)	1056 TJ
Hydro power (electrical energy)	233 GWh

## 2.6 Conclusions

It can be seen from Table 2.3 that to produce a kilogram of “made” tea in Sri Lanka, an average of 22.4 MJ of thermal energy and 0.76 kWh of electrical energy are needed. Ninety percent of the total energy requirement is thermal energy of which 85% is met by burning fuel wood. Therefore, in order to see if there are any variable substitutes or alternatives for this resource consumption, it is necessary to investigate the other available resources and evaluate their potential. The following chapter investigates the available energy resources in Sri Lanka to meet the energy demand outlined in this chapter. Then possible energy options and feasible technologies in order to harness those resources for tea process are discussed. It is followed by a life cycle energy cost of the feasible technologies.

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## **CHAPTER 3**

### **ENERGY OPTIONS FOR THE TEA PROCESS**

#### **3.1 Introduction**

The previous chapter outlines the energy needs of the tea industry in Sri Lanka. It further emphasises the necessity for a detailed study on the available energy resources to meet such energy needs. This chapter discusses all the energy options in terms of resources, potential and the quality of energy that can be produced by each of these resources. Then, the developments of local technologies available to harness those resources for tea processing are discussed in detail. A cost and benefit analysis for the available technologies are also carried out to assess their financial feasibility.

The quantity, quality of the energy required and the cost are the main criteria in selecting appropriate technologies. Only the resources which can contribute more than half of the annual energy demand are considered to be viable options for the tea industry as a whole. This will help establish the feasibility of incorporating locally available technologies with local energy resources to meet such energy demand of the industry. Based on the lowest life cycle energy cost, the most appropriate technology is identified for the tea drying process in Sri Lanka.

#### **3.2 Energy Resources of Sri Lanka**

There are three major sources of energy available in Sri Lanka. The two main indigenous sources of energy are hydropower, which is used almost exclusively for electricity generation, and biomass, which is generally used for thermal applications. The third source is petroleum products, which are totally imported. In 1992, the consumption of biomass, petroleum products and electricity accounted for 66.2%,

22.4%, and 11.4% of the gross energy consumption in the country respectively (Tariq & Purvis 1997).

Biomass sources are site specific and each source will have to be examined to determine its potential for utilization by the tea industry. Factors such as availability throughout the year, the amount of transformation such as collecting, shredding, transporting etc., required to convert the raw biomass into suitable forms for energy production, and the effect of these operations on local ecology and environment should also be given serious consideration. The following section discusses the potential of available biomass resources in detail and it is followed by other resources such as hydro power, solar energy, wind energy and fossil fuel.

### **3.2.1 Fuel wood**

Fuel wood is the main source of energy for cooking and heating in domestic households and for process heating in industries. In 1993, the share of biomass as compared to the total energy consumption was about 66% of which fuel wood accounted for 88% of the total biomass consumption (Kumaradasa et al. 1999). Approximately 56% of biomass comes from home gardens, crop lands and forests. The contribution of rubber plantations to the biomass supply is 7% and the remaining 37% comes from wastes mainly from coconut plantations and to a certain extent from tea plantations, and from paddy husks and straw (Kumaradasa et al. 1999). In Sri Lanka, the annual fuel wood potential is around  $16,500 \times 10^3$  tonnes (Howes & Endagama 1995).

### **3.2.2 Paddy husk**

Paddy is grown in nearly all regions of the island. In many areas there are two harvests a year. In 2000, paddy production reached  $2.86 \times 10^6$  tonnes (Central Bank of Sri Lanka 2001). Paddy husk and straw are by-products which have a potential use as a fuel. According to the FMP (1986), one kilogram of paddy provides 0.22 kg of

air-dried paddy husk. This gives a paddy husk potential of about  $629 \times 10^3$  tonnes per annum. But costs involved in processing paddy husk into transportable forms and fuel conversion are critical for large-scale applications. Paddy husks are available in the paddy growing areas, which are quite distant from the tea plantations. Thus, transportation cost will be vital. Straw is another under-utilized by-product of paddy. Despite its potential as a resource, it is not often used as fuel due to its high moisture content and low bulk density. But it is used as animal feed or as fertilizer in fields.

### **3.2.3 Coir dust**

Coir dust is another source of biomass available in abundance. It is an under-utilized waste product in the coconut industry and there is difficulty in disposing of this waste. Haskoning (1989) stated the annual production of coir dust is estimated to be  $140 \times 10^3$  tonnes with a moisture content of 20% (w.b.). Kumaradasa et al. (1999) has quoted the figure of  $750 \times 10^3$  tonnes with a moisture content of 85% (w.b.) Coir stocks are available in the coconut plantation areas and the distance between the tea and coconut plantation areas is more significant than the distance between the tea and paddy growing areas. Thus, the cost involved in converting coir dust into a transportable form such as briquettes and the actual transport cost is likely to be higher than that of rice husks. Haskoning (1989) stated that the main problem with coir dust is in the conversion technique necessary to dry the wet coir dust before briquetting takes place.

### **3.2.4 Annual biomass potential**

The annual fuel wood, paddy husk and coir dust potential in Sri Lanka is summarised in Table 3.1.

Table 3.1: The annual potential of biomass in Sri Lanka

Source	Quantity (Tonnes×10 <sup>3</sup> )	Moisture Content (% w.b.)	Gross Calorific Value (MJ/kg)	Potential (PJ/year)
Fuel Wood	16,500	35	10.4	171.60
Paddy Husk	629	15	14.4	9.06
Coir dust	140	20	12.6	1.76

### 3.2.5 Hydro power

Sri Lanka is a country rich in hydro power resources. According to recent statistics, the total installed capacity of hydro plants is 1142 MW, and this is augmented by thermal plants with a capacity of 606 MW (ColomboPage 2001). Generally the electricity demand in the country is met by hydropower. Thermal power plants are used only when the water capacities in the reservoirs are inadequate to meet the demand. Up to 1990, 99.8 per cent of the electricity demand in the country was met by hydro power (de Silva 1994). It is estimated that the hydro power potential in the country is approximately 2000 MW of which more than a half has already been harnessed (Daily News 1997). Further expansion of hydro resources may be difficult due to the impact on the environment and the eco-system of the country.

Although almost all the present electricity demand in the country is met by hydro power, the contribution of micro-hydro power to the national demand is insignificant. During the inception of the tea industry, many micro hydro units were installed at the tea factories to meet the electrical energy demand but after the national power supply grid was made available to these tea factories, the utilization of micro hydro was neglected. Haskoning (1989) stated that in 1984 a total of 140 mini hydro schemes were identified having the potential to generate between 50 to 200 kW of electrical energy in Sri Lanka. Some rehabilitation work on micro hydro has been carried out, but only a few are reported to be economically viable (Haskoning 1989).

Micro hydro power has the capability of producing electrical energy as well as mechanical energy for the tea industry, but it is necessary to consider its competitiveness with the present electricity prices. The power that can be generated by a micro hydro systems depends on the head and the flow range of the streams. As with other forms of renewable energy, it is essential to ascertain the variations in the energy source. This means measuring the flow variations during the course of the year. A consistent flow will produce steady power. But the water flows in the streams are seasonal and during the drought period, which lasts for approximately six months in Sri Lanka, power generation would be greatly reduced. A significant head of water is also essential for power generation.

### **3.2.6 Solar**

Sri Lanka is a tropical country where plenty of sunshine is available throughout the year. Tea plantation areas receive an annual sunshine ranging from 1400 hours in Nuwara Eliya to 2300 hours in Kandy (Arachchi 1997). The monthly mean daily global radiation over Sri Lanka is between 15 and 20 MJ/m<sup>2</sup>day (Samuel 1991). Solar energy can be used for thermal power applications but the quality and quantity of energy that can be produced by solar technology will need to be considered when assessing the viability of solar systems for the tea industry.

### **3.2.7 Wind**

The potential for wind energy is high in some parts of Sri Lanka. Sivasegaram (1983) has quoted from Fernando and Smulders (1976) that in Sri Lanka the wind speeds have an annual average value ranging from 2.2 m/s to 5.5 m/s. The higher values occur at the coastal areas like Hambantota, Trincomalee, Jaffna etc., whereas tea plantation regions are confined to the central hilly areas. Haskoning (1989) stated that although high wind speeds are also reported in the central hills, these winds blow only during a relatively short time of the season. He further stated that the locations of most of the tea estates are not the most viable places to install wind turbines.

Wind energy is more variable than solar radiation and mean wind speeds are highly site specific. The energy content of the wind is related to the cube of the wind speed, so site selection based on long-term wind speed data is essential. According to Sivasegaram (1983) no systematic study of the variation of wind velocity in Sri Lanka has been carried out. This is a major problem with respect to the use of wind energy, because wind speed data are essential to analyze the financial feasibility of the system. The higher the wind speed, the lower the cost of generating electrical energy will be (Krohn 1997). Thus, wind speed data needs to be collected particularly in the tea plantation areas on a regular basis to assess the potential for wind energy utilization.

### **3.2.8 Fossil fuel**

Fossil fuel resources in Sri Lanka are very limited. Wijesinghe (1988) has quoted from Ekono (1985) that a surface deposit of peat has been found in the Western coastal area with a thickness of one metre over an area of 240 ha. Apart from this, Sri Lanka does not have fossil fuel resources. All petroleum products are imported to meet the energy demand for the transport, industrial and domestic sectors. The motorized transport sector is almost totally dependent on petroleum products and kerosene is used mainly by the domestic sector for lighting.

### **3.3 Technologies and the Energy Costs**

The following section discusses the available technologies to harness the energy resources discussed in the previous section and a life cycle energy cost analysis is carried out in order to compare the economic feasibility of these systems for the tea industry. It is followed by a comparison of the energy cost and the selection of the technology based on lowest energy cost.

### **3.3.1 Wood combustion**

Fuel wood is the main source of energy for the tea industry and wood furnaces are being used in almost all the tea factories in the country to produce necessary hot air for withering and drying processes. But the main problems associated with wood furnaces are their low efficiency. A recent study carried out by Tariq and Purvis (1997) has reported that the average thermal efficiency of furnaces used in Sri Lanka is 50%. The study further stated that the low efficiency is due to poor maintenance of combustion plants, use of plants beyond its design capacity, use of fuel wood with high moisture content and lack of training and understanding of good combustion practice by the operators. A locally fabricated wood fired air heater of 15 years lifetime will cost US\$15,400 (Ziyad Mohamed 1998). If a plant thermal efficiency of 50%, an operating and maintenance cost of 10% of the capital cost, inflation rate of 10% for operating and maintenance and 5% for fuel wood and a market discount rate of 12% are assumed, then this gives a life cycle energy cost of US\$3.25/GJ. Details of this calculation are given in Appendix A. Any substitution for the traditional air heaters should be competitive with this energy cost.

### **3.3.2 Wood gasification**

Besides the traditional wood fired air heater system, a wood gasifier is another option for thermal energy production. Wood gasification can be used in the tea industry to produce the required thermal and electrical energy. A system with a wood processor, gas cleaning and cooling unit and gas engine-cum-generator will produce the energy necessary for the industry. Although this is an efficient way of using fuel wood, recent studies on biomass gasification have shown that power gasifiers are not economically viable but the commercial potential for heat gasifiers is significant and they are economically attractive (Stassen 1995). Stassen (1995) stated that in developing countries, the prospects of biomass power gasifiers appeared to be limited due to unfavourable economics compared to fossil fuels, operational difficulties resulting from low reliability and quality of equipment and difficulties in training sufficiently qualified personnel. Long term prospects depend on price increases in the world oil

market, and progress made in improving the quality of the equipment and simplifying operating procedures.

A locally fabricated wood gasifier of 15 years lifetime will cost US\$23,000 (Ziyad Mohamed 1998). If an operating and maintenance cost of 10% of the capital cost, inflation rate of 10% for operating and maintenance and 5% for fuel and a market discount rate of 12% are assumed, then this will give a life cycle energy cost of US\$2.93/GJ. This technology also saves 0.30 kg of fuel wood for every kilogram of tea produced, which corresponds to an annual fuel wood saving of 78,500 tonnes (Ziyad Mohamed 1998). Details of the calculation are given in the Appendix A.

### **3.3.3 Paddy husk combustion**

Generally, paddy husks are utilized by direct combustion. In direct combustion paddy husks are burnt in chambers to produce hot water, steam and hot air either with or without a heat exchanger. The potential for utilizing paddy husks to meet the hot air requirement in the tea drying and withering processes is significant provided the cost involved in the conversion of the fuel into a suitable form (like briquettes) for energy production is comparable with that of the price of fuel wood.

Lawand (1993) stated that the major problems associated with paddy husks are the high ash content and the difficulty of burning them in a combustion chamber. The moisture content of the paddy husks is also important because the lower the moisture content, the higher the combustion efficiency will be. Transport cost is another vital factor in estimating the overall cost. Thus, the potential for the utilization of paddy husks in the tea industry depends on factors like quantity, combustibility, moisture content, its availability throughout the year, storage and handling of fuel, operation and maintenance of the furnace, disposal of ash and residue and finally the economic viability of the system.

A project, undertaken in 1989 to assess the economic feasibility of biomass for industries has shown that low-density briquettes can be produced at ex-factory price of Sri Lankan rupees (Rs) 216/tonne at a calorific value of 14.4 MJ/kg (Haskoning

1989). The following equation was used to calculate the transport cost (Haskoning 1989).

$$\text{Transport cost (Rs/GJ)} = 1.79 + 0.429D$$

Where D is the distance in km. If a distance of 100 km is considered, then the total energy cost will be Rs. 45/GJ, which is equivalent to Rs. 648/tonne. This gives the total paddy husk briquette cost of Rs. 864/tonne. Over the past 11 years the average annual inflation in Sri Lanka was 9.7% (Central Bank of Sri Lanka 2001). Thus the cost of paddy husk briquettes for the year 2000 will be Rs. 2392/tonne, which is equivalent to US\$ 26.6/tonne for an average exchange rate of Rs. 90 = US\$1. If the same air heater used for wood combustion is used for paddy husk with the same assumptions, then this will give a life cycle energy cost of US\$3.00/GJ. Although this is a good option to replace fuel wood the availability of paddy husk in the tea growing area is limited to only about 61,000 tonnes per annum (Haskoning 1989). With a calorific value of 14.4 MJ/kg, the annual energy potential will be 878 TJ, which is about 15% of the annual thermal energy demand of the tea industry. Therefore it is not worth investing in a new technology which can meet only 15% of the annual energy demand at its best.

### **3.3.4 Coir dust combustion**

From the project undertaken to assess the economic feasibility of biomass for industries, Haskoning (1989) has shown that the cost of producing low-density coir dust briquettes was Rs. 367/tonne at a calorific value of 12.6 MJ/kg. If the same equation used for paddy husk is used to calculate the transport cost, then for an average distance of 150 km, the total cost of coir dust briquettes will be Rs. 1200/tonne. For the same average annual inflation rate (9.7%) and exchange rate (Rs. 90 = US\$1), the cost of coir dust briquettes for the year 2000 will be US\$36.9/tonne. If the same air heater used for wood combustion is used for coir dust as well with the same assumptions, then this will give a life cycle energy cost of US\$4.44/GJ.

### 3.3.5 Solar technology

Several studies have been carried out to assess the technical suitability of solar technology for the tea process. Such a study carried out by the Netherlands-Sri Lanka Energy Programme has shown that a thermal energy production of 5250 MJ/day from 700 m<sup>2</sup> of solar collectors generating a temperature of 70°C from an irradiation of 15 MJ/m<sup>2</sup> day is possible in Sri Lanka (Meel 1988). For their study, efficiencies of 50% for the collectors and 80% for the storage system were assumed. The cost of the system was US\$185/m<sup>2</sup> of the collector area. Another similar study carried out by a leading international construction company, has shown that 69% of the thermal energy requirement for tea processing in Sri Lanka could be met by solar (Millin 1993). They have shown that a collector area of 280 m<sup>2</sup> would be needed for a single drier capable of removing 185 kg of water per hour to give 100 kg of “made” tea per hour at an initial moisture content of 66% (w.b.). The cost of the system was US\$135/m<sup>2</sup> of the collector area and the life cycle cost of generating thermal energy was US\$5.6/GJ.

Kumar and Palaniappan (1994) stated that the use of high quality fuels like coal or oil for generating hot air between 60° and 140°C is wasteful and solar air collectors can be used for the drying process. They further stated that solar air collectors have been used for tea processing in India, and have saved conventional fuels without degrading the quality of the product. For temperatures below 100°C, flat plate collectors have been recommended. Since the temperature of the air required for drying is between 90° and 120°C, flat plate collector technology is not an ideal choice, because efficiency declines with increasing temperature. However, flat plat collector technology can be used to preheat the air to reduce the fuel requirements.

If flat plat collectors with a system cost of US\$175/m<sup>2</sup> of the collector area, efficiency of 40% and a lifetime of 20 years are considered, for an irradiation of 15 MJ/m<sup>2</sup> day, the life cycle thermal energy cost will be US\$5.8/GJ. For this calculation, an operating and maintenance cost of 3% of capital cost, an inflation rate of 10% and a market discount rate of 12% are assumed. Details of the calculation are given in the Appendix A. Assuming the above solar radiation level and collector efficiency, a solar

system with a collector area of 400 m<sup>2</sup> will produce 2400 MJ per day. This option would replace around 62,500 tonnes of fuel wood per annum.

### **3.3.6 Wind technology**

Wind energy technology is the most mature of renewable energy technologies currently available and has been in use to generate mechanical as well as electrical power. Although this technology is not widely used in Sri Lanka, several attempts have been made to produce electrical power in the coastal areas where the wind potential is high. A wind farm is in operation in Hambantota with a plant capacity of 3 MW (ColomboPage 2001).

The daily electrical energy requirement of a factory producing 2000 kg of “made” tea is 1520 kWh. According to Krohn (1997) a modern wind turbine with a rotor diameter of 43m and a 700 kW generator is capable of providing an average of 1600 kWh/day at a mean wind speed of 4.5 m/s. The cost of such system would be US\$300,000 excluding installation (Krohn 1997). A locally manufactured efficient wood fired heater, which provides 90% of the energy requirement for tea processing, costs approximately US \$ 40,000 in 1988 (Haskoning 1989). Thus, the high cost of wind turbines is unlikely to make them acceptable for local tea industries unless special incentives are granted.

Wind turbines are designed to generate electrical power. Although thermal energy can be generated using electrical power, the cost of generating electricity by wind turbines itself is high. Krohn (1997) stated that the cost of generating one kWh of electricity is US\$0.05 for a turbine at a mean wind speed of 4.5 m/s. If an operating and maintenance cost of 10% of the capital cost, inflation rate of 10% for operating and maintenance and 5% for fuel, a market discount rate of 12% and the cost of US\$15,000 for a heating system with 95% are assumed, then this will give a life cycle energy cost of US\$10.09/GJ (Appendix A).

### 3.3.7 Micro hydro power

If micro hydro schemes are to be introduced to the tea industry, factors such as site evaluation, initial cost of the equipment, payback period, maintenance cost, percentage of the contribution of micro-hydro power to the tea industry etc., should be considered. In Sri Lanka, the present electricity supply price stands at US\$ 0.035/kWh for industrial sectors. If an operating and maintenance cost of 10% of the capital cost, inflation rate of 10% for operating and maintenance and 5% for fuel, a market discount rate of 12% and the cost of US\$15,000 for a heating system with 95% are assumed, then this will give a life cycle energy cost of US\$7.14/GJ (Appendix A). The cost of producing electrical power using micro hydro is assumed to be same as present electricity price.

### 3.4 Discussion

Table 3.2 provides the life cycle cost of thermal energy production by different technologies.

Table 3.2: Life cycle cost of thermal energy production by different technologies

Technology	Cost (US\$/GJ)
Wood Gasifier	2.93
Paddy Husk Air Heaters	3.00
Wood Fired Air Heater	3.25
Coir Dust Air Heaters	4.44
Solar Collectors	5.80
Micro Hydro Powered Electric Air Heaters	7.14
Wind Turbines Powered Electric Air Heaters	10.09

It is evident from Table 3.2 that of all the renewable energy resources, burning of biomass either in a gasifier or conventional reactor is the cheapest source of energy

for the tea industry in Sri Lanka. Moreover, wood gasification appears to be a better option to the traditional air heaters as it is cheaper and can reduce the annual fuel wood consumption by as much as 15%. Paddy husk is next best option, but its limited availability in the tea growing areas, high ash content, and difficulty to burn in combustion chambers should be considered seriously before any decision is taken. Coir dust does not seem to be competitive with fuel wood due to high transport cost. The main problem in coir dust is in the conversion technique necessary to dry the wet coir dust with a high moisture content (as high as 85% on wet basis), before briquetting takes place.

Solar energy is another option to replace the traditional wood fired air heaters. Although it has the potential to reduce the greenhouse gas emission and save the valuable forest cover in the country, the solar energy system considered is twice as expensive as the wood gasifier option and also needs a large area. It could be financially viable for the withering process as the air temperature requirement is below 35°C. But this is not a critical issue for a tropical country like Sri Lanka where the average daily temperature lies around 25°C. Most of the time ambient air is used for withering. Another draw back in the solar system considered is its high initial cost. The initial high cost will not make solar energy attractive to the industry unless the fuel wood prices increase considerably or the local government grants special incentives for solar systems. Micro hydro power and wind turbines are also not attractive alternatives because of their high cost for heat generation.

### **3.5 Development of Gasifier Technology for Tea Industry**

Although feasibility studies have shown that the gasifiers can be technically and economically viable for developing countries like Sri Lanka, their acceptance by the local industries depends primarily on the reliability of the system. Most of the locally fabricated gasifier systems failed to attract the attention of the local industry because they did not perform consistently at their rated capacity. This could be due to their lack of knowledge and skill necessary to design such systems and the lack of enthusiastic personnel in this field. Lack of necessary funds for research and

development work is also another factor that prevents the development of this technology in Sri Lanka. Although several institutions have taken a keen interest to explore the possibility of enhancing this technology, progress has been slow.

The National Engineering Research and Development (NERD) Centre Sri Lanka is the pioneer research institution conducting extensive research on gasifiers in the country with the objective of introducing this technology to local industries. Research on gasifiers at this Centre was initiated in 1978. By 1999, 75 gasifiers of various capacities and types had been designed, fabricated and commissioned (Weerarathna 1999). Most of them were the downdraft type. Initially the gasifiers were used for power generation and vehicles, and later for crematoria and foundry work. However they are still in the research stage for drying of agricultural products. Their use has been a success in crematoria where 32 gasifiers have already been installed and commissioned.

The first gasifier fabricated for tea drying by the NERD Centre was in 1987 and a second gasifier was commissioned a year later. However those early attempts were not successful. There were several technical problems associated with them such as high heat losses from the gasifier body, gas and air leakage from the system, fast deterioration of materials used in the hearth zone, clogging of wood particles above the throat and severe corrosion of the inner wall of the body (Weerarathna 1999).

In 1988, in order to address the problem of the inefficient use of biomass in the tea drying process, the Tea Research Institute (TRI) of Sri Lanka focused research on biomass gasifiers to replace the traditional furnace system. As a result, a system consisting of a downdraft wood gasifier, a gas burner and a gas engine-cum-generator, fabricated by BECE of the Netherlands was tested between February 1988 to September 1990. The gasifier was designed to convert approximately 190 kg of fuel wood per hour into 440 m<sup>3</sup> of gas at 20°C with a lower heating value of 5 MJ/m<sup>3</sup> at maximum capacity. After cleaning, this gas was used in an internal combustion engine for power generation. The unit has a rated capacity of 500 kW thermal or 60 kW electrical power (de Silva 1994). The conversion efficiency (thermal power) of

the system was 83% for fuel wood at 17% moisture content (w.b.). During testing a fuel wood saving of 0.23 kg/kg of “made” tea was achieved.

Since the gasifier system fabricated by the BECE of the Netherlands was too expensive and complicated to operate and maintain (Ziyad Mohamed 1998), the NERD Centre with the collaboration of the TRI, fabricated a downdraft heat gasifier with similar capacity but using local technology and materials. The gasifier was installed in a tea factory in the TRI and testing began in 1994. Although the conversion efficiency (thermal power) of the system was not reported, a fuel wood saving of 0.30 kg/kg of “made” tea was apparently achieved during testing (Ziyad Mohamed 1998), but detailed test data were not available.

### **3.6 Conclusions**

When considering the financial feasibility, it can be concluded that wood gasification appears to be the best option and has the potential to replace the traditional wood fired furnace systems in the tea industry in Sri Lanka. It can also reduce the annual fuel wood consumption by 15% which equates to an annual forest cover saving of 416 ha at the rate of 15 tonnes  $\times$  25 years / ha and an annual reduction in emissions of 292,000 tonnes at the rate of 1.87 tonnes of emissions per tonne of fuel wood. Of the two gasifier systems discussed in this chapter, the NERD gasifier has reported a higher fuel wood saving of 0.30 kg/kg of “made” tea whereas the BECE gasifier has reported a saving of 0.23 kg/kg of “made” tea. The BECE gasifier system has a higher conversion efficiency of 83% but detailed information is not available to compare with the NERD system. Although the NERD gasifier system appears to be a better option, problems associated with the refractory lining used in the throat, gas leakage and the lack of knowledge of optimum design and desirable operating input parameters raised doubts of its applicability and productivity for the tea industry. Hence there is a need to investigate the performance of the local gasifier to improve its output, construction and use of better materials. A convenient way to investigate the gasifier is to make use of a computer simulations which enable the gasifier performance to be studied. The following chapter discusses the gasification process,

reviews the available gasifier models, and describes model selection and development to study the effects of operating parameters such as air-fuel ratio, chip size, inlet air temperature and fuel moisture content, and design parameters such as throat angle and level insulation on conversion efficiency of the gasifier.

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## **CHAPTER 4**

### **GASIFIER MODEL**

#### **4.1 Introduction**

The previous chapter discusses the energy resources and the technologies available to meet the energy demand of the tea industry in Sri Lanka. Gasifier technology using fuel wood appeared to be a better option compared to the traditional wood fired furnace and other technologies. Chapter Three also discusses the problems encountered with the local wood gasifier and then established the need to study the gasifier in detail as a part of the attempt to improve its output. Since the experimental approach is expensive, computer simulations are used for the investigation. This chapter reviews the existing gasifier models to see if any of them are suitable for this study. However in order to assist in the understanding and selection of a suitable model, the basic chemical kinetics of each gasification process based on the operation of a downdraft gasifier are initially described. The available models are next classified. The suitability of the models is reviewed against a set of criteria, which are also outlined in this chapter. These criteria were established based on the objective of this study and the importance of each zone in the model development. Then, the modification and development of the computer program are described in this chapter. Finally based on the available models, an appropriate computer code is developed which can be used to predict the conversion efficiency of the downdraft gasifier in terms of its operating and design parameters.

#### **4.2 Gasification Process**

Gasification is defined as the thermal degradation of carbonaceous materials in the presence of less air than the theoretical requirement for complete combustion. It consists of a variety of physical and chemical processes, and rate controlling

mechanisms. The location of these chemical processes depends on the type of the gasifier. The main three types are fixed bed, fluidized bed and entrained suspension flow. Of the three types, the fixed bed is the most common. Since this study focuses on downdraft gasifiers, the basic chemical kinetics of each gasification process based on the operation of a downdraft gasifier are described. The four main processes occurring in a gasifier are drying (preheating), pyrolysis, combustion (oxidation) and gasification (reduction). The schematic diagram of a typical downdraft gasifier with four main zones is shown in Figure 4.1

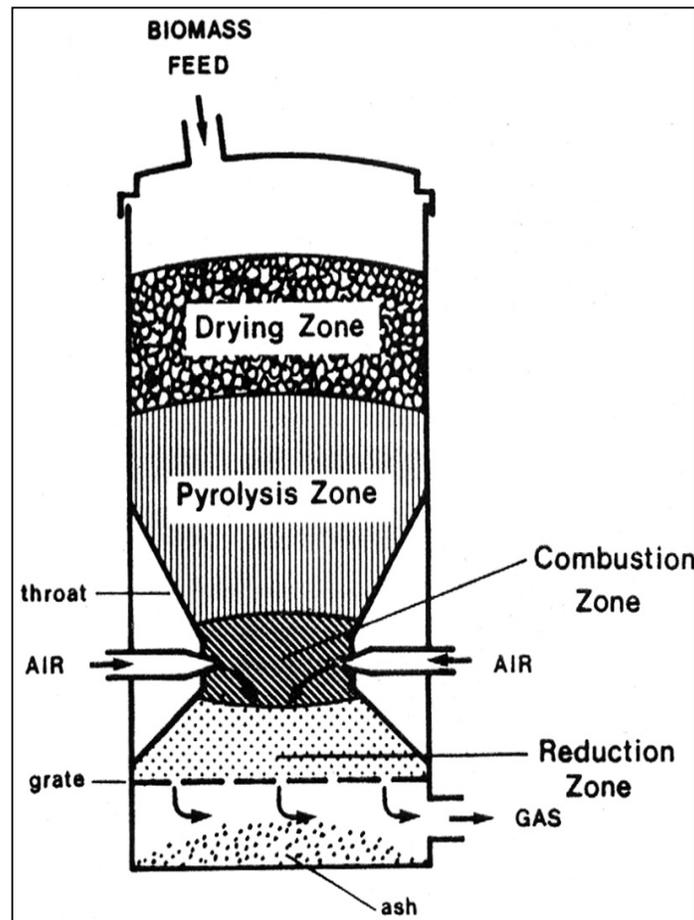


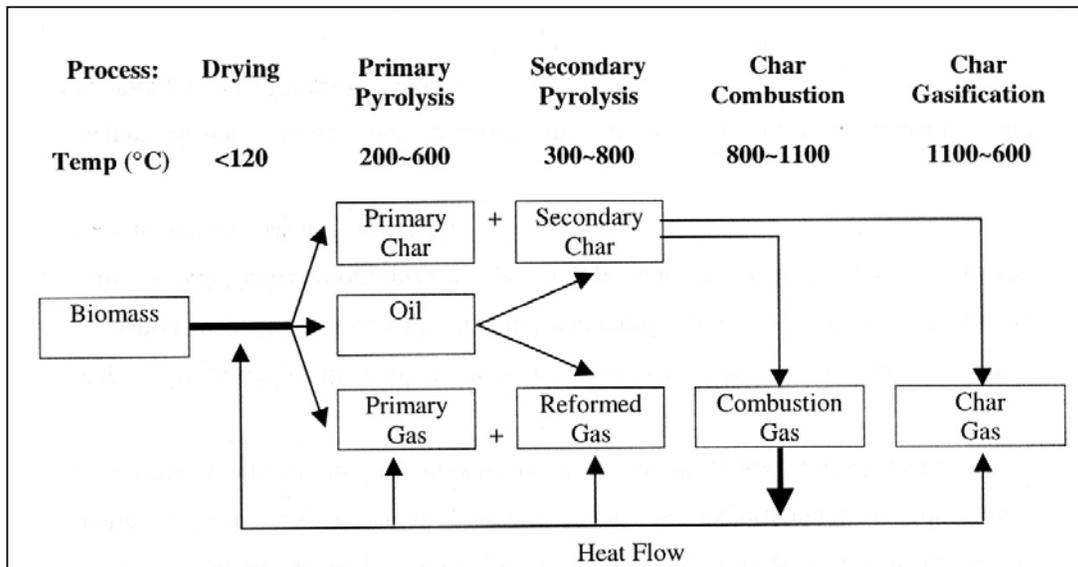
Figure 4.1: Schematic diagram of a downdraft gasifier

Source: Skov and Papworth (1974)

Each of these processes has its own energy requirements and they can be exothermic (energy releasing) or endothermic (energy absorbing). The main rate control mechanisms in the gasifier are heat transfer, mass transfer, chemical reaction kinetics

and pore diffusion. The heat and mass flows in the gasification process are shown in Figure 4.2. These mechanisms will vary in magnitude in each zone and the variations are caused primarily because of the different chemical and physical characteristics of the zones. Some of these characteristics include temperatures of gas and solid phases, air moisture, mass flow rate of air and gas, chip size, moisture content, feed rate, heat losses etc. The following section describes the processes in each zone of a gasifier.

Figure 4.2: Heat and mass flows in gasification processes



Source: Reed (1981)

#### 4.2.1 Drying

The drying zone lies at the top of the reactor and at this point fuel wood is fed into the gasifier. Wood then descends through the gasifier as a result of the consumption of wood particles in the gasification zone and there is no air movement through this region. The main function of this zone is to drive off the moisture in the form of water vapour from the wood. The heat necessary for drying is drawn from the adjacent pyrolysis and combustion zones. The rate at which drying takes place depends on the surface area, temperature and the internal diffusivity of wood particles, and the temperature, velocity and relative humidity of the surrounding air (Buekens & Schoeters 1985). The rate is governed by the external mass transfer as long as the

external surface is wet and by internal mass transfer when the former condition is no longer valid.

Drying occurs below 120°C and the necessary heat is transferred by conduction through wood particles. This heat transfer is not fast enough to remove all the water inside the particle in the drying zone due to the low thermal conductivity of wood which is in the range of 0.006 - 0.011 W/m K (Graboski & Bain 1981). Larger pieces can be burning on outside while there is still moisture inside. Thus drying is not confined to this region and it also occurs in the primary pyrolysis zone. Drying may be instantaneous for small particles exposed to high temperature as in a fluidized bed but it is generally a gradual process for moving bed gasifiers where large particles are descending at a slow pace. During drying, the variation of physical aspects and dimensions of wood particles can be neglected although some shrinking occurs (Buekens & Schoeters 1984). After drying, the dried fuel wood then descends to the pyrolysis zone.

#### **4.2.2 Pyrolysis**

The pyrolysis zone lies below the drying region. As in the case of drying, no air is admitted in this zone but it draws heat from the nearby hot combustion region. When fuel wood is heated to 200°- 600°C in the absence of air, the wood initially breaks down into oil vapour (tar), char (primary char) and gas (primary volatiles). The release of volatiles begins at 250°C (Reed & Das 1988) and this process is known as primary pyrolysis. Oil vapour generated near the surface of small particles can escape into the gas phase before being cracked into secondary char (Reed 1981). But these oil vapours are cracked at high temperatures (above 600°C) to form reformed gas such as hydrocarbons, mainly methane, and this process is called secondary pyrolysis.

For larger particles, the longer escape path provides more time for cracking the oil vapours thus resulting in higher char production. Pyrolysis converts 80 to 95% of original mass into liquid phase products such as water, tars and oils, and gaseous phase products including, CO, CO<sub>2</sub>, H<sub>2</sub> and hydrocarbons, leaving 5 to 20% of highly

reactive char (Reed & Markson 1983). The product composition and distribution of the pyrolysis process depends on factors such as the composition of the feed, heating rate, temperature, residence time of gaseous components and particle size (Buekens & Schoeters 1985). The production of char depends on the heating rate. The lower the heating rate, the higher the char production will be. At a low heating rate (50°C/min) the slow breakdown reaction first splits off water and carbon dioxide and makes the carbon content in the solid higher than in the feedstock. Further decomposition of carbon rich solid at higher temperature results in a relatively high fraction of char which is assumed to be pure carbon. The char produced in the pyrolysis zone then descends into the combustion zone.

#### **4.2.3 Combustion**

Below the pyrolysis region lies the combustion zone through which air is fed into the gasifier. Oxygen in input air reacts with char produced in the pyrolysis zone and produces combustion gas (CO<sub>2</sub>) and water vapour. Combustion of char is very rapid and the reaction is exothermic which results in a rapid rise in temperature. The temperature in this zone varies from 800°-1100°C (Reed 1981). Heat produced from char combustion is the main heat supplier to the other regions of the gasifier. The hot combustion gas and water vapour produced in the combustion zone are next drawn into the gasification zone.

#### **4.2.4 Gasification**

This zone lies at the bottom of the downdraft gasifier. The mechanism of the reaction between char and the surrounding gases can be described as follows. The surrounding gases such as carbon dioxide and water vapour initially react with the char particles at their outer surface. Then the zone of reaction moves into the solid. This is known as diffusion of gas down the pore towards the centre of the particle. Then absorption of these gases and the surface reaction takes place. Finally desorption of product gases such as carbon monoxide and hydrogen takes place on the pore wall. These product

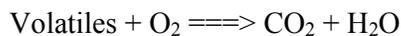
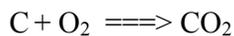
gases are also known as synthesis gas. A portion of the hydrogen combines with carbon to form methane. These reactions are endothermic and they cause the temperature to decline from 1500° to 600°C in the gasification zone.

#### 4.2.5 Chemistry of gasifier process

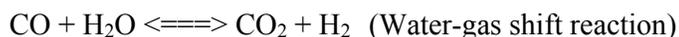
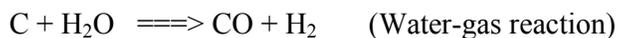
As described in the previous section, there is no chemical reaction occurring in the drying zone due to the prevailing low temperature. However the higher temperature in the pyrolysis zone causes the wood to degrade. The thermal conversion process for pyrolysis can be written as follows (Reed & Das 1988).



The main reaction to occur in the combustion zone is oxidation of char, in which oxygen reacts with char to produce carbon dioxide and water vapour. Other than this, the oxidation of volatiles also takes place. These reactions can be represented by



Besides the main char-carbon dioxide and char-water vapour reactions, hydrogasification and water-gas shift reactions also occur in the gasification zone. They can be written as follows.



### 4.3 Model Classification

In order to improve the design of the NERD gasifier, a model that allows the study of the gasifier performance is required. Thus, a survey of the available models in the literature was carried out. These are steady state, quasi-steady state and transient state. Steady state models are further classified as thermodynamic equilibrium, kinetics free equilibrium and rate models (Buekens & Schoeters 1985).

Most of the thermodynamic equilibrium models predict the outlet gas composition based on the assumption that the gasifier reactions are in thermodynamic equilibrium. In some models the reaction temperature is assumed to determine the gas composition. But according to Buekens and Schoeters (1985) the real gas composition deviates from the thermodynamic one because of imperfection in the reaction temperature. Moreover, since the models are based on homogeneous operating conditions, no concentration or temperature profiles can be predicted. It is also not possible to derive the required reaction volume due to absence of rate data. The main advantage of the thermodynamic model is its simplicity.

Kinetic models are more or less similar to thermodynamic equilibrium models but the reactor is divided into two zones. The temperatures of these two zones are calculated using an energy balance. Such models are also not recommended if feasible operating points are not available from experiments or rate models. If the operating conditions are not optimum, gasification reactions become slow to consume all the char. As a result some char is lost in the residue (Buekens & Schoeters 1985). The gas composition in the gasification zone is calculated from equilibrium data where the flow rate and composition of the solid fuel and the temperature and the composition of the air are known.

Steady state rate models are used to compute the gas composition and the temperature profile along the reactor axis but they cannot predict the effects of input parameters such as moisture content and size of solid fuel, inlet air temperature, reactor load, insulation and reactor size on conversion efficiency. These models include rate equations that consist of mass and energy balances for the heterogeneous and

homogeneous reactions and result in a set of algebraic and ordinary differential equations.

Quasi-steady and transient models do not neglect the time derivatives. They are based on the mass balances. Transients are calculated using pseudo-steady state assumptions based on the interpolation of the results of the steady state models. These models are capable of predicting small changes in operating parameters such as solid feed rate or air flow on reactor performance but these inputs need to be calculated from steady state models. A summary of the available models is given in Table 4.1.

#### **4.4 Model Criteria and Review of Gasifier Models**

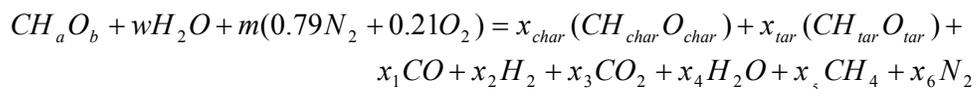
The objective of simulating the NERD gasifier is to identify the optimum design parameters and desirable operating parameters for a better output. In order to review the available models, it is necessary to establish a set of criteria based on the above objectives by which each model is evaluated. The criteria chosen are fixed bed downdraft type, steady state or quasi-steady, ability to predict the composition of product gas and the temperature profile in the gasification zone in terms of operating parameters. Based on these criteria, only the Milligan's and Chen's models will be reviewed in detail.

Table 4.1: Summary of the models

Model type	Objective	Reference
Steady state (Thermodynamic)	To predict outlet gas composition	Ruggiero and Manfrida (1998)
Steady state (Thermodynamic)	To predict outlet gas composition	Shand and Bridgwater (1984)
Steady state (Thermodynamic)	To predict outlet gas composition	Belleville and Capart (1984)
Steady state (Thermodynamic)	To predict the pyrolysis products and gas composition	Gibbins and Wilson (1983)
Steady state (Thermodynamic)	To predict outlet gas composition	Gumz (1950)
Steady state (Rate)	To predict the gas composition at different operating conditions and to evaluate the influence of temperature, pressure and air fuel ratio on outlet gas composition	Bettagali et al. (1995)
Steady state (Rate)	To investigate the effect of moisture content, size of fuel wood and degree of insulation on turndown ratio	Chandra and Payne (1986)
Steady state (Kinetic)	To predict the gas composition and axial temperature profile	Chowdhury et al. (1994)
Quasi-steady	To predict the gas composition and axial temperature profile in terms of input parameters	Di Blasi (2000)
Quasi-steady	To calculate the temperature, composition of pyrolysis gas, and the reactor height.	Milligan (1994)
Quasi-steady	To predict the temperature and gas composition of pyrolysis gas and to predict the composition of product gas	Chern (1989)
Quasi-steady	To investigate the effects of input air temperature, wood chip size, wood moisture content, reactor load and reactor insulation on gasifier performance	Chen (1987)
Quasi-steady	To investigate the effect of wood chip size, wood moisture content, reactor load and reactor insulation on gasifier performance and to predict the product gas composition	Groeneveld (1980)
Transient	To investigate the effect of flow conditions on gasifier performance	Kayal et al. (1994)
Transient	To predict the height of the gasification and pyrolysis zones	Reed and Markson (1983)

#### 4.4.1 Milligan's model

Milligan (1994) has developed a downdraft gasifier model to investigate the effects of process variables on gasifier performance. The objectives of the modelling were to predict the length of the gasification zone and the composition of the product gas. The model, which consisted of three sub-models, was developed using basic gasification reactions as well as experimental data. In the first stage, the modelling of the flaming pyrolysis zone was described. The aim of this sub-model was to calculate the composition of the product gas such as CO, H<sub>2</sub>, CO<sub>2</sub>, H<sub>2</sub>O, CH<sub>4</sub> and N<sub>2</sub> entering the gasification zone. The following basic wood reaction equation for carbon, hydrogen, oxygen and nitrogen balances for the flaming pyrolysis zone was used.



The remaining two equations used in this first sub-model were derived from a heat balance and the equilibrium constant of the water-gas shift reaction. Experimental char and tar yields from the pyrolysis zone were used as the amount of char leaving the combustion zone. The second sub-model described the modelling of char gasification based on external mass transfer, internal mass transfer, reaction kinetics, mass and balances. The aim of this sub-model was to estimate the length of the gasification zone necessary for complete char gasification. The third sub-model was an equilibrium model which was used to predict the composition of the product gas.

Milligan's model is capable of predicting the composition of product gas but not the temperature profiles. Instead the temperature profile has been measured to calculate the heat losses. Since the gasification zone sub-model is based on experimental data of the gasifier under review, the applicability of this model for other types of gasifiers with different capacity and characteristics is not possible without necessary experimental data.

#### 4.4.2 Chen's model

A detailed modelling program was developed by Chen (1987) for the simulation of a downdraft gasifier. The objectives of this model were to estimate the length of the gasification zone, reactor diameter and to investigate the dependence of reactor performance on operating parameters such as feedstock moisture, chip size, reactor insulation, input air temperature and the gasifier load.

The model consisted of three sub-models. In the first sub-model, the concept of the chemical reaction equilibrium together with mass and energy balance principles was applied to assess the amount of oxygen needed for a fixed input of fuel at a specific operating condition. In assessing the amount of oxygen required, the heat loss in the reactor wall has been neglected. In the second sub-model, the drying, pyrolysis and combustion zones were lumped together and considered as a single zone. The mass and energy conservation equations with the fuel to air ratio estimated from the first sub-model were used for the lumped zone to obtain the product concentrations and the temperatures of the gaseous and solid phases leaving the zone. It has been assumed that the amount of carbon monoxide, hydrogen and tar leaving the lumped zone to be negligible and the amount of methane and higher hydro carbons to be equal to 2.5% and 1.5% respectively. These calculated concentrations and temperatures were used as inputs to the third sub-model.

The third sub-model was based on the principles of thermodynamics, transport processes, hydrodynamics of solid and gas flow and mass and energy balances. This sub-model predicts the temperature profile along the axis of the gasification zone, gas composition, conversion efficiency of the product gas and the length of the gasification zone at any given time interval.

Milligan (1994) compared the results of his flaming pyrolysis zone sub-model with the lumped zone model developed by Chen (1987). Based on the results, he suggested that Chen's model over predicted the temperature of the gas leaving the combustion zone compared to his model. Chen's model predicted 1750 K as zone exit temperature whereas Milligan's model predicted 1296 K for the same wood moisture content (10% on dry basis) and air-fuel ratio (1.8). Milligan's conclusion was that the

assumption of a negligible amount of carbon monoxide and hydrogen leaving the zone was the reason for the high temperature predicted by Chen's model. Reed and Markson (1983) reported that gases like CO and H<sub>2</sub> are usually present in the pyrolysis products. The omission of carbon monoxide and hydrogen gases would result in a lower amount of chemical energy and therefore a high sensible heat content in the gas in the pyrolysis zone but the fraction of heat losses used in each model should also be considered before any conclusion is made. Chen's model calculates heat loss as 1.5% of input energy whereas Milligan (1994) has used 10.5% of input energy as heat loss for his model for comparison. If 10.5% heat loss is used for Chen's model the results would be as shown in Table 4.2.

Table 4.2: Comparison of Chen's and Milligan's models for 10.5% heat loss

Parameter	Chen's model	Milligan's model
Char yield, % mass	31.4	16.7 <sup>e</sup>
Zone exit temp. (K)	1375	1296
Composition of gas leaving flaming pyrolysis zone, % by vol.		
H <sub>2</sub>	0.0 <sup>a</sup>	10.6
CO	0.0 <sup>a</sup>	12.1
CO <sub>2</sub>	13.5	11.6
H <sub>2</sub> O	33.8	18.7
N <sub>2</sub>	50.7	45.0
CH <sub>4</sub>	2.0 <sup>a</sup>	2.0 <sup>a</sup>

(Feed analysis of CH<sub>1.46</sub>O<sub>0.65</sub>, moisture content of 10% on dry basis (d.b), an air fuel ratio of 1.8 by mass and tar yield of 1% by mass were used for comparison. a = assumed, e = experimental)

The product gas composition predicted by Chen's model is given in Table 4.3 and compared with typical experimental ranges reported by Graham and Huffman (1981) for commercial scale downdraft wood gasifiers. It can be seen from this table that the gas composition predicted by Chen's model is in the range typical to downdraft gasifiers, similar to several other models found in the open literature.

The model developed by Chen (1987) is applicable for most downdraft wood gasifiers. It is also capable of predicting the effects of operating and design parameters on gasifier performance and the length of the gasification zone. But the over prediction of the zone exit temperature due to low heat loss and the omission of CO and H<sub>2</sub> in the pyrolysis product gas is the main weaknesses of this model.

Table 4.3: Comparison of results of Chen's model with typical experimental range

Gas composition (% by vol.)	Chen's model	Typical range reported by Graham and Huffman (1981)
CO	18.5	11.9-26.3
H <sub>2</sub>	15.8	8.9-18.0
CO <sub>2</sub>	12.8	9.1-17.7
N <sub>2</sub>	48.7	41.5-54.7
CH <sub>4</sub>	2.6	2.1-3.5
Trace	1.6	1.5-2.6

#### 4.5 Model Selection and Improvements to Selected Model

Of the models reviewed, only Chen's model appears to be appropriate for this study. His model was basically developed for the downdraft wood gasifiers and can be used to study the effects of operating and design parameters on reactor performance. To overcome some of the weaknesses of his model, the following improvements were implemented.

- Inclusion of Milligan's flaming pyrolysis zone sub-model in place of the lumped zone model with experimental heat losses,
- Inclusion of the amount of water vapour in the incoming air which has an effect about 5% on the oxidation temperature,
- Inclusion of the wall temperature (initial assumption) for better estimation of heat losses.

#### **4.6 Model Development and Implementation**

The model development consists of two sub-models namely flaming pyrolysis and gasification zone sub-models. Flaming pyrolysis zone sub model is used to determine the product concentration and temperature of gas leaving the flaming pyrolysis zone. The concepts of equilibrium in chemical reactions with mass and energy balance principles are used in the model development. The calculated concentrations and temperatures of the flaming pyrolysis zone sub-model will then be used as inputs to the gasification zone sub-model. The gasification zone sub-model was developed by Chen (1987) on the basis of gasification of a single char particle which moves vertically downwards along the vertical axis of the gasifier. The gasification zone sub-model includes a description of the physical and chemical processes taking place in the multi-phase downdraft moving beds with flow equations, transport phenomena and conservation principles. The following section discusses the assumptions made in developing the flaming pyrolysis zone sub-model and the justification of those assumptions.

##### **4.6.1 Flaming pyrolysis zone sub-model assumptions**

The following assumptions were made in the development of the flaming pyrolysis zone sub-model.

- that Pseudo steady state conditions were approximated.

The objective of the pseudo steady state approximation is to eliminate the time derivatives from the mass and energy balance equations. A mathematical justification comparing the steady state and unsteady state solutions was performed by Wen (1968) showing the conditions for a pseudo steady state approximation to be valid. For most gasification reactions, the conditions are satisfied.

- that the ideal gas law is applicable.

The ideal gas law is used to calculate the gas concentrations and the velocities. The simplest modification of the ideal gas equation to fit the real gas behaviour is the introduction of a compressibility factor  $Z$ , where  $Z = PV/nRT$ . For an ideal gas,  $Z$  is equal to 1. Chen (1987) has quoted from Jones and Hawkins (1960) that at the gasification conditions, the  $Z$  values for the major components of the gas are close to unity. This justifies the validity of assuming the applicability of the ideal gas law.

- that char yield is equal to the amount of fixed carbon in fuel wood.

Other than the fixed carbon amount in the fuel, the actual char yield also depends on the heating rate and the temperature of the process. However Milligan (1994) stated that under the prevailing operating conditions of the pyrolysis zone, only a little variation in char yield can be expected. Therefore it is assumed that the amount of char leaving the pyrolysis zone is equal to the amount of fixed carbon in the fuel wood.

- that the methane content is 1.5%.

At atmospheric pressure, the hydrogasification reaction ( $C + 2H_2 \rightarrow CH_4$ ) produces a very small amount of methane (Walker et al. 1959). Results of the experiments carried out by Milligan (1994) show 1 to 2 % of methane concentration in the product gas. Thus the trace amount of methane existing at the end of the oxidation zone is assumed to be 1.5 %.

- that the tar content is 1.0%.

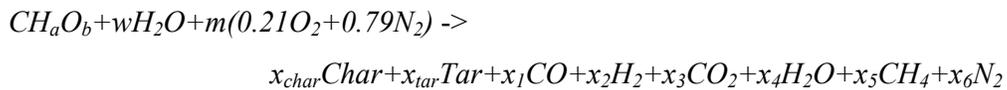
The experiments carried out by Graham and Huffman (1981) shows that the amount of tar found in the product output is around 1.0 % of total mass for chips 1.3 to 2.5 cm. Therefore the amount of tar is assumed to be equal to 1.0 % by mass.

- that the nitrogen and the ash contents of the wood can be neglected.  
The experiments carried out by HRL Technology Pty Ltd, Victoria Australia, to determine the proximate and ultimate analysis of rubber wood indicate that the amount of nitrogen and ash content as 0.22% and 0.67% respectively. So this assumption is acceptable.

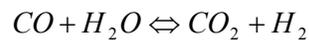
#### 4.6.2 Flaming pyrolysis zone sub-model

The particle temperature has been included in Milligan's flaming pyrolysis zone sub-model which is used to determine the composition and temperature of gas leaving the flaming pyrolysis zone. The basic equations of the model are as follows.

The general equation of reaction of wood in the flaming pyrolysis zone can be written as:



Char is taken as carbon and the ultimate analysis of tar is taken as  $CH_{1.03}O_{0.03}$  (Adams 1980). The equilibrium equation is obtained from the following reaction and the corresponding equilibrium constant.



$$K_3 = \frac{x_3 \times x_2}{x_1 \times x_4} \quad (4.1)$$

Gumz (1950) has given the correlation between the temperature and equilibrium constants for the above reactions in empirical formulae.

$$\text{Log}(K_3) = -36.72508 + \frac{3994.704}{T} - 4.46241 \times 10^{-3} T + 6.71814 \times 10^{-7} T^2 + 12.22028 \log(T) \quad (4.2)$$

Where  $T$  is temperature in Kelvin

By carbon mass balance:

$$1 = x_{char} + x_{tar} + x_1 + x_3 + x_5 \quad (4.3)$$

By hydrogen mass balance:

$$a + 2w = 1.03x_{tar} + 2x_2 + 2x_4 + 4x_5 \quad (4.4)$$

By oxygen mass balance:

$$b + w + 0.42m = 0.03x_{tar} + x_1 + 2x_3 + x_4 \quad (4.5)$$

By nitrogen mass balance:

$$0.79m = x_6 \quad (4.6)$$

The energy balance in flaming pyrolysis zone is

$$H_c \text{Wood} = H_c \text{Char} + H_c \text{Tar} + H_c \text{Gas} + H_s \text{Char} + H_s \text{Tar} + H_s \text{Gas} + \text{Heat Loss} \quad (4.7)$$

$w$  can be calculated as follows.

Moisture in fuel wood = dry matter in fuel wood  $\times$  moisture content on dry basis

$$w = (12 \times 1 + 1 \times a + 16 \times b) \times mc_{db} \text{ kg}$$

The values of  $a$  and  $b$  are given. The values of heat loss and  $m$  are taken from the experiment. The values of  $x_5$ ,  $x_{char}$  and  $x_{tar}$  are assumed. The higher heating value

(kJ/g) of wood, char and tar are calculated from the equation given by Channiwala (1992) as follows.

$$H_C \text{ Wood} = 0.3491f_C + 0.1783f_H - 0.1034f_O \text{ (} N_2, \text{ and ash contents are neglected)}$$

$$H_C \text{ Char} = 0.3491 \times f_{C, \text{char}}$$

$$H_C \text{ Tar} = 0.3491 \times f_{C, \text{tar}} + 0.1783f_{H, \text{tar}} - 0.1034f_{O, \text{tar}}$$

Chemical energy content of output gas, and sensible energy of char, tar and output gas are calculated as follows.

$$H_C \text{ Gas} = 241000x_1 + 283000x_2 + 802300x_5$$

$$H_S \text{ Char} = 12.15 \times x_{\text{char}} \times (T-300)$$

$$H_S \text{ Tar} = 21.95 \times x_{\text{tar}} \times (T-300)$$

$$H_S \text{ Gas} = x_1 \times H_{CO} + x_2 \times H_{H_2} + x_3 \times H_{CO_2} + x_4 \times H_{H_2O} + x_5 \times H_{CH_4} + x_6 \times H_{N_2}$$

The enthalpy data from JANAF Thermochemical Tables (1986) were used to correlate the enthalpies of the gases for temperature ranging from 200 to 2500 K and they are given in Appendix B.

There are six unknowns namely  $x_1$ ,  $x_2$ ,  $x_3$ ,  $x_4$ ,  $x_6$  and T to be determined using six equations. The successive approximation method with a Fortran programme is used to solve the equations. The subroutines used and the code of the computer programme are given in Appendix D and E respectively.

#### 4.6.3 Gasification zone sub-model assumptions

The following assumptions were made in the development of the gasification sub-model.

- that Pseudo steady state conditions were approximated.

Described in Section 4. 6.1

- that all the gas-solid reactions occur within the particle.  
The experiments conducted by Turkdogan et al. (1969) to measure the internal pore surface area of carbon shows that biomass char generally has an internal specific surface area of 350 to 1061 m<sup>2</sup>/g. Thus the external surface area of the particle will represent an insignificant fraction of the total surface area.
- that there were negligible convective transport effects.  
The dilute approximation of the reactant species concentrations which is in order of 1 to 20 kmol/m<sup>3</sup> around the particle permits the use of isobaric conditions within the particle (Chen 1987). Thus convection within the particle is neglected and diffusion is the only dominant mechanism responsible for the transport process.
- that char entering the gasification zone is pure carbon.  
Complete carbonization, which occurs at the end of the oxidation zone, is justified by the experimental results compiled by Blackwood and McCarty (1966). Their experimental results show that the wood char produced at 1173K contains 97.2% C, 1.0% H<sub>2</sub>, 1.4% O<sub>2</sub> and 0.4% N<sub>2</sub> in a dry and ash free basis. Moreover the high temperature in the oxidation zone increases the degree of carbonization. Thus the assumption that char entering the gasification zone is pure carbon is reasonable.
- that the ideal gas law is applicable.  
Described in Section 4.6.1
- that the nitrogen and the ash contents of the wood are negligible.  
Described in Section 4.6.1
- that there is negligible degradation of tar in the gasification zone  
Described in Section 4.6.1
- that the particles are assumed to be spherical in shape.

This assumption is acceptable since the diffusion calculation is typically insensitive to the geometry of the particle (Weisz & Prater 1954). Thus the effectiveness factor applicable to the sphere will be satisfactory for non-spherical particles as long as there are no large differences in three dimensions.

- that the amount of tar present in the output gas is negligible.  
The experiments carried out by Walawender et al. (1985) to determine the energy output of a downdraft gasifier for a feed rate of 48.8 kg/hr indicate that the amount of tar present is 0.9% of the total energy output. So this assumption is valid.
- that the bed porosity varies from 0.5 at the top to 0.3 at the bottom.  
The variation of bed porosity with particle size equation was modified from Perry et al. (1997).
- that the end of the gasification zone is reached when the carbon particle concentration is less than 5% of the initial char concentration.

#### **4.6.4 Gasification zone sub-model**

Modelling of this zone is performed by following a particle along the axis of reactor. The length co-ordinate is coupled with a time variable through the solid phase velocity. A small time increment approach is used in calculating the product composition of the zone. This approach involves the use of a step procedure starting from the gasification zone and marches axially through the reactor in appropriate time increments. The input values of the first interval are the output values of the flaming pyrolysis zone.

The model development results in a set of non-linear equations which are given in Appendix C. A computer programme has been formulated using Fortran language to calculate the characteristic profiles along the reactor axis. The profile includes temperature, concentrations, efficiency and distance the particle travelled. The correlated enthalpies, viscosity, diffusivity and thermal conductivity of gases used in

the program developed are given in Appendix B. The subroutines used and the code of the computer programme are given Appendix D and E respectively.

#### **4.7 Conclusions**

A model of a downdraft wood gasifier has been to predict the composition of the product gas. This model can also be used to estimate the required length of the gasification zone for complete reaction. The complete model consists of two sub models. The flaming pyrolysis zone sub-model is used to predict the composition of the gas leaving the combustion zone. The gasification zone sub-model has been used to predict the output of the product gas and the length of the gasification zone at any given time step. Although the model predictions were in the acceptable range (Table 4.3), it needs calibration in order to be used for parametric study for the NERD gasifier. This can be done by using experimental data. The following chapter describes the experimental work carried out to calibrate the model.

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## **CHAPTER 5**

### **EXPERIMENTAL WORK**

#### **5.1 Introduction**

Chapter Four describes the development of the downdraft gasifier model which is capable of predicting the effects of operating parameters such as air fuel ratio, chip size and fuel moisture content and design parameters such as insulation level and throat angle on the conversion efficiency of the gasifier. This chapter describes the details of the experimental work done to verify and calibrate the model. The chapter begins with a layout diagram of the experimental system. Then a description of the test gasifier, processing and characterization of rubber wood are discussed. It is followed by instrumentation, calibration and the experimental procedure. Finally the summary of the results gathered from a series of experiments is presented.

#### **5.2 Layout of the Experimental System**

The experimental test rig consists of a downdraft wood gasifier, air blower, air flow meter, data logger, computer, orifice plate, manometer, cooling system, gas sample collector, thermocouples and scales. Figure 5.1 shows the schematic diagram of the experimental system. An orifice plate was fabricated according to the Australian Standard (1993) to measure the gas flow rate. Appendix F describes the method of calculating the mass flow rate of the product gas. The inlet and outlet pipes of the orifice plate were made from galvanized iron and the flanges were made from mild steel. Stainless steel was used for the orifice plate. A manometer was used to measure the pressure drop across and at the upper stream of the orifice plate. The cooling system consists of a coiled tube made from copper. The coiled tube was immersed in ice to condense the water vapour in the gas. The gas was collected in a special type of

sample bag. A data logger coupled with the computer was used to collect the temperature measurements made inside and outside wall of the gasifier.

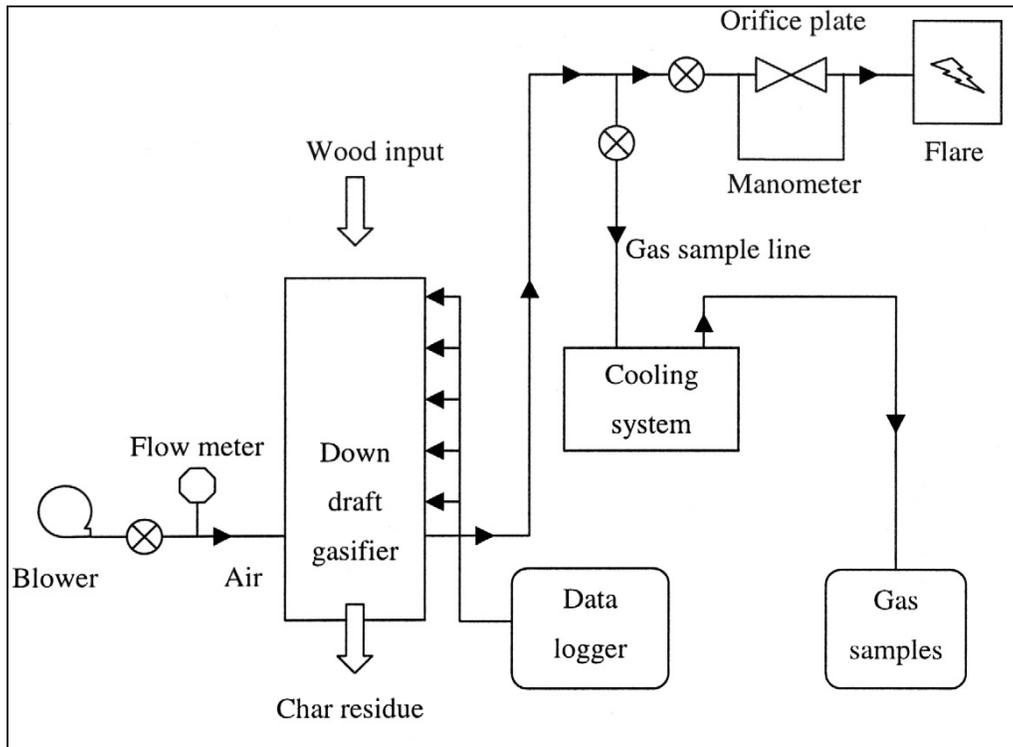


Figure 5.1: Schematic diagram of the experimental system

### 5.3 Details of the Test Gasifier

The gasifier used for this experiment was similar in design to the one used for the tea drying process but smaller in terms of capacity. The main reason for the selection of the smaller capacity gasifier was to reduce the cost of running the unit. The gasifier is of downdraft type with a reactor inner diameter of 0.92 m. Rubber wood is fed from the top and the charging door is kept closed throughout the operation. The outer surface of the drum is made from galvanized iron with a thickness of 1.4 mm and the inner surface is made from 22 gauge mild steel. A drain pipe of 50 mm in diameter is located at the end of the drum to extract any water that condenses out during the drying process. Air is sent to the combustion zone through 12 air nozzles of 6-mm diameter, located 100 mm above the throat. The throat is lined with high alumina castable and is surrounded by firebricks to withstand the high temperature in the hearth zone. The throat diameter is 100 mm and consists of a divergent angle of 57

degrees. The length of the gasification zone is 22 mm from the throat. Below the gasification zone lies the grate to hold unreacted char and the ash chamber is situated beneath the grate. A galvanized steel pipe of 50 mm in diameter is used to transport the product gas from the system. A high temperature resistant (up to 1700° C) ceramic tube was used to install the k-type thermocouples at the centre of the gasification zone. The picture of the test gasifier is shown in Figure 5.2 and the schematic diagram with the thermocouple positions and dimensions is shown in Figure 5.3.

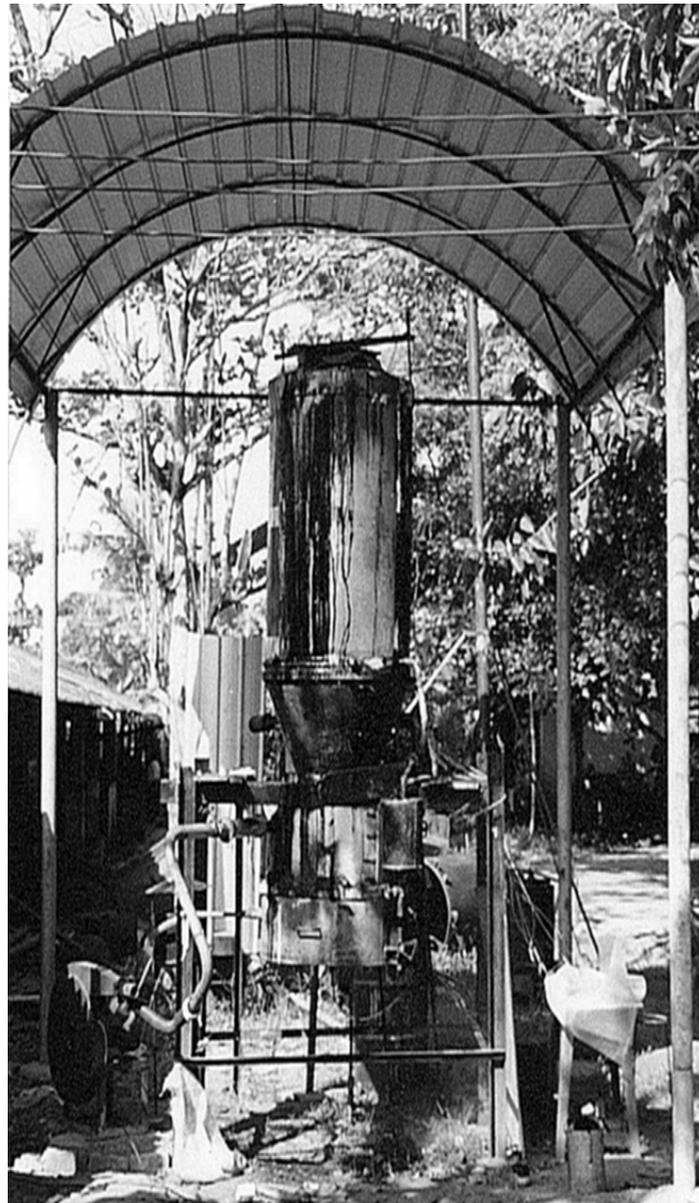


Figure 5.2: Picture of the test gasifier

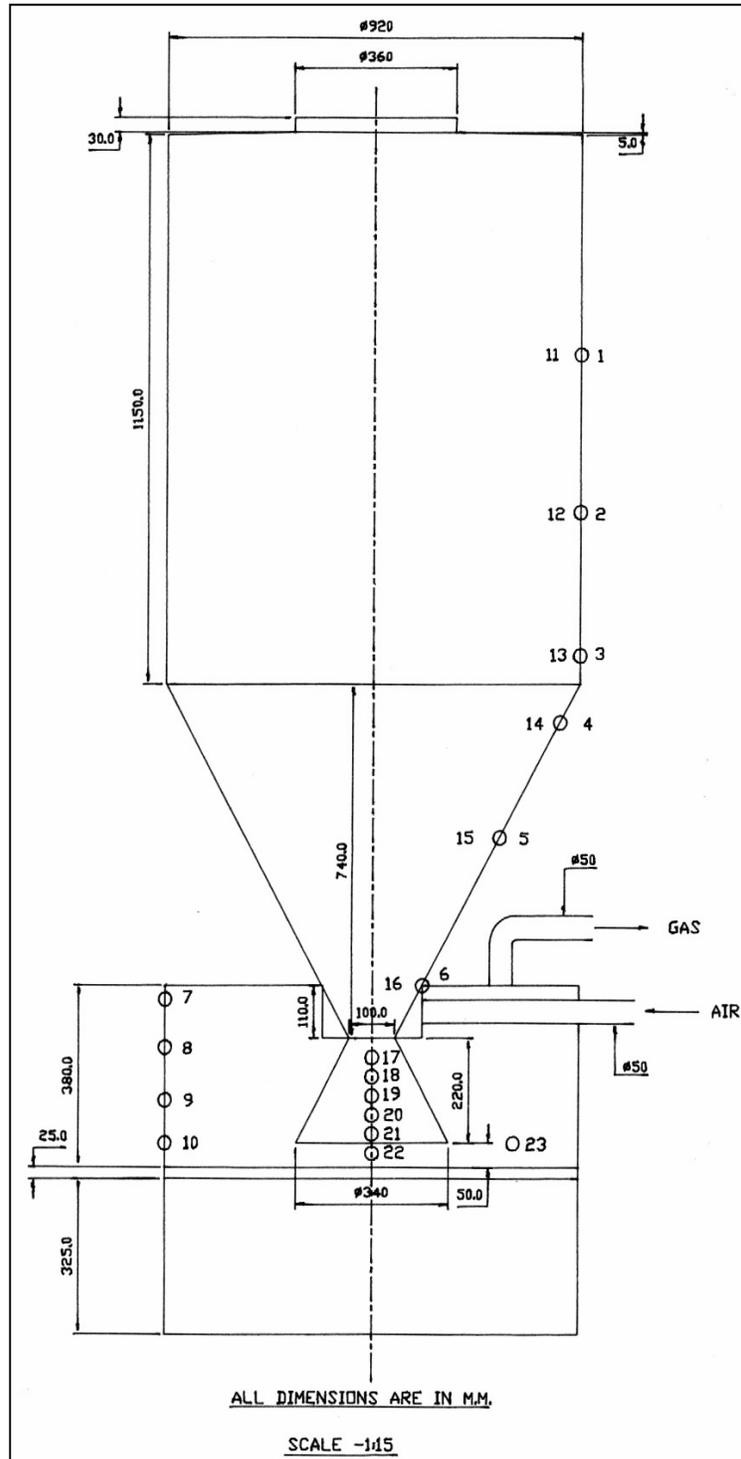


Figure 5.3: Schematic diagram of the gasifier

## 5.4 Feed Processing and Characterization

The main fuel wood used in the tea industry in Sri Lanka is rubber wood. It accounts for 70% of the total fuel wood consumption by the industry (Haskoning 1989). Thus for this study rubber wood has been selected as the feed material. The wood logs were purchased from local suppliers on a cubic metre basis and sun-dried for three weeks as wet wood with the moisture content of 35~50% (w.b.) was found to be difficult to cut using saws. The rubber wood was then cut into small cubes with sizes varying from 1.5 to 2.5" (3.3 to 5.5 cm) and they were dried again for a week until the desired moisture level 10~15% (w.b.) was achieved.

The process of gasification is governed by the characteristics of the feed materials used. An understanding of the properties of feed materials is necessary in order to evaluate their utility as feed stocks for the gasification process (Chen 1987). The moisture content of the feed material is one of the main characteristics, which affects the composition of the product gas (Hos & Groeneveld 1987). The effects of moisture on the recoverable heat are very significant due to the heat requirement for vapourizing the moisture and superheating the vapour (Graboski & Bain 1981). The moisture content of rubber wood was determined by the percent weight loss of a 10g sample at 105°C constant temperature for one hour. Three samples were tested for each run and the average value was taken. The range of moisture content of the samples varied 11.1-15.6% (w.b.).

The proximate and ultimate analysis of the rubber wood were carried out by HRL Technology Pty Ltd, Victoria, Australia. The following pieces of equipment were used to determine the rubber wood characteristics. Leco MAC400 Analyzer for moisture content and ash yield, Leco CHN600 Analyzer for C, H and N percentages and Australian Standard 2434.2 was used to determine the volatile matter. The proximate analysis of rubber wood is given in Table 5.1 and the measured value is compared with the figures quoted by Hoi et al. (1992). The ultimate analysis is given in Table 5.2.

Table 5.1: Proximate analysis of rubber wood

Parameter	Proximate analysis (% dry basis)	
	Measured	Hoi (1992)
Volatile matter	80.1	80.2
Fixed carbon	19.2	18.6
Ash content	0.7	1.2

Table 5.2: Ultimate analysis of rubber wood

Parameter	Ultimate analysis (%)	
	Dry basis	Dry and ash free basis
C	50.60	50.94
H	6.50	6.55
N	0.22	0.22
$O = \{100 - (C+H+N+Ash)\}$	42.01	42.29
Ash content	0.67	-

The higher heating value of rubber wood was determined by HRL Technology Pty Ltd, Victoria Australia, using Leco AC350 Calorimeter. A number of researchers have found that the elemental composition found in the ultimate analysis of biomass is closely related to the heat of combustion (Graboski & Bain 1981 and Reed & Das 1988). Gaur & Reed (1998) have quoted from Channiwala (1992) that the heat of combustion can be predicted from the ultimate analysis according to

$$HHV = 0.3491C + 1.1783H + 0.1005S - 0.1034O - 0.0151N - 0.0211Ash \quad (kJ/g)$$

Where HHV is higher heating value of the sample, and C, H, S, O, N and Ash are weight percentage of carbon, hydrogen, sulphur, oxygen, nitrogen and ash. It is reported that the calculated value agrees with the measured value with an absolute error of 1.45%, better than Reed's (1981) correlation which reports an absolute error of 2.1%. The measured higher heating value of rubber wood is given in Table 5.3 and is compared with the estimated figure and the figure quoted by Hoi et al. (1992).

Table 5.3: Higher heating value of rubber wood

Source	Higher heating value on dry basis (MJ/kg)
Estimated	20.96
Measured	19.55
Hoi et al. (1992)	18.44

The bulk density of rubber wood was obtained by dividing the weight of air dried samples by the volume it occupied in a cylinder of 44 cm in diameter and 39 cm in height. The range of the bulk density of the samples varied from 313.5 – 330.0 kg/m<sup>3</sup>.

### 5.5 Instrumentation

The instruments used in the field experiments a data logger, computer, thermocouples, air flow meter, orifice plate, manometer, dry bulb-wet bulb thermometer, digital thermometer and scales. The gas composition of the product gas was analyzed by Ceylon Institute of Scientific and Industrial Research (CISIR) using gas liquid chromatography. A molecular sieve column was used to analyze H<sub>2</sub>, O<sub>2</sub>, N<sub>2</sub>, CO and CH<sub>4</sub> while a Carboxine 1000 column with temperature program 50°-225°C at 20°C/min was used to determine CO<sub>2</sub>. The data logger was used to monitor the axial temperatures in the gasification zone and the outer surface of the gasifier. All the temperatures monitored by data logger were then recorded in the computer. K-type and T-type thermocouples were used for temperature measurements. Input air flow rate was measured using the air flow meter. An orifice plate was used to determine the flow characteristics of product gas to calculate the gas flow rate. The method of calculation is given Appendix F. A manometer was used to obtain the pressure readings at the orifice plate. Ambient conditions were measured using a dry bulb-wet bulb thermometer and the outlet gas temperature was measured using a digital thermometer. Finally rubber wood intake was measured using scales. Table 5.4 gives the details of the instrument used at the field site.

Table 5.4: Instruments used for the experiment

Equipment	Description	Accuracy
Data logger	Data Electronics acquisition unit Model 500, Serial No. 16136	-
Computer	IBM PC compatible, ACER-355, Serial No. M001F2	-
Thermocouples	K-type, T-type	$\pm 0.01\%$
Air flow meter	Pitot tube type	$\pm 0.1\%$
Manometer	Water	-
Dry bulb-wet bulb thermometer	Octagon-20	-
Digital thermometer	Temperature measuring meter Model number HI 91530 K	$\pm 0.5\%$

## 5.6 Calibration

The accuracy of the thermocouples given by the manufacturer was  $\pm 0.01\%$ . Since the range of the temperature readings measured lay between 600 to 1500 K, the maximum absolute error that can occur was  $\pm 15\text{K}$ . This error was too small and would not have any effect on output gas composition. Therefore greater accuracy for temperature readings was not required for this study and the accuracy of the thermocouple measurement given by the manufacturer was assumed to be adequate. The air flow meter was calibrated by the NERD Centre to  $\pm 1.0\%$  of full scale and the digital thermometer was calibrated by the manufacturer at purchase. The readings of a dry bulb-wet bulb thermometer were used to calculate the amount of moisture in the air. The air moisture represented only 0.5% of total input by weight and its effect on output was not significant. Therefore the dry bulb-wet bulb thermometer was not calibrated, but its reading were compared with a calibrated mercury thermometer.

## 5.7 Experimental Procedure

The aim of the experiments were to evaluate the effects of operating parameters such as air-fuel ratio, chip size and fuel moisture content on the thermal performance of the gasifier. Initially all the instruments necessary for the experiments were installed. Next, the moisture content and the bulk density of wood chips were measured. Three samples selected randomly were tested for each run and the average value was taken. Then the gasifier was loaded with wood cubes and a measurement of the load was recorded. The blower was switched on and air was allowed to flow. Then the air flow was set to the desired level. The chips were ignited through the ignition port at the throat. As the production of gas started, the gas was ignited. After the system started gas production, readings such as air flow rate, temperatures and pressure measurements were recorded. A sample of product gas was collected for gas analysis. The experiments were conducted for three chip sizes and for four air-fuel ratios.

## 5.8 Summary of Results

Table 5.5 and 5.6 give the key results of the experiment conducted for chip size of 3.3 cm and three air-fuel ratios. The results include the axial temperature profile of the gasification zone and the gas composition of the product gas. It can be seen from Table 5.6 that the composition of the gas measured are in the range typical to downdraft gasifier (Table 4.3) reported by Graham and Huffman (1981) except for methane content, which is low. The detailed results are given in Appendix G.

Table 5.5: Axial temperature variations in gasification zone for chip size 3.3cm

Chip size (cm)	Ave. m/c % (d.b.)	Air/fuel ratio	Temperature (K)					
			Distance from throat (cm)					
			4	8	12	16	20	24
3.3	18.5	2.03	1237	1252	1174	1155	1002	762
	16.0	2.20	1255	1276	1188	1170	1032	826
	14.7	2.37	1311	1321	1263	1190	1043	848
	13.6	2.54	1335	1351	1294	1212	1063	865

Table 5.6: Dry gas analysis for chip size 3.3 cm

Chip size (cm)	Ave. m/c % (d.b.)	Air/fuel Ratio	Dry gas analysis (% by vol.)				
			CO	H <sub>2</sub>	CO <sub>2</sub>	CH <sub>4</sub>	N <sub>2</sub>
3.3	18.5	2.03	19.6	17.2	9.9	1.4	51.9
	16.0	2.20	20.2	18.3	9.7	1.1	50.7
	14.7	2.37	19.5	17.2	9.6	1.1	52.6

## 5.9 Conclusions

This chapter described the details of the experimental work including a description of the test gasifier, the characteristic of the rubber wood, instrumentation and experimental procedure. Finally a summary of the experiments is presented. The previous chapter stated the need to calibrate the gasifier model developed to predict the gas composition. These experimental results have been used to calibrate the gasification sub-model. The next chapter discusses the verification and calibration of the flaming pyrolysis and gasification zone sub-models. It is followed by a parametric study to evaluate the effects of operating and design parameters on conversion efficiency of the gasifier.

## 5.10 Reference

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## CHAPTER 6

### MODEL CALIBRATION AND PARAMETRIC STUDY

#### 6.1 Introduction

Chapter Four states the need to calibrate the flaming pyrolysis and gasification zone sub-models developed to study the NERD gasifier. Chapter Five describes the details of the test gasifier, characteristics of the rubber wood, instrumentation, experimental procedure and finally presents a summary of the experimental data gathered. This chapter begins with a verification of the predictions of the flaming pyrolysis zone sub-model. This sub-model is used to calculate the input parameters to the gasification zone sub-model. The results of a parametric study to investigate the effects of air fuel ratio, heat loss and char yield on the temperature of the gas leaving the flaming pyrolysis zone are then discussed. In order to present the predictions of the flaming pyrolysis zone sub-model for the NERD gasifier, it is necessary to calculate the actual heat loss of the system. Therefore material and energy balances of the system are presented to calculate the heat losses. It is then followed by the verification and calibration of the gasification zone sub-model. Another parametric study is presented to investigate the effects of wood chip size, moisture content, inlet air temperature, heat loss and the throat angle on conversion efficiency. Finally based on the simulation studies, the desirable wood chip size, moisture content, heat loss and the optimum length of gasification zone are discussed.

#### 6.2 Verification of Flaming Pyrolysis Zone Sub-Model

Since no experimental measurements of the pyrolysis gas composition within a gasifier could be found in the literature, the results of the flaming pyrolysis zone sub-model are compared with the results of the theoretical models given in the literature. The composition and temperature of the product gas predicted by the flaming

pyrolysis sub-model are given in Table 6.1 and compared with the results given by Milligan (1994) and Chern (1989). It can be seen from Table 6.1 that the gas temperature predicted by the flaming pyrolysis zone sub-model is almost the same but the gas compositions vary within 8.5% compared to Milligan's results. Model predictions vary depending on the assumptions and the equations used. For this study the solid temperatures were calculated from the correlation given by Chen (1987) whereas Milligan (1994) has assumed the same temperature for solid and gas. Di Blasi (2000) and Reed and Das (1988) stated that these temperatures cannot be assumed to be same. For this study, char is assumed to be pure carbon whereas Milligan (1994) has assumed that char consists of 95% carbon and the remainder is hydrogen and oxygen. The absolute error of equations used to calculate the heating value of feed also varies and affects the final composition. However despite the discrepancies, it was felt that the predictions of the flaming pyrolysis zone sub-model were close enough to those of other models to conduct a parametric study.

Table 6.1: Comparison of results of flaming pyrolysis zone sub-model with others

Parameter	Flaming pyrolysis zone sub-model	Chern (1989)	Milligan (1994)
Char yield, % mass	16.7	20.0	16.7
Tar yield, % by mass	1.0 <sup>a</sup>	-	1.0 <sup>a</sup>
Zone exit temp. (K)	1299	1300	1296
Gas composition, % by vol.			
CO	11.3	13.0	12.1
H <sub>2</sub>	9.7	13.0	10.6
CO <sub>2</sub>	12.1	12.0	11.6
H <sub>2</sub> O	19.3	21.0	18.7
N <sub>2</sub>	45.6	41.0	45.0
CH <sub>4</sub>	2.0 <sup>a</sup>	0.0	2.0 <sup>a</sup>

(Moisture content of 10% (d.b.) and an air fuel ratio of 1.8 by mass were used for comparison. a = assumed)

### 6.3 Parametric Study of Flaming Pyrolysis Zone Sub-Model

One of the problems of the NERD gasifier was the failure of the insulation materials to withstand the temperature in the oxidation zone. It was therefore considered necessary to study the parameters which affect the gas temperature of the combustion zone. Chern (1989) identified four such parameters that affect the gas temperature and they are air fuel ratio, heat loss of gasifier, char yield and moisture content of fuel wood. A parametric study has been conducted to investigate the effects of three of these variables on the temperature of the product gas leaving the flaming pyrolysis zone. The effect of moisture content on the temperature and product gas composition is discussed in the section describing the gasification zone sub-model. Of the three operating parameters investigated in this section, air fuel ratio is the main variable affecting downdraft gasifiers, and the heating value of the product gas mainly depends on this ratio. Heat loss is another important parameter which effects the reactor temperature. The rate of heat loss is determined primarily by the level of the insulation on the gasifier. Char yield, which mainly depends on the nature of feed analysis has a greater influence on the pyrolysis reactions and thus on the heating value of the output gas. Chern (1989) stated that the yield could also change slightly depending on the operating conditions of the gasifier.

Walawender et al. (1985) used an air fuel ratio of 1.4 for a commercial type downdraft gasifier to study the material balance whereas Milligan (1994) has used higher air fuel ratios ranging from 2.0 to 3.0 for his downdraft gasifier experiments. An air fuel ratio of 3.0 is not typical for downdraft gasifiers because lean mixtures produce gas with poor combustible properties due to high dilution of nitrogen concentration. Milligan (1994) has used lean mixtures for some of his experiments probably to investigate their effects on gasifier performances. An air fuel ratio range of 1.2 to 2.4 has been therefore assumed for this parametric study. Chern (1989) stated that heat loss of 10% is acceptable for a commercial downdraft gasifier although a lower figure is desirable. Milligan has reported from the results of the experiments that heat losses can be as high as 16.6%. He further stated that 48% of the total heat loss comes from the flaming pyrolysis zone. Char yield mainly depends on the amount of fixed carbon in the fuel wood. Generally this figure varies between

12.8 to 21.0% by weight depending on the nature of the fuel wood. (Reed & Das 1988). Table 6.2 shows the ranges of operating parameters used for this parametric study.

Table 6.2: Typical range of parameters used for parametric study

Parameter	Range
Air fuel ratio	1.2-2.4
Heat loss (% input energy)	0-20
Char yield (% by mass)	10-30

### 6.3.1 Effects of operating parameters on gas temperature

The effect of air fuel ratio on gas temperature is given in Figure 6.1. It can be seen from the figures that the gas temperature increases with increasing air fuel ratio. High air fuel ratio increases the amount of oxygen input and thus the degree of oxidation of volatiles. These volatiles in return convert more chemical energy into sensible energy and this results in a higher temperature.

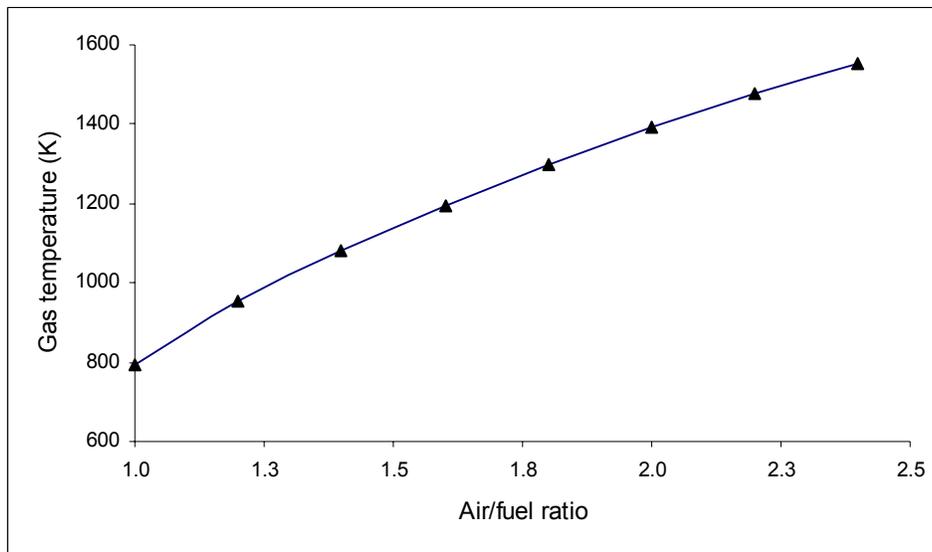


Figure 6.1: Effects of air fuel ratio on gas temperature

The effect of heat loss on gas temperature is given in Figure 6.2. It shows that there is a temperature drop of 40K when heat loss increases by 1%. High heat loss is not desirable for complete pyrolysis as lower temperatures tend to reduce pyrolysis reaction rates. Reed & Das (1988) have shown that temperature of around 900K is favourable for complete pyrolysis. Thus a heat loss of higher than 20% does not seem to be favourable for pyrolysis.

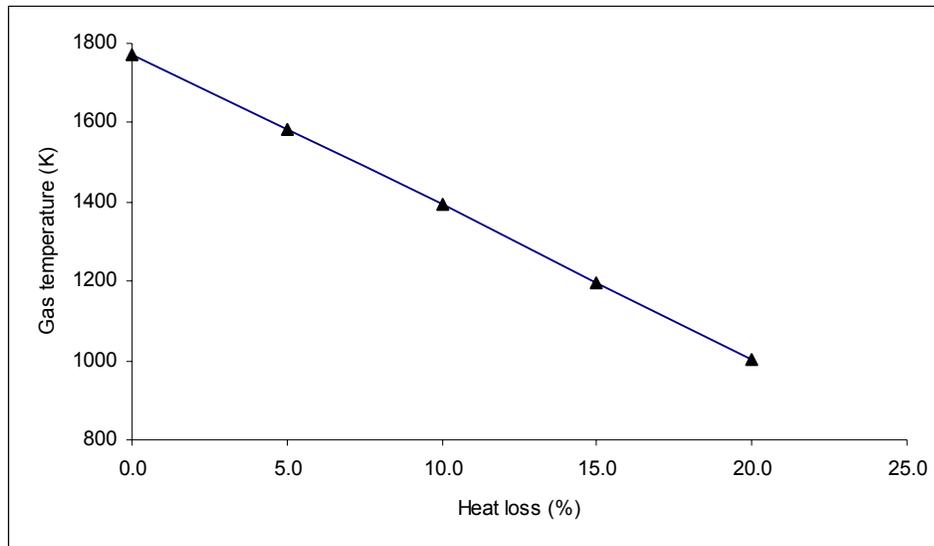


Figure 6.2: Effects of heat loss on gas temperature

Figure 6.3 shows the relationship between char yield and gas temperature. As mentioned in Section 6.3, char yield mainly depends on the amount fixed carbon present in the fuel and higher fixed carbon corresponds with high temperature of the gas. Char yield also depends on the rate at which pyrolysis reactions take place. Milligan (1994) stated that slow pyrolysis increases the char yield and, being an exothermic reaction, results in higher temperatures. Moreover, high char yield increases the chemical and sensible energy of char but decreases the chemical energy of volatiles due to the low amount of volatiles in the gas. This may also result in a higher temperature of the system.

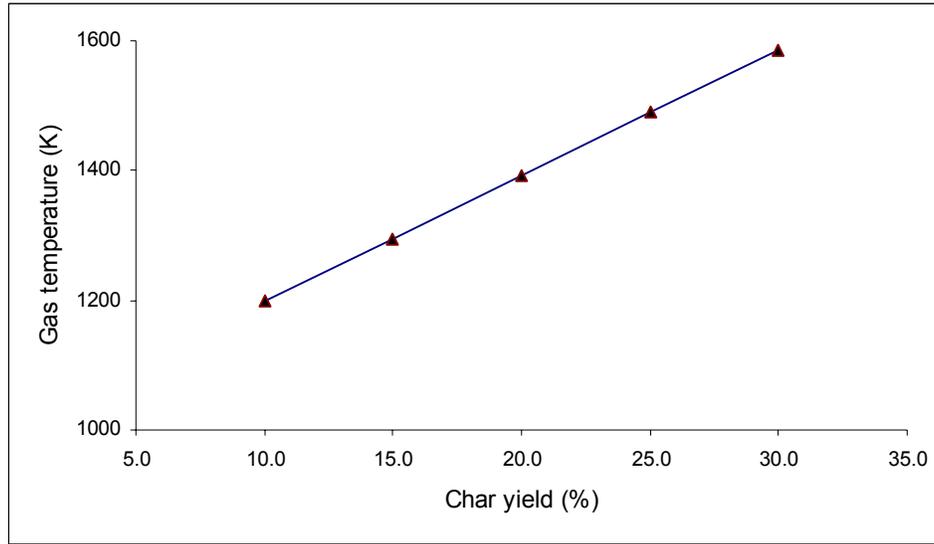


Figure 6.3: Char yield vs gas temperature

### 6.3.2 Summary of parametric study

Based on the simulation results of the flaming pyrolysis zone sub-model, for the NERD gasifier, it can be concluded that low air fuel ratios below 1.4 should be avoided as they tend to decrease the temperature of the flaming pyrolysis zone and result in incomplete pyrolysis. On the other hand, higher air fuel ratios make the product gas lean due to the dilution of the higher nitrogen concentration. Experiments conducted by Milligan (1994) with an air fuel ratio of 3.0 show a lower heating value of the product gas with  $3.5 \text{ MJ/Nm}^3$  whereas the typical range for commercial downdraft gasifiers is  $4.5\text{-}5.0 \text{ MJ/Nm}^3$  (Quaak et al. 1999). So an air fuel ratio more than 2.4 is not desirable as it increases the temperature above  $1500\text{K}$  which has adverse effects on the throat insulation and lining and also makes the gas lean. High fixed carbon content increases the temperature of the flaming pyrolysis zone. Thus feeds with fixed carbon content more than 30% are not desirable for downdraft (throated) gasifiers. High heat losses decrease the flaming pyrolysis zone temperature and lead to incomplete pyrolysis as in the case of low air fuel ratio. So a heat loss of more than 20% of input energy is also not acceptable as it tends to decrease the gas temperature below  $900\text{K}$ . The following section discusses the energy balance and

material balance to determine the actual heat loss of the system. This heat loss figure is then used in the flaming pyrolysis zone sub-model to present the predictions of the sub-model for the NERD gasifier.

## **6.4 Energy and Material and Balances**

The energy and material balances are carried out to study the input and output energy flows and material distribution in the gasifier. The output energy of the gasifier consists of both chemical and sensible energy of the product gas, char, tar, water and heat losses. Thus an energy balance can identify the actual heat loss of the system, if all the other products can be quantified accurately. As mentioned in Section 6.3, the heat loss, which depends on the extent of insulation and the gasifier design has significant effects on the gasifier performance. The overall heat loss of a particular gasifier is reflected in the gasifier efficiency. So it is vital to know the magnitude of heat loss in order to assess its effect on gasifier performance.

The following section describes the detailed calculation of the energy and material balances. In calculating the energy balance, the sensible energy of wood, wood moisture (mc) and air moisture is neglected as the input temperature ( $\sim 27^{\circ}\text{C}$ ) is approximately equal to the reference temperature ( $25^{\circ}\text{C}$ ). Tar production, heat losses through water vapour in the output gas and ash are also not considered. For the material balance, tar and water vapour in the product gas have been neglected. Ash is assumed to be inert. Apart from these assumptions, actual measurements were used for other parameters. Due to the high cost of running the gasifier, energy and material balances are carried out only for two test runs. Test runs 12 and 13 were used for these analysis and the results of all the test runs are given in Appendix G.

### **6.4.1 Energy balance**

The energy balance equation of the gasifier is written as

Input energy = Output energy + Heat loss

The energy inputs to the gasifier are chemical energy of wood, latent heat of wood moisture and air moisture. The wood moisture and air moisture are calculated from the wet wood feed rate and air flow rate as follows.

Wood moisture in = Wet wood feed rate  $\times$   $m_{c_{wb}}$  (kg/hr)

Dry wood feed rate = Wet wood feed rate – wood moisture in (kg/hr)

Air moisture in = Air flow rate  $\times$  air moisture (kg/hr)

(Air moisture is calculated from the psychometric chart for given wet and dry bulb temperatures).

Chemical energy of wood = Higher heating value (HHV) of wood (d.b.)  $\times$   
dry wood feed rate (MJ/hr)

Latent heat of air moisture = Latent heat (LH) of water  $\times$  air moisture in (MJ/hr)

Latent heat of wood moisture = LH  $\times$  wood moisture in (MJ/hr)

The measured HHV of wood and latent heat of water are:

HHV of wood (d.b.) = 19.55 MJ/kg

LH of water = 2.55 MJ/kg

The energy outputs from the gasifier are the chemical and sensible energy in the product gas, tar and char, latent heat of condensation of water and tar, and heat losses. Since the tar production has been neglected for this study, the chemical, sensible and latent heats of tar are not considered. Perry et al. (1997) has given the HHV values of CO, H<sub>2</sub> and CH<sub>4</sub> gases at 298 K as follows.

HHV of CO = 11.97 MJ/m<sup>3</sup>

HHV of H<sub>2</sub> = 12.10 MJ/m<sup>3</sup>

HHV of CH<sub>4</sub> = 37.69 MJ/m<sup>3</sup>

The chemical energy of the dry gas is calculated by summing the products of volume fractions of CO, H<sub>2</sub> and CH<sub>4</sub>, and their respective heating values. The chemical energy flow rate is then calculated by multiplying the heating value of the dry gas by the gas flow rate. The heating value of char is taken as 0.344 MJ/kg (Gaur & Reed, 1998). The chemical energy of char is calculated by multiplying the char output with its heating value.

Sensible energy of the product gas is calculated by the summing the products of volume fractions of CO, H<sub>2</sub>, CO<sub>2</sub>, H<sub>2</sub>O, CH<sub>4</sub> and N<sub>2</sub>, and their respective mean specific heat capacities given in Appendix B. The specific heat of char is taken as 1.013 kJ/kg K (Perry et al. 1997). The sensible energy of char is then calculated by multiplying the char output by its specific heat. The latent heat of water is calculated by multiplying water output by its specific heat which is 4.184 kJ/kg K, and temperature.

The main heat losses from the gasifier wall to the surroundings are by convection and radiation. The basic equations used in calculating the heat losses are given in Appendix H. The total heat loss is taken as the summation of convective and radiation heat losses.

The summary of the results of the energy balance is given in Tables 6.3. It can be seen from this table that the calculated heat loss of the system is in the range of 10.5 – 12.5 % of the energy input. Chern (1989) stated that a heat loss of 10% is acceptable for a downdraft gasifier system of commercial type. Moreover experiments carried out by Milligan (1994) show that the heat losses of downdraft gasifiers are in the range of 6 to 16.6% of the energy output. The unaccounted energy is between 5.0-5.5% of input energy. In calculating the output energy, the amount of energy carried by tar, water vapour in product gas and ash were not considered although tar is said to be responsible for around 4% of the output energy (Milligan 1994). So it can be concluded that the heat loss of the NERD gasifier is between 11.5–14.0% of the input energy of the system.

Table 6.3: Energy balance of the gasifier

Energy input (MJ/hr)			Energy output (MJ/hr)			Calculated heat loss (MJ/hr)	Unaccounted energy (MJ/hr)
Feed	Wood Moisture	Air moisture	Gas	Char	Water		
363.0	8.7	2.4	311.5	2.7	0.1	39.39	20.41
402.7	7.7	1.9	337.0	2.9	0.1	51.13	21.17

#### 6.4.2 Material balance

The material inputs to the gasifier are dry air, dry wood (ash free), wood moisture, air moisture and ash. The material flow of inputs is calculated from the wet wood feed rate and air flow rate as follows.

$$\begin{aligned} \text{Wood moisture in} &= \text{Wet wood feed rate} \times mc_{wb} \text{ (kg/hr)} \\ \text{Dry wood feed rate} &= \text{Wet wood feed rate} - \text{wood moisture in (kg/hr)} \\ \text{Ash in} &= \text{Dry wood rate} \times \text{ash fraction (kg/hr)} \\ \text{Net dry wood feed rate} &= \text{Dry wood feed rate} - \text{ash in (kg/hr)} \\ \text{Air moisture in} &= \text{Air flow rate} \times \text{air moisture (kg/hr)} \end{aligned}$$

(Air moisture is calculated as mentioned in the energy balance).

The material outputs from the gasifier are dry gas, water, char, tar and ash. Since tar, water vapour in the product gas and ash production have been neglected for this analysis, the outputs of these products have been given as the materials unaccounted for. The mass flow rates of char and water are calculated by the division of yield and test duration. The summary of the results of material balance is given in Tables 6.4. It can be seen from this table that material closure is 91 to 93% and the unaccounted materials are in the range of 7 to 9% of total output mass.

Table 6.4: Material balance of the gasifier

Material input (kg/hr)					Material output (kg/hr)				
Air	Feed	Wood moisture	Air moisture	Ash	Gas	Water	Char	Ash	Unaccounted material
55.6	18.6	3.4	1.0	0.1	71.0	1.4	0.7	0.1	5.5
55.6	20.9	2.6	1.0	0.1	69.9	1.2	2.1	0.1	6.9

### 6.5 Flaming Pyrolysis Zone Sub-Model Output for NERD Gasifier

Having established that the predictions of the flaming pyrolysis zone sub-model are satisfactory in Section 6.2, this sub-model is then used to predict the gas compositions and the gas temperature of the NERD gasifier for a moisture content of 18.5% (d.b.), air fuel ratio of 2.03 by mass and char yield of 19.2% by mass. The moisture content and air fuel ratio were taken from the first set of experimental results given in Table 5.5 and the char yield was taken from Table 5.1 in Chapter Five. The energy balance carried out in Section 6.4 shows that the average heat loss of the gasifier is 12.8% of the total energy input. The mass of tar was assumed to be 1% feed input based on the average experimental value reported by Milligan (1994). The values of input parameters and the results predicted by the flaming pyrolysis zone sub-model are given in Table 6.5 and Table 6.6 respectively.

Table 6.5: Input parameters used for NERD gasifier

Parameter	Value
Rubber wood feed analysis	$\text{CH}_{1.542}\text{O}_{0.623}$
Moisture, % d.b.	18.5
Air fuel ratio by mass	2.03
Heat loss, % of energy in	10.5
Char yield, %	19.2
Tar yield, %	1.0

Table 6.6: Sub-model predictions for NERD gasifier

Parameter	Sub-model prediction
Zone exit temp. (K)	1275
Gas composition, % by vol.	
CO	5.1
H <sub>2</sub>	11.1
CO <sub>2</sub>	13.6
H <sub>2</sub> O	25.5
N <sub>2</sub>	43.3
CH <sub>4</sub>	1.4

## 6.6 Verification and Calibration of Gasification Zone Sub-Model

Having verified the flaming pyrolysis zone sub-model predictions of the input data required for the gasification zone sub-model, this section outlines the verification and calibration of the gasification zone sub-model. The sub-model is first verified to compare its predictions with other models. The verification is done using the experimental results reported by Milligan (1994), Hoi et al. (1992) and Groeneveld (1980). Next the sub-model is calibrated with the experimental data collected from the NERD gasifier. Sub-model calibration is carried out for the output gas composition and axial temperature profile in the gasification zone and then it is used to study to investigate the effects of the main operating and design parameters on conversion efficiency of the NERD gasifier.

### 6.6.1 Verification of gasification zone sub-model

In order to verify the results, the predictions of the gasification zone sub-model are compared with the experimental results cited in the literature. The composition of the product gas predicted by this sub-model is given in Tables 6.7, 6.8 and 6.9, and compared with the experimental results reported by Milligan (1994), Hoi et al. (1992) and Groeneveld (1980) respectively for different feed materials and operating

conditions. It can be seen from these tables that the gas composition predicted by the gasification zone varies within 5% except for CO<sub>2</sub> composition, which has a variation between 5 to 9%. One of the reasons for this variation could be the amount of heat loss assumed for the comparison. Except for Milligan (1994), the amount of heat loss was not reported by others. Milligan (1994) has used a heat loss of 10.5% of the total input energy, and the same figure was used for this comparison. Other factors that lead to this variation were the unavailability of reactor diameter, bulk density of feed and the reactor load which were assumed. However despite these differences, the predictions of the gasification zone sub-model is satisfactory when compared to other results.

Table 6.7: Comparison of sub-model results with Milligan’s experimental data

Parameter	Gasification zone sub-model	Milligan (1994)
Dry gas composition (% by vol.)		
CO	20.1	19.7
H <sub>2</sub>	17.4	16.8
CO <sub>2</sub>	11.4	12.3
N <sub>2</sub>	49.2	49.3
CH <sub>4</sub>	1.9 <sup>a</sup>	1.9
Conversion efficiency (%)	59.4	59.2

(Feed analysis of CH<sub>1.46</sub>O<sub>0.65</sub>, moisture content of 11.9% (d.b.) and an air fuel ratio of 2.11 by mass were used for comparison. a = assumed)

Table 6.8: Comparison of sub-model results with Hoi's experimental data

Parameter	Gasification zone sub-model	Hoi et al. (1992)
Dry gas composition (% by vol.)		
CO	19.6	18.9
H <sub>2</sub>	17.5	18.4
CO <sub>2</sub>	11.4	12.3
N <sub>2</sub>	50.9	49.8
CH <sub>4</sub>	0.6 <sup>a</sup>	0.6
Conversion efficiency (%)	53.3	53.1

(Feed analysis of CH<sub>1.54</sub>O<sub>0.62</sub>, moisture content of 7.9% (d.b.) and an air fuel ratio of 1.95 by mass were used for comparison. a = assumed)

Table 6.9: Comparison of sub-model results with Groeneveld's experimental data

Parameter	Gasification zone sub-model	Groeneveld (1980)
CO	20.2	19.6
H <sub>2</sub>	17.8	18.6
CO <sub>2</sub>	11.4	12.4
N <sub>2</sub>	49.6	48.4
CH <sub>4</sub>	1.0 <sup>a</sup>	1.0
Conversion efficiency (%)	56.2	57.9

(Feed analysis of CH<sub>1.49</sub>O<sub>0.68</sub>, moisture content of 14.0% (d.b.) and an air fuel ratio of 2.15 by mass were used for comparison. a = assumed)

### 6.6.2 Calibration of gasification zone sub-model

Having verified the predictions of the gasification zone sub-model against the experimental results given in the open literature, this section discusses the calibration of the sub-model against the experimental data collected during the field experiments for the NERD gasifier. The sub-model was calibrated for the composition of gas

leaving the gasification zone and for the axial temperatures along the gasification zone. The results of the gas compositions for chip sizes varying from 3.3 to 5.5cm were used for calibration. In calibrating the sub-model, the prediction of the amount of methane was adjusted in such way that it is equal to the amount of methane present in the product gas. Tables 6.10, 6.11 and 6.12 compare the sub-model's predictions with experimental results for rubber wood. It can be seen from these tables that the gas composition predicted by the gasification zone sub-model vary within 5% except for H<sub>2</sub> composition of chip size 3.3 cm ,which is 10.5%.

Table 6.10: Comparison of sub-model and experiment results for 5.5 cm chip size

Dry gas composition (% by vol.)	Gasification sub- model	Experiment
CO	18.3	19.1
H <sub>2</sub>	16.4	15.5
CO <sub>2</sub>	11.1	11.4
N <sub>2</sub>	53.2	52.9
CH <sub>4</sub>	1.1 <sup>a</sup>	1.1

(Moisture content of 14.7% (d.b.) and an air fuel ratio 1.89 by mass were used. a = assumed)

Table 6.11: Comparison of sub-model and experiment results for 4.4 cm chip size

Dry gas composition (% by vol.)	Gasification zone sub- model	Experiment
CO	19.1	18.4
H <sub>2</sub>	16.1	17.0
CO <sub>2</sub>	11.1	10.6
N <sub>2</sub>	52.6	52.7
CH <sub>4</sub>	1.3 <sup>a</sup>	1.3

(Moisture content of 16.0% (d.b.) and an air fuel ratio 1.96 by mass were used. a = assumed)

Table 6.12: Comparison of sub-model and experiment results for 3.3 cm chip size

Dry gas composition (% by vol.)	Gasification zone model	Experiment
CO	20.7	19.6
H <sub>2</sub>	15.4	17.2
CO <sub>2</sub>	10.2	9.9
N <sub>2</sub>	52.3	51.9
CH <sub>4</sub>	1.4 <sup>a</sup>	1.4

(Moisture content of 18.5% (d.b.) and an air fuel ratio 2.03 by mass were used. a = assumed)

The gasification zone sub-model is also calibrated for axial temperatures for air fuel ratios of 2.03, 2.20 and 2.37. Figures 6.4, 6.5 and 6.6 compare the sub-model predictions with experimental results for chip size of 3.3 cm. It can be seen from these figures that the experimental temperatures deviate from the sub-model predictions around the exit point of the NERD gasifier, which is 22 cm below the throat, and the correlation coefficient lies between 0.75 to 0.78. The main reason for this temperature deviation is the high heat loss occurring in the bottom of the ash pit. Heat loss calculated for the energy balance in Section 6.4.1 shows that around 50% of the heat loss of the gasifier is taking place in the gasification zone. Apart from this trend at the exit, temperature deviation can also be seen in the middle of the zone which has an average error of 6%. Despite these discrepancies, the predictions of the gasification zone sub-model were deemed to be satisfactory to study the performance trends of the NERD gasifier. The following section discusses the parametric study to investigate the effects of operating and design parameters of the NERD gasifier.

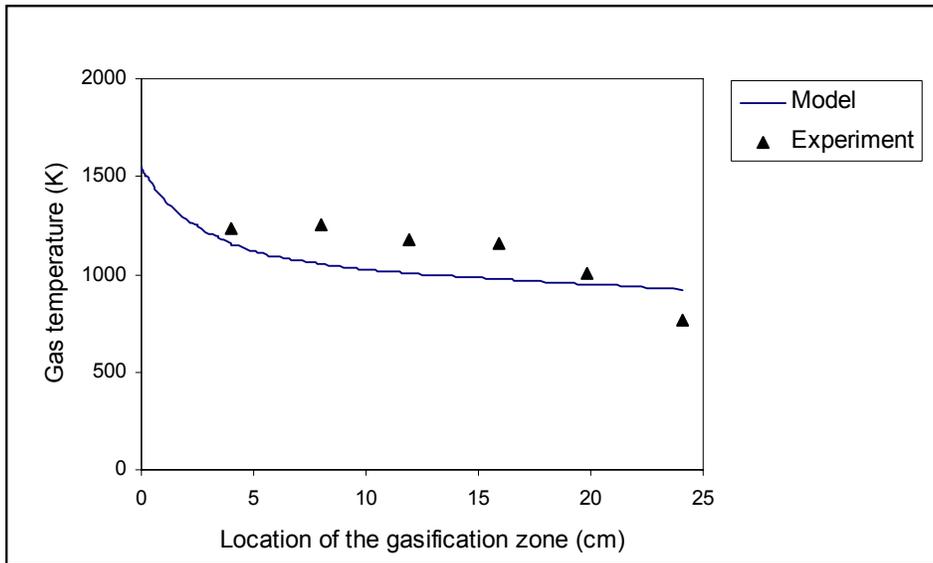


Figure 6.4: Comparison of axial temperature profile for air fuel ratio of 2.03

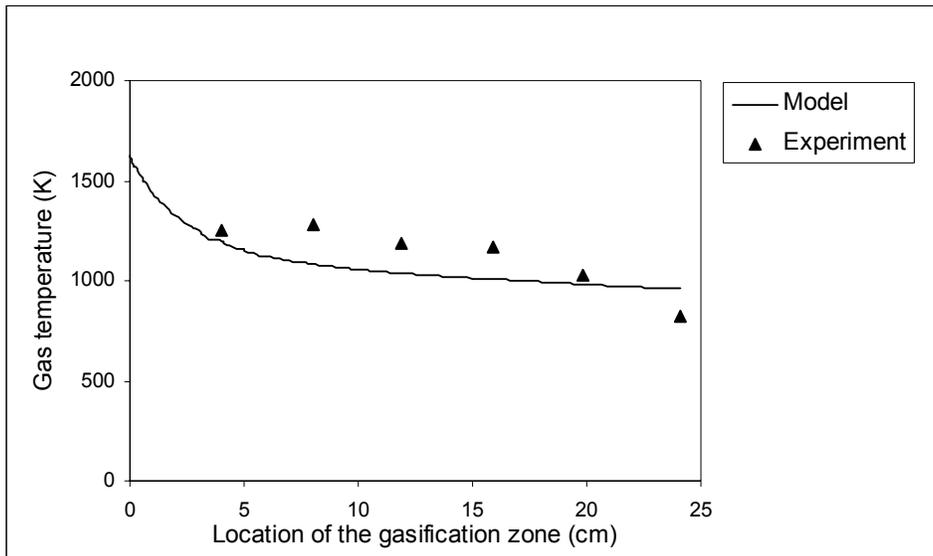


Figure 6.5: Comparison of axial temperature profile for air fuel ratio of 2.20

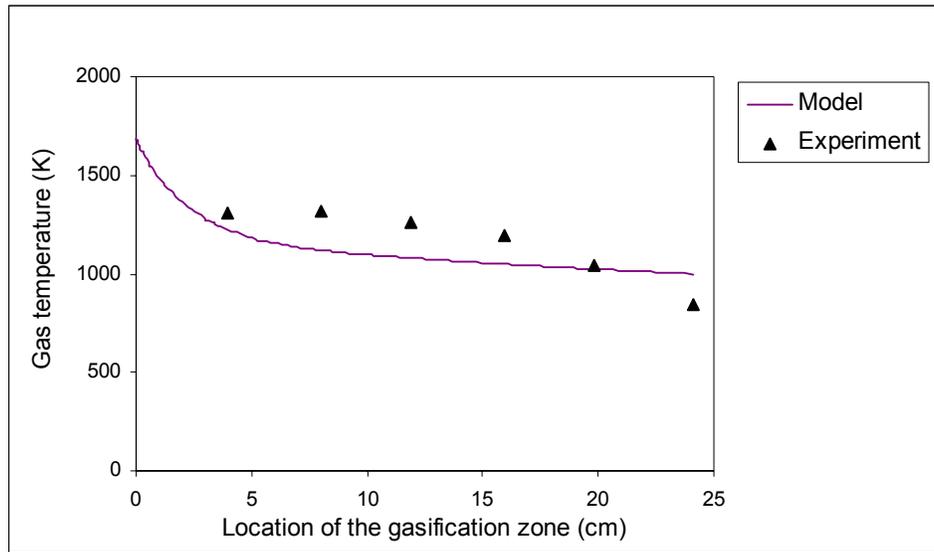


Figure 6.6: Comparison of axial temperature profile for air fuel ratio of 2.37

### 6.7 Parametric Study of Gasification Zone Sub-Model

The main problem associated with the NERD gasifier was the lack of knowledge of the parameters for a maximum efficiency. Conversion efficiency depends on the operating and design parameters of the gasifier, so it is necessary to study the effects of these parameters in detail. A set of computer simulations have been conducted to investigate the effects of operating parameters such as chip size, moisture content, inlet air temperature and design parameters such as heat loss and throat angle on conversion efficiency. Of the five parameters investigated in this section, the moisture content of fuel wood and heat loss are the main variables affecting downdraft gasifiers. Moisture content and heat loss have greater effects on reactor temperature and hence on the conversion efficiency. The size of fuel wood chip is another important parameter which affects the char conversion rate and hence the conversion efficiency. Chen (1987) mentioned that higher inlet air temperature is beneficial for conversion efficiency but the cost of this benefit needs to be investigated. Throat angle is a special unique feature of downdraft gasifiers and the effect of this on conversion efficiency is important.

Generally for gasifier operations, fuel wood with a low moisture content is preferred because of its higher gross energy content. Reed and Das (1988) have reported that moisture contents higher than 40% (w.b.) make the product gas too lean for ignition. Although lower moisture contents are preferred to produce a high quality gas, they further stated that it is desirable to maintain a level at least below 25% (w.b.) to produce a combustible gas. In practice, it is not possible to achieve a 0% moisture content but this figure has been included for this study to determine the maximum possible efficiency. According to Chern (1989) for a commercial gasifier, the heat loss is around 10% although the results of the energy balance conducted for the NERD gasifier found the heat loss to be as much as 13%. The sizes of the chips used in gasification process mainly depend on the diameter of the gasifier. Milligan (1994) has used chips of size 0.6 cm for a lab scale gasifier, which has a reactor diameter of 7.5 cm diameter. Whereas Walawender et al. (1985) used chips as large as 4.7 cm for a commercial scale gasifier with a reactor diameter of 60 cm. Generally ambient air at 300K is used for gasification. Besides this, the effect of input air at temperatures of 450K and 600K is also investigated for this study. In selecting the range of the throat angles, the values of 30, 60 and 90 degrees were selected to include the existing angle of 57 degrees. Table 6.13 presents the values of the parameters used for this study.

Table 6.13: Range of parameters used for parametric study

Parameter	Range
Wood moisture content (% d.b.)	0-30
Heat loss (% input energy)	5-15
Wood size (cm)	1-5
Air temperature (K)	300-600
Throat angle (°)	30-90

### 6.7.1 Effect of chip size

The variation of conversion efficiency along the gasification zone for different chip sizes is illustrated in Figure 6.7. It can be seen from this figure that as the chip size increases the length of the gasification zone must also increase before the conversion reaches its maximum. Thus larger particles need longer reactor lengths to achieve the maximum conversion. Char conversion consists of two processes namely, fast conversion and slow conversion. Fast conversion of char takes place at the entrance of the gasification zone due to the fast reaction rate (Chen 1987). Smaller particles are more likely to get converted to gases completely before the slow conversion begins because of their size and fewer diffusional limitations during the reaction process (Chen 1987). Thus gasifiers with shorter reactor lengths need small chips. With the same environmental conditions, larger chips also undergo the same fast conversion but because of their size, complete conversion may not be possible. Larger chips undergo the remaining slow conversion and thus need a longer reactor length prior to leaving the gasification zone. As a result of faster char conversion, smaller chips increase the conversion efficiency compared to larger chips. Sizes of chips used in the NERD gasifier varied from 1 to 5 cm. But in general, 5 cm chips are preferred due to their lower preparation cost which is significant for commercial gasifiers. The gasification zone length of the NERD is 22 cm. Thus it can be seen from the Figure 6.7 that for a gasification zone length of 22 cm, the conversion efficiency is 55% for chip size of 5 cm. If this size is reduced to 3 cm, the conversion efficiency increases by 1%. The optimum gasification zone length is defined as the length at which the efficiency is 99.5% of the ultimate conversion efficiency. For a 5 cm chip size, the optimum length of the gasification zone is approximately 33cm.

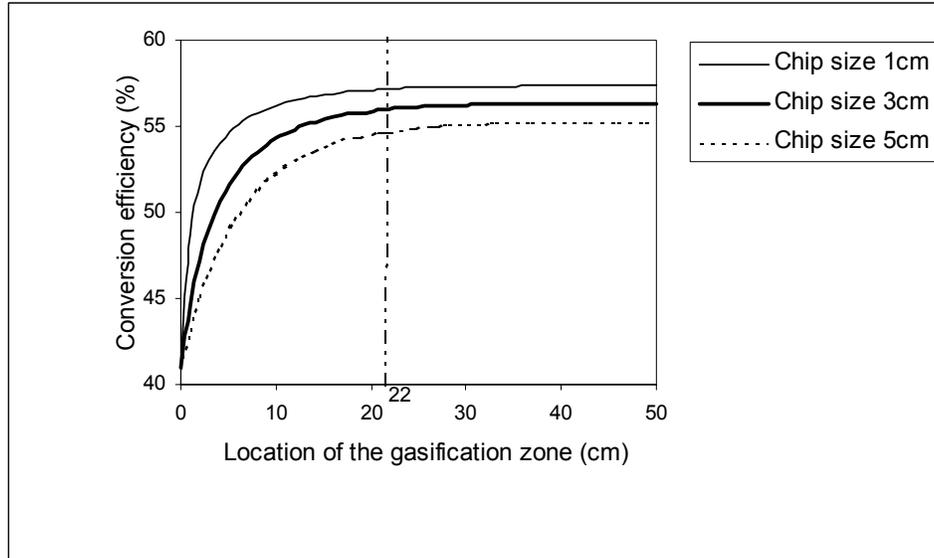


Figure 6.7: Axial variation of conversion efficiency for different chip sizes

### 6.7.2 Effect of moisture content

Figure 6.8 shows the variation of conversion efficiency along the gasification axis for various moisture contents. It can be seen from this figure that the conversion efficiency decreases with increasing moisture contents. This is due to the fact that a higher amount of energy is consumed in evaporating the moisture in the wood which subsequently reduces the temperature of the gas. The lower temperature reduces the reaction rate and thereby the conversion efficiency. The moisture content of the wood chips used in NERD gasifier varied from 12.5 to 18.5 (d.b.). It can be seen from Figure 6.8 that for an average moisture content of 15% (d.b.), the conversion efficiency of the NERD gasifier is 56% at gasification zone length of 22 cm whereas the optimum length occurs at approximately 32 cm.

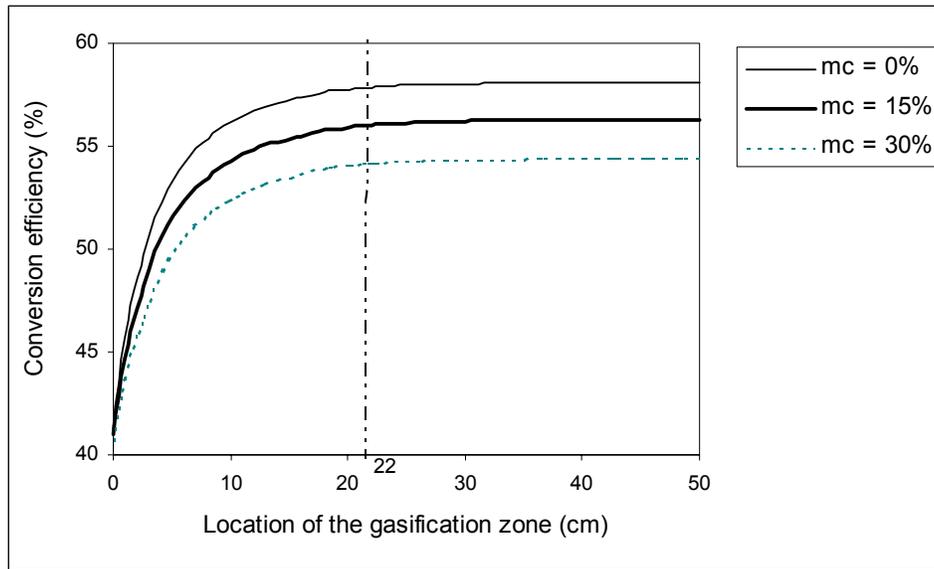


Figure 6.8: Axial variation of conversion efficiency for different moisture contents

### 6.7.3 Effect of inlet air temperature

Gasifiers are generally operated using ambient air at 300K. But Figure 6.9 shows that higher inlet air temperatures are beneficial to gasifier performances. Conversion efficiency increases as the inlet air temperature increases because hot air provides additional enthalpy necessary for reaction thereby decreasing the air fuel ratio. The conversion efficiency increases from 56.0 to 56.5% when the inlet temperature increases from 300 to 450K. For an air input of 50 kg per hour (a typical figure for the NERD gasifier), around 2 kW of energy is required to raise the temperature of air from 300 to 450K. This is approximately 3% of the 70 kW gasifier capacity. It appears that the increase in conversion efficiency is not economical when compared to the amount of energy needed to raise the temperature of input air.

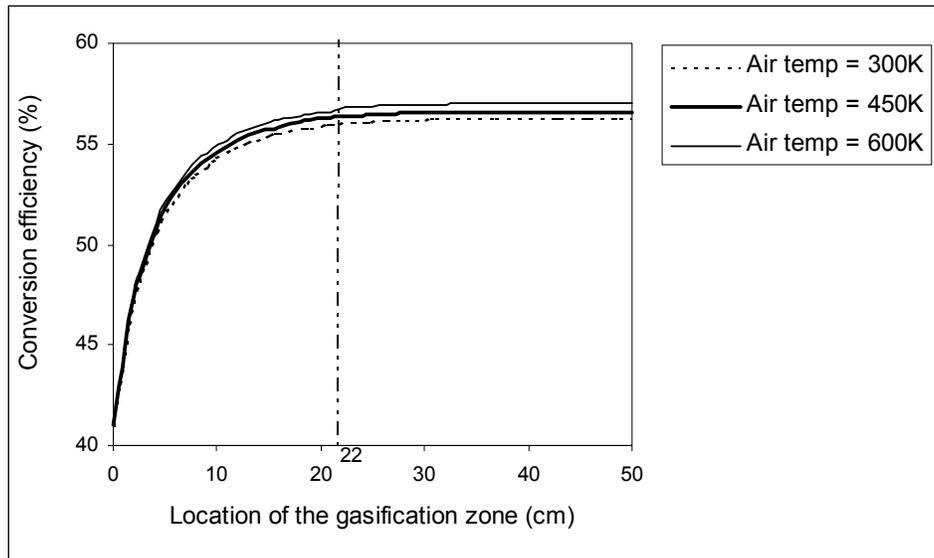


Figure 6.9: Axial variation of conversion efficiency for different air temperatures

#### 6.7.4 Effect of heat loss

The variation of conversion efficiency for varying degrees of heat losses is given in Figure 6.10. It can be seen from the figure that the conversion efficiency decreases by approximately 11 % when the heat loss increases from 5% to 10% or from 10% to 15%. This is because the high heat loss lowers the temperature which in return reduces the gasification reaction rates and thereby the conversion efficiency. The results of the energy balance conducted in Section 6.4.1 indicates a heat loss of up to 12.8% for the NERD gasifier. It can be seen from Figure 6.7 that 12.8% heat loss corresponds to a conversion efficiency of approximately 56% at a gasification zone length of 22 cm but the optimum gasification zone length is approximately 26 cm. Although the decrease in conversion efficiency is considerable for any increase in heat loss, the cost associated in reducing the heat loss (like having the gasifier insulated) must be considered to ensure economic benefit. This is more significant for smaller reactors because in terms of fluxes (based on reactor cross section area) the heat loss is high for gasifiers with smaller reactor diameters as the surface to volume ratio is higher.

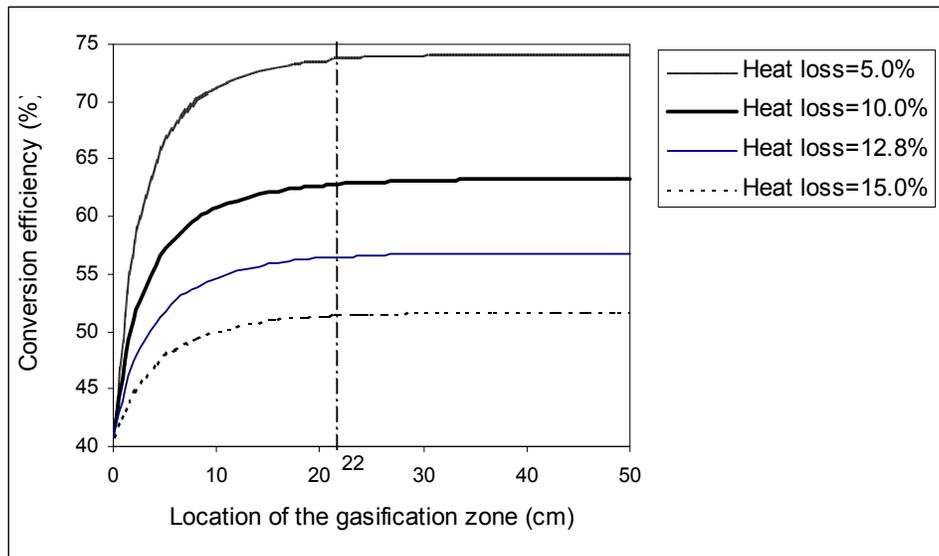


Figure 6.10: Axial variation of conversion efficiency for different heat losses

### 6.7.5 Effect of throat angle

Figure 6.11 shows the variation of conversion efficiency along the gasification zone axis for different throat angles. The smaller angles tend to increase the conversion efficiency whereas larger angles decrease the conversion efficiency, because the latter decreases the temperature due to the divergent effect and hence the reaction rate. Although smaller angles increase the conversion efficiency, it also needs a longer gasification zone length to reach a higher efficiency. The effect of throat angles is not significant until the conversion process reaches the gasification zone length of 10 cm. For a throat angle of 57 degrees (the typical figure of the NERD gasifier) at the gasification zone length of 22 cm, the conversion efficiency is 56%.

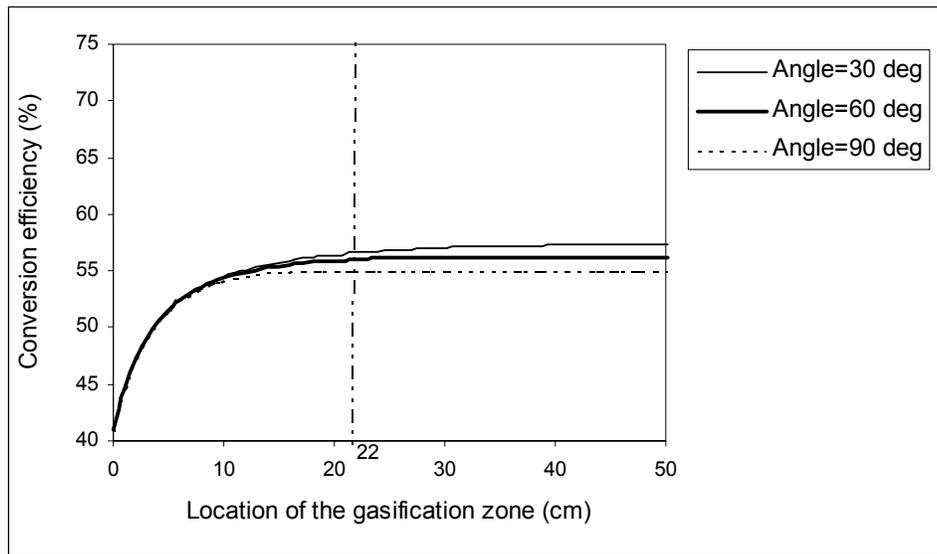


Figure 6.11: Axial variation of conversion efficiency for different throat angles

## 6.8 Conclusions

Chips size has a significant effect on the required length of the gasification zone for complete gasification as well as on conversion efficiency. Although smaller chips increase the conversion efficiency, for a commercial type gasifier they may not be economical due to high fuel wood preparation cost. According to Ziyad Mohamed (1998), fuel wood preparation cost for blocks of 5 cm for the gasifier tested at TRI (BECE gasifier) was 10% of the cost of fuel wood. Larger chips reduce the preparation cost but they would need longer reactor lengths to achieve the same char conversion as smaller chips. So in selecting the appropriate chip size, the cost of wood preparation and the length of the gasification zone need to be considered. Table 6.14 compares the operating and design parameters with efficiency achievable at 22 cm and at optimum gasification zone length. In this comparison the effects of air temperature is not considered as it is not significant (Figure 6.9). It can be seen from this table that chips of 3 cm tend to decrease the maximum efficiency by only 1% compared to chips of 1 cm and also reduces the wood preparation cost.

Table 6.14: Comparison of efficiencies with gasification zone lengths

Parameter	Chip size (cm)			Moisture content (% d.b.)			Heat loss (%)			Throat angle (Deg)		
	1	3	5	0	15	30	5	10	15	30	60	90
Efficiency at the exit (22 cm)	57.2	56.0	54.7	57.8	56.0	54.2	73.8	62.8	51.5	56.6	56.0	55.1
Optimum length (cm)	24	32	33	31	32	35	29	28	25	28	24	19
Efficiency at optimum length	57.3	56.3	55.2	58.1	56.3	54.5	74.2	63.2	51.7	57.4	56.2	55.1

Moisture content of fuel wood also has a significant effect on conversion efficiency. Lower moisture content increases the conversion efficiency and hence it is preferred for gasification process. Moisture in fuel wood can be reduced to a desirable level through external drying before it is fed into the gasifier. But it is not possible to maintain a low moisture level of below 10% as high wood inputs are needed for a commercial type gasifier with a wood capacity of 300 kg. High cost of wood drying is also not economical for commercial gasifiers. If wood chips are given sufficient sun exposure, the moisture content can be reduced to around 15% which gives a conversion efficiency of 56% (Table 6.14).

High inlet air temperature is beneficial for gasifier performance but heating the inlet air is not probably economical when compared with the small gain in conversion efficiency. However, if the temperature of inlet air can be increased by an external means such as waste heat from factories, it may be beneficial for the conversion efficiency of the plant. In tea drying, the temperature of the exhaust gas is around 150° C (Keegel 1983) and this could be used with an aid of an economiser to raise the temperature of inlet air. Lower heat loss is very beneficial for conversion efficiency and this can be achieved by insulating the gasifier reactor. The NERD gasifier appeared to have a reasonable heat loss of 12.8% of the total input energy which is a common figure for this type of reactor. This figure can be reduced further but the cost involved in increasing the level of insulation on the gasifier needs to be compared with the efficiency gain. Smaller throat angle favours the conversion efficiency but at the same time it increases the required length of the gasification zone to achieve the

same conversion efficiency. So the selection of the throat angle has to be made accordance with the length of the gasification zone which is discussed in the next.

From the above study it can be concluded that the length of the gasification zone is a vital design parameter for downdraft gasifiers. A longer length enables the gasifier to operate at its maximum efficiency but it also increases the fabrication cost. Shorter length is not desirable as the gasifier may perform below its design capacity due to insufficient length for char conversion. Optimum gasification zone length has to be selected for maximum output for a given range of operating and design parameters. For the present NERD gasifier configuration with a throat angle of 57 degrees, heat loss of below 15%, chip size between 3 to 5 cm and moisture content of below 15% (w.b.) are the optimum ranges. For these ranges, the optimum gasification zone length of 33 cm would be appropriate for maximum efficiency.

## 6.9 References

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## CHAPTER 7

### CONCLUSIONS

#### 7.1 Summary of Research Findings

This research was conducted with an overall goal of contributing to the improved cost competitiveness of the tea industry and to the reduction in the use of natural resources and CO<sub>2</sub> emissions in Sri Lanka. The main objective to achieve this goal was to assist in the improvement of the downdraft gasifier that has been developed by the NERD Centre for tea drying in Sri Lanka. The main objective was achieved by using a computer program to study of the effects of parameters such as wood chip size, moisture content, inlet air temperature, level of insulation and the throat angle on the conversion efficiency of the downdraft gasifier. The computer program was calibrated using experimental data gathered from the NERD gasifier. Based on the scope of this study, the research found the followings:

1. Gasifiers have the potential to reduce the fuel wood consumption in the tea industry in Sri Lanka by 15%, which equates to a saving in forest cover of 416 ha and carbon dioxide emissions of 307,000 tonnes per annum.
2. The gasifier should be operated at an air fuel ratio of above 1.4 and possibly below 2.4. Air fuel ratios below 1.4 should be avoided as they tend to decrease the temperature of the flaming pyrolysis zone and result in incomplete pyrolysis. Air fuel ratios more than 2.4 make the product gas lean and incombustible due to the diluting effect of nitrogen.
3. High fixed carbon in fuel wood increases the temperature of the pyrolysis zone, which will have undesirable effects on the throat insulation and linings. Thus feed stocks such as barks, with a high percentage of fixed carbon (more than 30%) are not desirable for downdraft (throated) gasifiers.

4. Wood chip size has a significant effect on the length required for char gasification in the gasification zone as well as on conversion efficiency. Although smaller sizes increase the conversion efficiency, it is probably more economical to use larger sizes due to a lower wood preparation cost. Moreover the difference between the conversion efficiencies of smaller (1 cm) and larger (3 cm) chips is only 1%. Therefore chips in the range of 3 to 5 cm can be used for the gasification process with high conversion efficiency.
5. Moisture content of fuel wood also has a significant effect on conversion efficiency. Lower moisture content increases the conversion efficiency and hence it is beneficial to reduce the moisture of fuel wood through sun drying. A moisture content of below 15% is sufficient to achieve an acceptable conversion efficiency.
6. A higher inlet air temperature results in a higher oxidation temperature and a faster reaction rate which in return increases the conversion efficiency. But the increase in conversion efficiency is not significant compared to the energy required to increase the temperature of inlet air. However, in tea drying, a considerable amount of heat is wasted in the drier exhaust gas. This can be used to increase the inlet air temperature with the aid of an economizer.
7. High heat losses decrease the flaming pyrolysis zone temperature and lead to incomplete pyrolysis in the case of a low air fuel ratio. Consequently a high heat loss of more than 15% of the input energy is not desirable for commercial gasifiers and should be avoided by having proper insulation.
8. The NERD gasifier has a heat loss of up to 14.0% of the input energy. If this figure can be reduced further it will increase the conversion efficiency but the cost involved in insulating a commercial type gasifier needs to be balanced against the increased conversion efficiency.
9. The length of the gasification zone is a vital design parameter for downdraft gasifiers. Although longer length increases the possibility of achieving a higher conversion efficiency, care should be taken not to increase the length excessively

as it also increases the reactor fabrication cost. A gasifier may perform below its design capacity if the length is too short. From the simulation results, it appears that a gasification length of 33 cm is appropriate for the NERD gasifier for maximum efficiency.

## **7.2 Limitations of Present Study**

1. A more rigorous theoretical approach should be developed to determine the product gas distribution at the end of the oxidation zone as the output values of the oxidation zone were calculated using energy and mass conversion principles. Although the computational system is acceptable, the product distributions and temperature also depend on values assumed for tar, char yield (same as fixed carbon) and methane.
2. More realistic calculations of heat loss from the zones would improve the predictability of the sub-model. The accuracy of heat loss calculations can be improved by having more external temperature measurements around the gasifier. The actual heat loss figures are very important as they affect the air fuel ratio, the energy balance and the composition of the product gas.
3. The amount of tar was assumed at the end of oxidation zone. The existence of this component may have appeared in the incomplete carbonized char particle and later driven off during the gasification process.

## **7.3 Conclusions**

This study focused on the second most important factor, namely energy use, that causes high production costs to the tea industry in Sri Lanka. It was found that to produce a kilogram of “made” tea in Sri Lanka, an average of 22.4 MJ of thermal energy was used. Eighty five percent of the thermal energy requirement is met by burning fuel wood. This fuel wood dependence is creating concern because local forests are being lost at an alarming rate. In order to reduce the fuel wood

consumption, alternative resources and technologies need to be identified. The available resources have been evaluated in terms of their potential and the quality of energy that can be produced. It was found that of all the energy options that can be used to produce the thermal energy required for the tea industry in Sri Lanka, fuel wood has the highest potential with an annual value of 171.6 PJ.

The development of local technologies available to harness those resources for tea processing were then described. A life cycle cost analysis for the available technologies was carried out to assess their financial feasibility. Based on the lowest life cycle energy cost, it was found that of all the renewable energy resources, burning of biomass in a gasifier is the cheapest source of energy for the tea industry. Moreover wood gasification was the best option. It has the potential to replace the traditional wood fired furnace systems and to reduce the fuel wood consumption in the tea industry in Sri Lanka by 15%. This figure represents a saving in forest cover of 416 ha and carbon dioxide emissions of 307,000 tonnes per annum. Although the local gasifiers appeared to be a better option, problems associated with the refractory lining used in the throat and the lack of knowledge of optimum design and desirable operating parameters have raised doubts of its applicability and productivity for the tea industry. Hence there was a necessity to investigate the performance of the local gasifier to improve its output, construction and use of better materials.

In order to overcome the problems associated with the local gasifier, a computer program was used to investigate gasifier performances. The model consists of two sub-models namely, flaming pyrolysis and gasification zone sub-models. The flaming pyrolysis zone sub-model was used to study the parameters which affect the temperature of the gas leaving that zone. Whereas the gasification zone sub-model was used to investigate the effects of operating and design parameters on gasifier's conversion efficiency. The gasification zone sub-model was calibrated using the experimental data gathered from the NERD gasifier.

Based on the simulation study of the NERD gasifier, it was found that the wood chip size of 3 to 5 cm with moisture content of below 15% is desirable for a commercial version of this downdraft gasifier. Feed material with a higher fixed carbon content of more than 30% and higher heat losses of more than 15% should be avoided. For the

above range of parameters, a gasification zone length of 33 cm was found to be sufficient to achieve a higher conversion efficiency for the NERD gasifier.

#### **7.4 Potential for Future Studies**

1. More studies should be conducted in order to investigate the effects of different throat sizes and ratios of throat to reactor area on the performance of the gasifier. Optimum throat size is necessary to create favourable conditions for tar cracking in the combustion zone.
2. A two dimensional model approach should be adopted to investigate the effects of operating parameters on gasifier performance. Apart from the axial variation, temperature also varies in the radial direction due to the heat transfer through the wall. Since the reaction rate is dependent on temperature, model predictions can be improved if temperature in the radial direction is also considered.

## APPENDIX A

### LIFE CYCLE ENERGY COST OF DIFFERENT SYSTEMS

The following assumptions were used for all systems except for solar system.

Overhead & maintenance cost	=	10% of capital cost
Fuel price inflation( $i_1$ )	=	5% pa
Overhead & maintenance inflation( $i_2$ )	=	10% pa
Market discount (d)	=	12% pa
Life time (N)	=	15 years
Drier capacity	=	200 kg “made” tea/hr
Number of operating hours a day	=	10 hr
Number of working days a year	=	250 days

#### 1. Energy cost of wood fired air heater system

Thermal efficiency of the plant	=	50%
HV of wood at m/c content of 17% (w.b.)	=	14.0 MJ/kg
Wood preparation cost	=	US\$3.00/ton
Wood price	=	US\$30.00/ton
Wood price including preparation cost	=	US\$33.00/ton
Wood consumption by air heater system (de Silva 1994)	=	1.38 kg/kg “made” tea
Fuel wood requirement	=	200*1.38 kg/hr 276 kg/hr
Thus, annual fuel wood cost	=	276*10*250*33/1000 US\$22,770
Energy production over life time	=	276*10*250*15*14.0*0.50 MJ = 72,450 GJ
Initial cost of the air heater	=	US\$15,400
PWF(N=15, $i_2$ =10,d=12)	=	13.64

$$\begin{aligned}
\text{PWF}(N=15, i_1=5, d=12) &= 10.09 \\
\text{Life cycle cost} &= 15400 + 1540 * \text{PWF}(N=15, i_2=10, d=12) \\
&\quad + 22770 * \text{PWF}(N=15, i_1=5, d=12) \\
&= \text{US\$}235,379 \\
\text{Thus energy cost} &= 235,379 / 72,450 \\
&= \mathbf{\text{US\$}3.25/\text{GJ}}
\end{aligned}$$

## 2. Energy cost of the gasifier system

Assumptions for the gasifier system:

$$\begin{aligned}
\text{Drier efficiency} &= 50\% \\
\text{Efficiency of the burner and heat exchanger} &= 80\% \\
\text{Heating value of the gas (at STP)} &= 4.5 \text{ MJ/Nm}^3 \\
\text{Gas yield per kg of wood (at 17\% m/c by w.b.)} &= 2.3 \text{ Nm}^3 \\
\text{Latent heat of water} &= 2.4 \text{ MJ/kg} \\
\text{Conversion efficiency of the gasifier} &= 70\%
\end{aligned}$$

$$\text{The initial moisture content of tea entering the drier} = 66\% \text{ (w.b.)}$$

$$\text{The final moisture content of tea leaving the drier} = 3\% \text{ (w.b.)}$$

For one hour operation

$$\begin{aligned}
\text{Weight of water to be removed} &= 200 * (0.66 - 0.03) / (1.00 - 0.66) \\
&= 370.6 \text{ kg}
\end{aligned}$$

$$\begin{aligned}
\therefore \text{Energy required} &= 370.6 * 2.4 \text{ MJ} \\
&= 889 \text{ MJ}
\end{aligned}$$

$$\begin{aligned}
\text{Then energy input to the drier} &= 889 / 0.50 \\
&= 1778 \text{ MJ}
\end{aligned}$$

$$\begin{aligned}
\text{Energy input to the burner and heat exchanger} &= 1778 / 0.80 \\
&= 2223 \text{ MJ}
\end{aligned}$$

$$\begin{aligned}
\text{Gas requirement} &= 2223 / 4.5 \\
&= 494 \text{ Nm}^3
\end{aligned}$$

$$\begin{aligned}
\text{Thus, wood consumption} &= 494 / 2.3 \\
&= 215 \text{ kg}
\end{aligned}$$

Wood consumption per kg of "made" tea	= 215/200
	= <b>1.07 kg</b>
HV of wood at m/c content of 17% (w.b.)	= 14 MJ/kg
Wood preparation cost	= US\$8.00/ton
Wood price	= US\$30.00/ton
Wood price including preparation cost	= US\$38.00/ton
Wood consumption	= 215 kg/hr
Thus, annual wood cost	= 215*10*250*38/1000
	US\$20,425
Energy production over life time	= 215*10*250*15*14*0.70 MJ
	= 79,013 GJ
Initial cost of the gasifier	= US\$23,000
Life cycle cost	= 23000+2300*PWF(N=15, i <sub>2</sub> =10, d=12)
	+ 20425*PWF(N=15, i <sub>1</sub> =5, d=12)
	= US\$231,202
Thus energy cost	= 231,202/79,013
	= <b>US\$2.93/GJ</b>

### 3. Energy cost of paddy husk combustion system

Thermal efficiency of the plant	= 50%
HV of paddy husk at m/c content of 15% (w.b.)	= 14.4 MJ/kg
Paddy husk cost	= US\$26.6/ton
Paddy husk consumption	= 1.34 kg/kg "made" tea
Paddy husk requirement	= 200*1.34 kg/hr
	268 kg/hr
Thus, annual paddy husk cost	= 268*10*250*33/1000

$$\begin{aligned}
& \text{US\$17,822} \\
\text{Energy production over life time} &= 268*10*250*15*14.4*0.50 \text{ MJ} \\
&= 72,360 \text{ GJ} \\
\text{Initial cost of the air heater} &= \text{US\$15,400} \\
\text{Life cycle cost} &= 15400+1540*\text{PWF}(N=15, i_2=10, d=12) \\
&\quad + 17,822*\text{PWF}(N=15, i_1=5, d=12) \\
&= \text{US\$216,223} \\
\text{Thus energy cost} &= 216,223/72,360 \\
&= \mathbf{\text{US\$3.00/GJ}}
\end{aligned}$$

#### 4. Energy cost of coir dust combustion system

$$\begin{aligned}
\text{Thermal efficiency of the plant} &= 50\% \\
\text{HV of coir dust at m/c content of 20\% (w.b.)} &= 12.6 \text{ MJ/kg} \\
\text{Coir dust cost} &= \text{US\$36.9/ton} \\
& \\
\text{Coir dust consumption} &= 1.53 \text{ kg/kg "made" tea} \\
\text{Coir dust requirement} &= 200*1.53 \text{ kg/hr} \\
&\quad 306 \text{ kg/hr} \\
\text{Thus, annual coir dust cost} &= 306*10*250*36.9/1000 \\
&\quad \text{US\$28,229} \\
\text{Energy production over life time} &= 306*10*250*15*12.6*0.50 \text{ MJ} \\
&= 72,293 \text{ GJ} \\
\text{Initial cost of the air heater} &= \text{US\$15,400} \\
\text{Life cycle cost} &= 15400+1540*\text{PWF}(N=15, i_2=10, d=12) \\
&\quad + 28,229*\text{PWF}(N=15, i_1=5, d=12) \\
&= \text{US\$321,230} \\
\text{Thus energy cost} &= 321,230/72,293 \\
&= \mathbf{\text{US\$4.44/GJ}}
\end{aligned}$$

## 5. Energy cost of micro hydro powered electric air heaters

Efficiency of the electric heater	= 95%
Energy cost	= US\$0.035/kWh
Energy requirement	= 19.3 MJ/kg “made” tea = 5.36 kWh/kg “made” tea
Thus, annual energy cost	= $5.36 * 200 * 10 * 250 * 0.035$ US\$93,800
Energy production over life time	= $19.3 * 200 * 10 * 250 * 15 * 0.95$ MJ = 137,513 GJ
Initial cost of the air heater	= US\$15,000
Life cycle cost	= $15000 + 1500 * PWF(N=15, i_2=10, d=12)$ $+ 93,800 * PWF(N=15, i_1=5, d=12)$ = US\$981,896
Thus energy cost	= $981,896 / 137,513$ = <b>US\$7.14/GJ</b>

## 6. Energy cost of wind powered electric air heaters

Efficiency of the electric heater	= 95%
Energy cost	= US\$0.05/kWh
Energy requirement for drying	= 19.3 MJ/kg “made” tea = 5.36 kWh/kg “made” tea
Thus, annual energy cost	= $5.36 * 200 * 10 * 250 * 0.05$ US\$134,000
Energy production over life time	= $19.3 * 200 * 10 * 250 * 15 * 0.95$ MJ = 137,513 GJ
Initial cost of the air heater	= US\$15,000
Life cycle cost	= $15000 + 1500 * PWF(N=15, i_2=10, d=12)$ $+ 134,000 * PWF(N=15, i_1=5, d=12)$ = US\$1,387,514

Thus energy cost = 1,387,514/137,513  
= **US\$10.09/GJ**

## 7. Energy cost of solar system

Assumptions:

Overhead & maintenance cost	= 3% of capital cost
Overhead & maintenance inflation( $i_2$ )	= 10% pa
Market discount (d)	= 12% pa
Thermal efficiency of the plant	= 40%
Life time	= 20 years
Irradiation	= 15 MJ/m <sup>2</sup>
Energy production over life time/m <sup>2</sup>	= 15*365*20*0.4 43.8 GJ
Collector cost/m <sup>2</sup>	= US\$ 175
PWF(N=20, $i_2$ =10,d=12)	= 15.05
Life cycle cost	= 175+5.25*PWF(N=20, $i_2$ =10,d=12) = US\$254
Thus energy cost	= 254/43.8 = <b>US\$5.80/GJ</b>

## References

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## APPENDIX B

### PROPERTIES OF GASES

Table B.1 Binary diffusivity of various gases

Gas	Binary diffusivity / (cm <sup>2</sup> /s)	Reference temperature (K)
CO <sub>2</sub> - CO	1.480	1200.0
CO <sub>2</sub> - H <sub>2</sub>	0.550	273.0
CO <sub>2</sub> - H <sub>2</sub> O	0.198	307.4
CO <sub>2</sub> - CH <sub>4</sub>	0.153	273.0
CO <sub>2</sub> - N <sub>2</sub>	0.167	298.0
H <sub>2</sub> O - CO	2.690	1200.0
H <sub>2</sub> O - H <sub>2</sub>	1.020	307.2
H <sub>2</sub> O - CO <sub>2</sub>	0.198	307.4
H <sub>2</sub> O - CH <sub>4</sub>	0.292	307.6
H <sub>2</sub> O - N <sub>2</sub>	0.256	307.5

(Source: Satterfield 1970)

Table B.2 Viscosity of various gases

Gas	Viscosity (g/cm s)	Temperature Range (K)	Degree of fit (%)
CO	$(78.9+0.389T-7.50 \times 10^{-5}T^2) \times 10^6$	973-1273	99.9
H <sub>2</sub>	$(28.3+0.220T-5.76 \times 10^{-5}T^2+1.01 \times 10^{-8}T^3) \times 10^6$	773-2000	100.0
CO <sub>2</sub>	$(34.6+0.471T-1.30 \times 10^{-4}T^2+2.00 \times 10^{-8}T^3)$	700-2000	100.0
H <sub>2</sub> O	$(-169.0+0.774T-2.31 \times 10^{-4}T^2+3.33 \times 10^{-8}T^3) \times 10^6$	773-2000	100.0
CH <sub>4</sub>	$(186+0.211(T-573.0)-3.77 \times 10^{-5}(T-573.0)^2) \times 10^6$	573-772	100.0
N <sub>2</sub>	$(78.2+0.422T-1.18 \times 10^{-4}T^2+1.82 \times 10^{-8}T^3) \times 10^6$	700-2000	100.0

(Source: Vargaftik 1996), T in Kelvin

Table B.3 Thermal conductivity of various gases

Gas	Thermal conductivity (W/cm K)	Temperature Range (K)	Degree of fit (%)
CO	$(14.0+0.0526T-2.49\times 10^{-7}T^2)\times 10^{-5}$	700-2000	100.0
H <sub>2</sub>	$(-212.0+1.21T-7.64\times 10^{-4}T^2+2.18\times 10^{-7}T^3)\times 10^{-5}$	773-2000	99.9
CO <sub>2</sub>	$(-134.0+0.46T-3.63\times 10^{-4}T^2+1.07\times 10^{-7}T^3)\times 10^{-5}$	700-2000	99.8
H <sub>2</sub> O	$(-89.6+0.248T-9.25\times 10^{-5}T^2+3.65\times 10^{-8}T^3)\times 10^{-5}$	773-2000	99.6
CH <sub>4</sub>	$1.005\times 10^{-3}$	300-500	100.0
N <sub>2</sub>	$(20.6+0.0372T+6.19\times 10^{-6}T^2+8.0\times 10^{-10}T^3)\times 10^{-5}$	700-2000	100.0

(Source: Vargaftik 1996), T in Kelvin

Table B.4 Enthalpy of various gases

Gas	Enthalpy (kJ/kmol)	Degree of fit (%)
CO	$-8162.0+26.02T+4.49\times 10^{-3}T^2-6.38\times 10^{-7}T^3$	100.0
H <sub>2</sub>	$-8372.0+27.91T+9.41\times 10^{-4}T^2+2.10\times 10^{-7}T^3$	100.0
CO <sub>2</sub>	$-10856.0+32.38T+1.41\times 10^{-2}T^2-2.36\times 10^{-6}T^3$	100.0
H <sub>2</sub> O	$-8998.0+27.91T+7.77\times 10^{-3}T^2-6.61\times 10^{-7}T^3$	100.0
CH <sub>4</sub>	$(-1937.0+04.03T+8.24\times 10^{-3}T^2-1.19\times 10^{-6}T^3)\times 4.184$	100.0
N <sub>2</sub>	$-8127.0+26.02T+4.15\times 10^{-3}T^2-5.47\times 10^{-7}T^3$	100.0

(Source: JANAF 1986), T in Kelvin

## References

JANAF Thermochemical Tables. (1986), ed. M. W. Chase, Third Edition, American Chemical Society and the American Institute of Physics.

Satterfield, C. N. (1970), *Mass Transfer in Homogeneous Catalyst*, MIT Press, Cambridge.

Vargaftik, N. B. (1996), *Handbook of Physical Properties of Liquid and Gases: Pure Substances and Mixtures*, Hemisphere Publishing Corporation, NY.

## APPENDIX C

### GASIFICATION ZONE SUB-MODEL EQUATIONS

The kinetics of gasification of char particle result in the following set of equations (Chen 1987).

$$C_{CO_2}(r) = C_{CO_2,b} \left\{ (r^2 - c_1 r^3) / 3 + g_1(r) - g_2(r) c_1 \right\} B_1 \quad (C.1)$$

$$C_{H_2O}(r) = C_{H_2O,b} \left\{ (r^2 - c_2 r^3) / 3 + g_1(r) - g_2(r) c_2 \right\} B_2 \quad (C.2)$$

$$T(r) = T_b - \left\{ (r^2 - c_3 r^3) / 3 + g_1(r) - g_2(r) c_3 \right\} A \quad (C.3)$$

$$C_c = C_c(0) \exp \left\{ -Kt \left[ (C_{CO_2})^{0.7} + (C_{H_2O})^{0.7} \right] \right\} \quad (C.4)$$

Where:

$$A = \frac{KC_c(r) \left\{ [C_{CO_2}(r)]^{0.7} \Delta H_1 + [C_{H_2O}(r)]^{0.7} \Delta H_2 \right\}}{k_e} \quad (C.5)$$

$$B_1 = \frac{KC_c(r) [C_{CO_2}(r)]^{0.7}}{D_{e,CO_2}} \quad (C.6)$$

$$B_2 = \frac{KC_c(r) [C_{H_2O}(r)]^{0.7}}{D_{e,H_2O}} \quad (C.7)$$

$$g_1(r) = (r_0^2 - r^2) / 2 \quad (C.8)$$

$$g_2(r) = (r_0^3 - r^3) / 3 \quad (C.9)$$

$$c_1 = \frac{r_0 k_{m,CO_2} - D_{e,CO_2}}{r_0^2 k_{m,CO_2}} \quad (C.10)$$

$$c_2 = \frac{r_0 k_{m,H_2O} - D_{e,H_2O}}{r_0^2 k_{m,H_2O}} \quad (C.11)$$

$$c_2 = \frac{r_0 h - k_e}{r_0^2 h} \quad (C.12)$$

$$K = 1.019 \times 10^{11} \exp(-E/RT)$$

$$E = 217.1 \text{ kJ}$$

$$\Delta H_1 = 139.9 \text{ kJ/kmol}$$

$$\Delta H_2 = 172.4 \text{ kJ/kmol}$$

Equations C.1 to C.4 are solved simultaneously to calculate the values of C, CO<sub>2</sub>, H<sub>2</sub>O and temperature. These values are then used to calculate the gas concentration. The gasification zone sub-model formulation at any time step of Δt is given as follows.

The superficial velocity can be written as

$$v_g = \frac{RT_g}{P} (N_{H_2O} + N_{CO_2} + N_{CH_4} + N_{H_2} + N_{CO} + N_{N_2} + N_{trace}) \quad (C.13)$$

The bulk concentration of CO<sub>2</sub> and H<sub>2</sub>O are calculated as follows:

$$C_{CO_2,b} = \frac{N_{CO_2}}{v_g \epsilon_b} \quad (C.14)$$

$$C_{H_2O,b} = \frac{N_{H_2O}}{v_g \epsilon_b} \quad (C.15)$$

Where:

$$\varepsilon_b = 0.5 - 0.2(r_p - r_0) / r_p \quad (\text{C.16})$$

'b' is referred to as gas bulk condition

The mass transfer coefficients are calculated from the equation developed by Sherwood et al. (1975).

$$k_{mA} = 1.17 \text{Re}^{0.585} \text{Sc}^{0.333} D_A / (2r_0) \quad (\text{C.17})$$

Where:

$$\text{Re} = \frac{2r_0 G}{\mu} \quad \text{and} \quad \text{Sc} = \frac{\mu}{\rho D_A}$$

The superficial mass velocity (G) is computed by summing the products of mole flux and molecular weight for all gas components while the gas density is calculated from the division of mass velocity and gas velocity.

Wilke (1950) has proposed the following equations to calculate the viscosity of a gas mixture.

$$\mu_{mix} = \sum_{i=1}^n \frac{\mu_i}{1 + \frac{1}{Y_i} \sum_{j=1}^{j=n} Y_j \varepsilon_{ij}} \quad (\text{C.18})$$

$$\varepsilon_{ij} = \frac{[1 + (\mu_i / \mu_j)^{0.5} (M_j / M_i)^{0.25}]^2}{[8(1 + M_i / M_j)]^{0.5}} \quad (\text{C.19})$$

The diffusivities of CO<sub>2</sub> and H<sub>2</sub>O are calculated according to Wilke's formula as follows.

$$D_{1m} = \frac{1 - Y_1}{\frac{Y_2}{D_{12}} + \frac{Y_3}{D_{13}} + \frac{Y_4}{D_{14}} + \dots} \quad (\text{C.20})$$

$$D_{12} = \frac{0.001858T^{1.5}[(M_1 + M_2)/(M_1M_2)]^{0.5}}{P\sigma_{12}^2\Omega_D} \quad (C.21)$$

$$\Omega_D = \frac{K_B T}{\varepsilon_{12}} \quad (C.22)$$

Where

The binary diffusivity to other temperatures is extrapolated using the following equation.

$$D_{ij,T} = D_{ij,T_r} (T/T_r)^{1.75} \quad (C.23)$$

Heat transfer coefficient is calculated using the following empirical correlation developed by Petrovic & Thodos (1968).

$$j_H = \frac{h}{C_p G} \text{Pr}^{0.667} \quad (C.24)$$

$$j_H = \frac{0.357}{\text{Re}^{0.359} \varepsilon_b} \quad (C.25)$$

Where:

$$\text{Pr} = \frac{C_p \mu}{k}$$

The heat capacity of the mixture ( $C_p$ ) is obtained by summing the product of  $C_p$  and molar fraction of each component. Chen (1987) has quoted from Friend and Adler (1958) that the thermal conductivity of the gas mixture can be calculated using the following.

$$k_m = \frac{\sum Y_i k_i (M_i)^{0.333}}{\sum Y_i (M_i)^{0.333}} \quad (C.26)$$

The concept of gasification of char particle will be used here to calculate the temperature profile and the concentration in the particle. From these profiles the mass and heat fluxes (N) flowing into the particle can be determined. They can be written as follow.

$$N_{CO_2,s} = k_{m,CO_2} (C_{CO_2,b} - C_{CO_2,s}) \quad (C.27)$$

$$N_{H_2O,s} = k_{m,H_2O} (C_{H_2O,b} - C_{H_2O,s}) \quad (C.28)$$

$$N_{heat,s} = h(T_b - T_s) \quad (C.29)$$

Based on particle surface, the production of CO and H<sub>2</sub> are computed by:

$$N_{CO,s} = N_{H_2O,s} + 2N_{CO_2,s} \quad (C.30)$$

$$N_{H_2s} = N_{H_2O,s} \quad (C.31)$$

‘s’ is referred to as condition at particle surface

The particle flux can be written as

$$N_p = N_{char} / (V_p C_C) \quad (C.32)$$

The number of particles per unit cross section area (n<sub>p</sub>) during the given time step and the total particle surface, A<sub>pT</sub> are,

$$n_p = N_p \Delta t \quad (C.33)$$

$$A_{pT} = n_p A_p \quad (C.34)$$

The component molar fluxes based on reactor cross section area can be deduced as follows.

$$N_{CO}(i) = N_{CO}(i-1) + N_{CO,s}A_{pT} \quad (C.35)$$

$$N_{H_2}(i) = N_{H_2}(i-1) + N_{H_2,s}A_{pT} \quad (C.36)$$

$$N_{CO_2}(i) = N_{CO_2}(i-1) + N_{CO_2,s}A_{pT} \quad (C.37)$$

$$N_{H_2O}(i) = N_{H_2O}(i-1) + N_{H_2O,s}A_{pT} \quad (C.38)$$

$$N_C(i) = N_C(i-1) - (N_{CO_2,s} + N_{H_2O,s})A_{pT} \quad (C.39)$$

The mole fluxes of N<sub>2</sub>, CH<sub>4</sub> and traces remain unchanged. The amount of shift ( $w_s$ ) to restore the equilibrium is calculated from the equilibrium constant of the water-gas shift reaction. The equations are as follows.

$$K_3 = \frac{(N_{CO_2}(i) + w_s)(N_{H_2}(i) + w_s)}{(N_{CO}(i) - w_s)(N_{H_2O}(i) - w_s)} \quad (C.40)$$

Gumz (1950) has given the correlation between the temperature and equilibrium constant ( $K_3$ ) for the water gas shift reaction in empirical formula as follows.

$$\text{Log}(K_3) = -36.72508 + \frac{3994.704}{T} - 4.46241 \times 10^{-3} T + 6.71814 \times 10^{-7} T^2 + 12.22028 \log(T) \quad (C.41)$$

So the final fluxes leaving the time step  $i$  are as follows.

$$N_{CO,e} = N_{CO}(i) - w_s \quad (C.42)$$

$$N_{H_2,e} = N_{H_2}(i) + w_s \quad (C.43)$$

$$N_{CO_2,e} = N_{CO_2}(i) + w_s \quad (C.44)$$

$$N_{H_2O,e} = N_{H_2O}(i) - w_s \quad (C.45)$$

The distance that the particle travels during a given time step is calculated from the average velocity expressed as follows.

$$v_p = \frac{N_{C,avg}}{C_{C,avg}(1 - \varepsilon_b)} \quad (C.46)$$

$$\Delta L = v_p \Delta t \quad (C.47)$$

The gas phase temperature at the end of time step  $i$  is computed from a heat balance on solid and gas. The overall heat balance at the time step  $i$  can be written as,

$$(H_{gas} + H_{solid})_i + \Delta H_3 w_s = (H_{gas} + H_{solid})_e + H_{g-s} + H_L \quad (C.48)$$

$$H_{g-s} = (T_b - T_s) A_{pT} \quad (C.49)$$

$$\Delta H_3 = -40.5 \text{ kJ/kmol}$$

' $i$ ' is referred to as inlet condition and ' $e$ ' is referred to as exit condition.

## References

Chen, J. S. (1987), Kinetic Engineering Modelling of Co-Current Moving Bed Gasification Reactors for Carbonaceous Material, PhD thesis, Cornell University, New York.

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Sherwood, T. K., Pigford, R. L. & Wilke, C. R. (1975), *Mass Transfer*, McGraw-Hill Book Company, New York.

Wilke, C. R. (1950), 'Diffusional properties of multicomponent gases', *Chemical Engineering Progress*, vol. 46, pp. 95-104.

## APPENDIX D

### SUBROUTINES USED IN THE COMPUTER PROGRAM

```
SUBROUTINE newt(x,n,check)
  INTEGER n,nn,NP,MAXITS
  LOGICAL check
  REAL x(n),fvec,TOLF,TOLMIN,TOLX,STPMX
  PARAMETER (NP=40,MAXITS=200,TOLF=1.e-4,TOLMIN=1.e-6,TOLX=1.e-7,
*STPMX=100.)
  COMMON /newtv/ fvec(NP),nn
  SAVE /newtv/
CU  USES fdjac,fmin,lnsrch,lubksb,ludcmp
  INTEGER i,its,j,indx(NP)
  REAL d,den,f,fold,stpmax,sum,temp,test,fjac(NP,NP),g(NP),p(NP),
*xold(NP),fmin
  EXTERNAL fmin
  nn=n
  f=fmin(x)
  test=0.
  do 11 i=1,n
    if(abs(fvec(i)).gt.test)test=abs(fvec(i))
11  continue
    if(test.lt..01*TOLF)return
    sum=0.
    do 12 i=1,n
      sum=sum+x(i)**2
12  continue
    stpmax=STPMX*max(sqrt(sum),float(n))
    do 21 its=1,MAXITS
      call fdjac(n,x,fvec,NP,fjac)
      do 14 i=1,n
        sum=0.
        do 13 j=1,n
          sum=sum+fjac(j,i)*fvec(j)
13  continue
        g(i)=sum
14  continue
      do 15 i=1,n
        xold(i)=x(i)
15  continue
      fold=f
      do 16 i=1,n
```

```

        p(i)=-fvec(i)
16      continue
        call ludcmp(fjac,n,NP,indx,d)
        call lubksb(fjac,n,NP,indx,p)
        call lnsrch(n,xold,fold,g,p,x,f,stpmax,check,fmin)
        test=0.
        do 17 i=1,n
            if(abs(fvec(i)).gt.test)test=abs(fvec(i))
17      continue
        if(test.lt.TOLF)then
            check=.false.
            return
        endif
        if(check)then
            test=0.
            den=max(f,.5*n)
            do 18 i=1,n
                temp=abs(g(i))*max(abs(x(i)),1.)/den
                if(temp.gt.test)test=temp
18      continue
            if(test.lt.TOLMIN)then
                check=.true.
            else
                check=.false.
            endif
            return
        endif
        test=0.
        do 19 i=1,n
            temp=(abs(x(i)-xold(i)))/max(abs(x(i)),1.)
            if(temp.gt.test)test=temp
19      continue
        if(test.lt.TOLX)return
21     continue
        pause 'MAXITS exceeded in newt'
        END

```

(Source: Press et al. 1997)

```

*****
SUBROUTINE fdjac(n,x,fvec,np,df)
    INTEGER n,np,NMAX
    REAL df(np,np),fvec(n),x(n),EPS
    PARAMETER (NMAX=40,EPS=1.e-4)
CU   USES funcv
    INTEGER i,j
    REAL h,temp,f(NMAX)

```

```

do 12 j=1,n
  temp=x(j)
  h=EPS*abs(temp)
  if(h.eq.0.)h=EPS
  x(j)=temp+h
  h=x(j)-temp
  call funcv(n,x,f)
  x(j)=temp
  do 11 i=1,n
    df(i,j)=(f(i)-fvec(i))/h
11  continue
12  continue
return
END

```

```

FUNCTION fmin(x)
INTEGER n,NP
REAL fmin,x(*),fvec
PARAMETER (NP=40)
COMMON /newtv/ fvec(NP),n
SAVE /newtv/
CU  USES funcv
INTEGER i
REAL sum
call funcv(n,x,fvec)
sum=0.
do 11 i=1,n
  sum=sum+fvec(i)**2
11  continue
fmin=0.5*sum
return
END

```

(Source: Press et al. 1997)

```

*****
SUBROUTINE lnsrch(n,xold,fold,g,p,x,f,stpmax,check,func)
INTEGER n
LOGICAL check
REAL f,fold,stpmax,g(n),p(n),x(n),xold(n),func,ALF,TOLX
PARAMETER (ALF=1.e-4,TOLX=1.e-7)
EXTERNAL func
CU  USES func
INTEGER i
REAL
a,alam,alam2,alamin,b,disc,f2,fold2,rhs1,rhs2,slope,sum,temp,

```

```

*test,tmplam
check=.false.
sum=0.
do 11 i=1,n
    sum=sum+p(i)*p(i)
11  continue
    sum=sqrt(sum)
    if(sum.gt.stpmax)then
        do 12 i=1,n
            p(i)=p(i)*stpmax/sum
12  continue
        endif
        slope=0.
        do 13 i=1,n
            slope=slope+g(i)*p(i)
13  continue
        test=0.
        do 14 i=1,n
            temp=abs(p(i))/max(abs(xold(i)),1.)
            if(temp.gt.test)test=temp
14  continue
        alamin=TOLX/test
        alam=1.
1  continue
        do 15 i=1,n
            x(i)=xold(i)+alam*p(i)
15  continue
        f=func(x)
        if(alam.lt.alamin)then
            do 16 i=1,n
                x(i)=xold(i)
16  continue
            check=.true.
            return
        else if(f.le.fold+ALF*alam*slope)then
            return
        else
            if(alam.eq.1.)then
                tmplam=-slope/(2.*(f-fold-slope))
            else
                rhs1=f-fold-alam*slope
                rhs2=f2-fold2-alam2*slope
                a=(rhs1/alam**2-rhs2/alam2**2)/(alam-alam2)
                b=(-alam2*rhs1/alam**2+alam*rhs2/alam2**2)/(alam-alam2)
                if(a.eq.0.)then
                    tmplam=-slope/(2.*b)

```

```

        else
            disc=b*b-3.*a*slope
            if(disc.lt.0.) pause 'roundoff problem in lnsrch'
            ttmplam=(-b+sqrt(disc))/(3.*a)
        endif
        if(tmplam.gt..5*alam)tmplam=.5*alam
    endif
endif
alam2=alam
f2=f
fold2=fold
alam=max(tmplam,.1*alam)
goto 1
END

```

(Source: Press et al. 1997)

```

*****
SUBROUTINE lubksb(a,n,np,indx,b)
INTEGER n,np,indx(n)
REAL a(np,np),b(n)
INTEGER i,ii,j,ll
REAL sum
ii=0
do 12 i=1,n
    ll=indx(i)
    sum=b(ll)
    b(ll)=b(i)
    if (ii.ne.0)then
        do 11 j=ii,i-1
            sum=sum-a(i,j)*b(j)
11        continue
        else if (sum.ne.0.) then
            ii=i
        endif
        b(i)=sum
12    continue
do 14 i=n,1,-1
    sum=b(i)
    do 13 j=i+1,n
        sum=sum-a(i,j)*b(j)
13    continue
        b(i)=sum/a(i,i)
14    continue
return
END

```

(Source: Press et al. 1997)

```
*****
      SUBROUTINE ludcmp(a,n,np,indx,d)
      INTEGER n,np,indx(n),NMAX
      REAL d,a(np,np),TINY
      PARAMETER (NMAX=500,TINY=1.0e-20)
      INTEGER i,imax,j,k
      REAL aamax,dum,sum,vv(NMAX)
      d=1.
      do 12 i=1,n
         aamax=0.
         do 11 j=1,n
            if (abs(a(i,j)).gt.aamax) aamax=abs(a(i,j))
11          continue
            if (aamax.eq.0.) pause 'singular matrix in ludcmp'
            vv(i)=1./aamax
12          continue
         do 19 j=1,n
            do 14 i=1,j-1
               sum=a(i,j)
               do 13 k=1,i-1
                  sum=sum-a(i,k)*a(k,j)
13              continue
               a(i,j)=sum
14              continue
            aamax=0.
            do 16 i=j,n
               sum=a(i,j)
               do 15 k=1,j-1
                  sum=sum-a(i,k)*a(k,j)
15              continue
               a(i,j)=sum
               dum=vv(i)*abs(sum)
               if (dum.ge.aamax) then
                  imax=i
                  aamax=dum
               endif
16              continue
            if (j.ne.imax)then
               do 17 k=1,n
                  dum=a(imax,k)
                  a(imax,k)=a(j,k)
                  a(j,k)=dum
17              continue

```

```

        d=-d
        vv(imax)=vv(j)
    endif
    indx(j)=imax
    if(a(j,j).eq.0.)a(j,j)=TINY
    if(j.ne.n)then
        dum=1./a(j,j)
        do 18 i=j+1,n
            a(i,j)=a(i,j)*dum
18        continue
        endif
19    continue
    return
    END

```

(Source: Press et al. 1997)

## References

Press, W. H., Teukolsky, S. A., Vetterling, W. T. & Flannery, B. P. (1997), *Numerical Recipes in Fortran 77, The Art of Scientific Computing*, Second Edition, Press Syndicate, University of Cambridge, Cambridge.

## APPENDIX E

### CODES OF THE COMPUTER PROGRAM

```
C *****
C *
C *           MODELLING PROGRAM FOR DOWNDRAFT GASIFIERS *
C *
C *****
C *
C * THIS PROGRAM PREDICTS THE PROFILES OF GAS COMPOSITION *
C * AND TEMPERATURE ALONG THE AXIS OF THE GASIFICATION ZONE*
C *
C *****

C *****
PROGRAM MODEL PROGRAMME FOR DOWNDRAFT GASIFIER
IMPLICIT NONE
INTEGER I, I1, I2, I3, I4, I5, I6, I7, I8, I9, I10, I11, J, J3, J5, JJ, KK,
$      N1, N2, N3, NN

LOGICAL CHECK

REAL X2(3), F2(3), X1(6), F1(6), X3(1), F3(1), Z, NMOL, KEFF, KMCO2,
$     KMH2O, R(21), R2(21), R3(21), T(21), CCO2(21), CH2O(21),
$     CCI(21), WM(6), EPS(6,6), FRN(6), FX(6), G(6), VS(6), WM1(6),
$     FX1(6), H(6), CP(6), D(6), TC(6), WK2(100), WK(68), WK3(22),
$     TCINSU, THICKI, FXO2W, FUR, FXFUEL, FOMR, FXN2ML, FXO2ML, CF,
$     AIRINPUT, OMOL1, CMOL, HMOL, OMOL, FMOLWT, HHVFD, DNSTY, LHV,
$     DP, RP, DREA, PREA, MC, TAIR, HWET, TWAT, WATMOL, HEVAP, THEVAP,
$     PHLOSS, HHVNET, FRNMOL, TNMOL, TOTMOL, CH4, HCG, P, GC, TL, FI,
$     RREA, VOL1, FXO2W1, FXAIR1, FXFUE, FXCMOL, FXHMOL, FXOMOL,
$     FXNMOL, FXWATM, AH2OML, TARS, EMSVTY, SIGMA, TAMB, PI, PLL, TW,
$     DTT, TDCOMB, HCF, TOTR, HU, DTALL, ASUF, HLO1, ASUFO, HLO, FCTOR1,
$     FXHLO, FXHAIR, HAIR, FXTNL, TB, EPI, DCHAR, CCHPOW, HLORAD, DT, TT,
$     TMFXG, VG, V1, CCO2B, CH2OB, GM, TB2, TB3, FXHCOM, VOIDB, WMR, WMR1,
$     RE, TBR1, TBR2, TBR3, TBR4, SUM, DCO2M, VSR, VSCO, YKM, YM, TCM, CPM,
$     AWM, CPG, PR, RE1, HJ, HCOEFF, TCCHAR, EP, FCTOR, SC1, DECO2, DEH2O,
$     TWOUT, TCAIR, VISK, PRAIR, TF, CVEXP, GRAV, GRD, RA, A1, FXCH4,
$     CPOW2, CNU, HTFREE, HW, THRW, TCSTL, TDREA, TRREA, CIRREA, HTCIN,
$     B30, TDREA1, TRREA1, RR, FXCO2P, FXH2OP, DH2OM, C1, CCH, POW,
$     FXHCH, FXCOP, FXH2P, VOLP, FXPTCL, PN, AP, APPN, FXCML1, SPCH, HCH,
$     SPEEL, TP, VP, DTL, DWA, DTEMP, CROSSA, TW1, HLCC, HLOSS, VSCO1,
```

```

$      CR,HVCO,HFLUX,EFF,DNSTYG,CO,H2,H2O,CO2,TN2,HVH2,HVCH4,
$      FXTNML,HLCC1, HFXW,CHAR,TDCOM,HVFUFX,THVTRC,A2,A3,RP2,
$      RP3,AVGCCH,TG,XCH4,TTL,MWFD,NTRGN,AFR,AIRIN,HL,WPC,WPH,
$      WPO,WPN,DTHRT,RTHRT,ANGLTHRT,FLX,FC,TFX

PARAMETER (N1=6,N2=3,N3=1,PI=3.14159)

COMMON/BK1/ TAIR,CMOL,HMOL,OMOL
$      /BK2/ xCH4,CHAR,MWFD,MC,DP,HHVFD,AFR,HL,FC
$      /BK3/ A1,A2,A3,R,R2,R3,CCI,CCO2B,CH2OB,RP,RP2,RP3,KEFF,
$      DECO2,DEH2O,DT,B30,KMCO2,KMH2O,NN,J
$      /BK4/ TB
$      /BK5/ FX
$      /BK6/ SUM,TP,APPN,HCOEFF,HLOSS,HCH,FXCML1

C      *****
WRITE(*,50)
50     FORMAT(/1X,'WELCOME TO DOWNDRAFT GASIFIER MODELLING PROGRAM',/)

WRITE(*,60)
60     FORMAT(/1X,'THE MODEL WAS ORIGINALLY DEVELOPED BY J. S. CHEN
$(1987) THEN MODIFIED BY TUAN HARIS JAYAH TO SUIT DOWNDRAFT
$THROATED GASIFIERS',/)

WRITE(*,70)
70     FORMAT(/1X,'THIS CODE WAS DEVELOPED BY TUAN HARIS JAYAH UNDER
$THE SUPERVISION OF DR. BOB FULLER, DR. LU AYE & A/PROF. DON
$STEWART, THE UNIVERSITY OF MELBOURNE AUSTRALIA'/)

WRITE(*,80)
80     FORMAT(/1X,'*****',/)

C      THE ULTIMATE ANALYSIS OF 100 KG WOOD

WRITE(*,100)
100    FORMAT(/1X,'ENTER CARBON CONTENT IN FEED IN % BY WEIGHT')
READ(*,*)WPC

WRITE(*,101)
101    FORMAT(/1X,'ENTER HYDROGEN CONTENT IN FEED IN % BY WEIGHT')
READ(*,*)WPH

WRITE(*,102)
102    FORMAT(/1X,'ENTER OXYGEN CONTENT IN FEED IN % BY WEIGHT')
READ(*,*)WPO

```

```

WRITE(*,103)
103  FORMAT(/1X,'ENTER NITROGEN CONTENT IN FEED IN % BY WEIGHT')
      READ(*,*)WPN

      CMOL=1.0
      HMOL=12.0*WPH/WPC/1.0
      OMOL=12.0*WPO/WPC/16.0
      NMOL=12.0*WPN/WPC/14.0

C    HIGHER HEATING VALUE (KJ/100KG)

      MWFD=12.0*CMOL+1.0*HMOL+16.0*OMOL+14.0*NMOL
      HHVFD=(34.91*12.0*CMOL+117.83*HMOL-10.34*16.0*OMOL)/MWFD

C    CHAR YIELD IS ASSUMED TO BE AS SAME AS % OF FIXED CARBON IN
FEED

      WRITE(*,105)
105  FORMAT(/1X,'ENTER FIXED CARBON IN FEED IN % BY WEIGHT (BETWEEN
$10.0 TO 30.0)')
      READ(*,*)FC

C    DENSITY (G/CM3) OF 100 KG WOOD

      WRITE(*,110)
110  FORMAT(/1X,'ENTER BULK DENSITY OF FEED AT DRY CONDITION IN
$G/M3')
      READ(*,*)DNSTY

C    LOWER HEATING VALUE OF FEED IS

      LHV=19.55*1.0E5-2440.*18.0*HMOL*4.255/2.0

C    WOOD CHIP SIZE IN CM

      WRITE(*,115)
115  FORMAT(/1X,'ENTER DIAMETER OF WOOD PARTICLE IN CM')
      READ(*,*)DP

      RP=DP/2.0

C    THROAT DIAMTER IN CM

      WRITE(*,120)
120  FORMAT(/1X,'ENTER THROAT DIAMETER IN CM')
      READ(*,*)DTHRT

```

```

C      THROAT ANGLE

      WRITE(*,125)
125   FORMAT(/1X,'ENTER THROAT ANGLE USUALLY 30.0 TO 90.0 DEG')
      READ(*,*)ANGLTHRT

C      THICKNESS (IN CM AND BETWEEN 5 TO 15) OF THE INSULATING
C      MATERIAL

      WRITE(*,130)
130   FORMAT(/1X,'ENTER INSULATION THICKNESS IN CM, IF NOT INSULATED,
$ENTER 0.0')
      READ(*,*)THICKI
      IF (THICKI.EQ.0.0) GOTO 136

C      THERMAL CONDUCTIVITY (IN W/CM K AND BETWEEN 0.0003 TO 0.001)

      WRITE(*,135)
135   FORMAT(/1X,'ENTER THERMAL CONDUCTIVITY OF INSULATION
$IN W/CM K, IF NOT INSULATED, 0.0')
      READ(*,*)TCINSU
      GOTO 139
136   TCINSU=0.0

C      MOISTURE CONTENT OF RUBBER WOOD (DRY BASIS BY %)

139   WRITE(*,140)
140   FORMAT(/1X,'ENTER MOISTURE CONTENT (BY DB) OF FEED IN
$PERCENTAGE')
      READ(*,*)MC

C      TOTAL INTEGRAL HEAT OF WETTING OF WOOD IS RANGING FROM
C      18.9 TO 20.5 CAL/G

      HWET=20.0*4.184*MC

C      TOTAL KG OF WATER INPUT

      TWAT=MC
      WATMOL=TWAT/18.0

C      HEAT OF VAPOURIZATION OF WATER

      HEVAP=2440.0*MC

```

```

C      TOTAL HEAT OF VAPOURIZATION

          THEVAP=HEVAP+HWET

C      TEMPERATURE OF INPUT AIR (K)

WRITE(*,145)
145    FORMAT(/1X,'ENTER TEMPERATURE OF INPUT AIR IN K')
      READ(*,*)TAIR

C      FUEL WOOD INPUT IN KG/HOUR

WRITE(*,150)
150    FORMAT(/1X,'ENTER FEED INPUT IN KG/HR')
      READ(*,*)FXFUEL

C      AIR INPUT KG PER HOUR

WRITE(*,155)
155    FORMAT(/1X,'ENTER AIR INPUT IN KG/HR')
      READ(*,*)AIRIN

          AFR=AIRIN/FXFUEL

C      HEAT LOSS FROM THE GASIFIER

WRITE(*,160)
160    FORMAT(/1X,'ENTER HEAT LOSS IN %, GENERALLY BETWEEN 0.0 TO
$20.0')
      READ(*,*)HL

OPEN(UNIT=8,FILE='COMPSOFTWARE.OUT')

C      THE AMOUNT OF CH4 ASSUMED

          xCH4=0.035

C      THIS SECTION CALCULATES THE COMPOSITION AND THE TEMPERATURE OF
C      THE GAS LEAVING THE PYROLYSIS ZONE. INITIAL GUESSES FOR NUMBER
C      OF MOLES OF GAS COMPONENTS AND TEMPERATURE ARE ENTERED HERE

          X1(1)=0.5
          X1(2)=0.5
          X1(3)=0.5
          X1(4)=0.5
          X1(5)=1.2

```

```

X1(6)=1300.0

CALL NEWTS(X1,N1,check)
CALL FUNCVS(N1,X1,F1)
IF (check) THEN
WRITE(*,*) 'Convergence problems.'
ENDIF

C THE UNITS OF X1(1) TO X1(5) ARE IN KMOL/M2 HR. THEY ARE
C CONVERTED TO MOL/CM2 SEC BY MULTIPLYING 2.7778E-5. AND THEN
C EXPRESSED AS FLUXES

CF=2.7778E-5
FLX=(100.0/MWFD)*(FXFUEL/100.0)*CF

FX(1)=X1(2)*FLX
FX(2)=X1(1)*FLX
FX(3)=X1(3)*FLX
FX(4)=X1(4)*FLX
FX(5)=XCH4*FLX
FX(6)=X1(5)*FLX

FXCMOL=CHAR*FLX

TB=X1(6)

FUR=FXFUEL/100.0

Z=FXCMOL*3.6E4

EMSVTY=0.6
SIGMA=5.669E-12
FXFUE=FXFUEL/3.6E4

TT=0.0
NN=1
TL=0.0

DO 490 J=1,21
VOL1=RP**3
R3(J)=VOL1*REAL(21-J)/20.0
R(J)=R3(J)**(1.0/3.0)
R2(J)=R(J)*R(J)

490 CONTINUE

```

```

GC=82.05
WM(1)=28.0
WM(2)=2.0
WM(3)=44.0
WM(4)=18.0
WM(5)=16.0
WM(6)=28.0

C THE APPARENT CONCENTRATION OF CHAR IS OBTAINED DIVIDING THE
C SUPERFICIAL CONCENTRATION BY(1-VOID FRACTION OF PARTICLE)

EPI=0.8
DCHAR=DNSTY*Z*12.0/FXFUEL
CCH=DCHAR/12.0

DO 500 J=1,21
CCI(J)=CCH
500 CONTINUE
AVGCCH=CCH

WRITE(8,502)
502 FORMAT (1X,'DRY GAS COMP. BY % & TEMP. ALONG GASIFIER AXIS
$ARE AS FOLLOWS:'/)

WRITE(8,504)
504 FORMAT(1X,'TIME',3X,'LOC.CM',4X'CO',4X,'H2',3X,'CO2',3X,
$ 'CH4',4X,'N2',4X,'EFF',1X,'TEMP' )

DO 900 KK=0,200

C REACTOR DIAMETER IN CM
DREA=DTHRT+2.0*TAN(ANGLTHRT/2.0/57.3)*TL
RREA=DREA/2.0

DT=3.0*REAL(KK)**0.8
TT=DT+TT
TMFXG=0.0

DO 515 I1=1,6
TMFXG=TMFXG+FX(I1)

515 CONTINUE
P=1.0
VG=GC*TB*TMFXG/P

```

```

C      BED VOID FRACTION

      VOIDB=0.5-0.2*REAL(NN-1)/20.0
      V1=VG*VOIDB
      CCO2B=FX(3)/V1
      CH2OB=FX(4)/V1

C      THE INITIAL GUESS FOR SOLVING THE PARTICLE CONCENTRATION AND
C      TEMPERATURE PROFILES ARE AS FOLLOWS

      X2(1)=0.9*TB
      X2(2)=0.9*CCO2B
      X2(3)=0.7*CH2OB

C      THE MASS VELOCITY OF GAS PHASE

      GM=0.0

      DO 520 I2=1,6
      GM=GM+FX(I2)*WM(I2)

520      CONTINUE

      DNSTYG=GM/VG

C      ESTIMATING VISCOSITY OF GAS MIXTURE

      TB2=TB*TB
      TB3=TB2*TB
      POW=1.0E-6

      VS(1)=(78.9+0.389*TB-7.50E-5*TB2)*POW
      VS(2)=(28.3+0.220*TB-5.76E-5*TB2+1.01E-8*TB3)*POW
      VS(3)=(34.6+0.471*TB-1.30E-4*TB2+2.00E-8*TB3)*POW
      VS(4)=(-169.0+0.774*TB-2.31E-4*TB2+3.33E-8*TB3)*POW
      VS(5)=(186.+0.211*(TB-573.)-3.77E-5*(TB-573.)**2)*POW
      VS(6)=(78.2+0.422*TB-1.18E-4*TB2+1.82E-8*TB3)*POW

      DO 540 I3=1,6
      DO 530 J3=1,6
      IF(I3.EQ.J3)GO TO 530
      VSR=VS(I3)/VS(J3)
      WMR=WM(J3)/WM(I3)
      WMR1=1.0/WMR

      EPS(I3,J3)=(1.0+VSR**0.5*WMR**0.25)**2/2.8284/(1.0+WMR1)**0.5

```

```

530      CONTINUE
540      CONTINUE

      DO 550 I4=1,6
      FRN(I4)=FX(I4)/TMFXG

550      CONTINUE

      DO 570 I5=1,6
      G(I5)=0.0

      DO 560 J5=1,6
      IF(I5.EQ.J5)GOTO 560
      G(I5)=G(I5)+FRN(J5)*EPS(I5,J5)

560      CONTINUE
570      CONTINUE

      VSCO=0.0

      DO 580 I6=1,6
      IF(FRN(I6).EQ.0.0) GOTO 580
      VSCO=VSCO+VS(I6)/(1.0+G(I6)/FRN(I6))

580      CONTINUE

      RE=2.0*R(NN)*GM/VSCO

C      DIFFUSIVITIES OF CO2 AND H2O IN THE GAS MIXTURE

      TBR1=(TB/1200.0)**1.75
      TBR2=(TB/273.0)**1.75
      TBR3=(TB/298.0)**1.75
      TBR4=(TB/307.6)**1.75

      D(1)=1.48*TBR1
      D(2)=0.55*TBR2
      D(4)=0.198*TBR4
      D(5)=0.153*TBR2
      D(6)=0.167*TBR3
      SUM=0.0

      DO 590 I7=1,6
      IF(I7.EQ.3) GOTO 590
      SUM=SUM+FRN(I7)/D(I7)

```

```

590      CONTINUE

      DCO2M=(1.0-FRN(3))/SUM

      D(1)=2.69*TBR1
      D(2)=1.02*TBR4
      D(3)=D(4)
      D(5)=0.292*TBR4
      D(6)=0.256*TBR4

      SUM=0.0

      DO 600 I8=1,6
      IF(I8.EQ.4) GOTO 600
      SUM=SUM+FRN(I8)/D(I8)

600      CONTINUE

      DH2OM=(1.0-FRN(4))/SUM

C      CALCULATING MASS TRANSFER COEFFICIENTS

C      SC (SCHMIDT NO.):VSCO/(DNSTY*DM)
C      SH (SHERWOOD NO.): (KM*DIA)/DM

      VSC01=VSCO/DNSTYG
      SC1=VSC01/DCO2M
      RE1=RE**0.585*1.17
      SC1=SC1**0.3333
      KMCO2=RE1*SC1*DCO2M/(2.0*R(NN))
      SC1=VSC01/DH2OM
      SC1=SC1**0.3333
      KMH2O=RE1*SC1*DH2OM/(2.0*R(NN))

C      CALCULATING THERMAL CONDUCTIVITY OF THE GAS MIXTURE

      POW=1.0E-5
      TC(1)=(14.0+0.0526*TB-2.49E-7*TB2)*POW
      TC(2)=(-212.0+1.21*TB-7.64E-4*TB2+2.18E-7*TB3)*POW
      TC(3)=(-134.0+0.46*TB-3.63E-4*TB2+1.07E-7*TB3)*POW
      TC(4)=(-89.6+0.248*TB-9.25E-5*TB2+3.65E-8*TB3)*POW
      TC(5)=1.005E-3
      TC(6)=(20.6+0.0372*TB+6.19E-6*TB2+8.0E-10*TB3)*POW

      YKM=0.0

```

YM=0.0

DO 610 I9=1,6  
WM1(I9)=WM(I9)\*\*0.3333  
YKM=YKM+FRN(I9)\*TC(I9)\*WM1(I9)  
YM=YM+FRN(I9)\*WM1(I9)

610 CONTINUE

TCM=YKM/YM

C ESTIMATING SPECIFIC HEAT OF THE GAS MIXTURE

CP(1)=26.02+8.98E-3\*TB-1.914E-6\*TB2  
CP(2)=27.91+1.882E-3\*TB+6.3E-7\*TB2  
CP(3)=32.38+2.82E-2\*TB-7.08E-6\*TB2  
CP(4)=27.91+1.554E-2\*TB-1.983E-6\*TB2  
CP(5)=(4.03+1.648E-2\*TB-3.57E-6\*TB2)\*4.184  
CP(6)=26.02+8.3E-3\*TB-1.641E-6\*TB2

CPM=0.0  
AWM=0.0  
DO 620 I10=1,6  
CPM=CPM+FRN(I10)\*CP(I10)  
AWM=AWM+FRN(I10)\*WM(I10)

620 CONTINUE

CPG=CPM/AWM  
PR=CPG\*VSCO/TCM  
RE1=RE\*\*0.359  
HJ=0.357/RE1/VOIDB  
HCOEFF=CPG\*GM\*HJ/PR\*\*0.66667

C ESTIMATING EFFECTIVE THERMAL CONDUCTIVITY OF CHAR IN W/CM K

TCCHAR=5.4E-4  
  
EP=(1.0-EPI)\*AVGCCH/CCH  
EP=1.0-EP  
FCTOR=7.692\*EP-5.615  
IF(EP.LT.0.839)DECO2=DICO2M\*FCTOR  
  
IF(EP.LT.0.839)DEH2O=DH2OM\*FCTOR  
IF(EP.GT.0.839)DECO2=EP\*DICO2M  
IF(EP.GT.0.839)DEH2O=EP\*DH2OM

```

        FCTOR1=1.0+(TCCHAR/TCM-1.0)/3.0
        FCTOR1=1.0/FCTOR1
        FCTOR1=(1.0-EP)*FCTOR1
        KEFF=(EP*TCM+FCTOR1*TCCHAR)/(EP+FCTOR1)

C      HEAT LOSS IN THE TIME INTERVAL IS CALCULATED AS FOLLOWS

        IF(TCINSU.NE.0.0) GOTO 621
        TWOUT=1050.0
        TAMB=300.0

C      AT THE AVERAGE FILM TEMPERATURE OF 675K THERMAL CONDUCTIVITY OF
C      AIR IS 0.00051 W/CM K, KINETIC VISCOSITY IS 0.624 CM2/SEC AND
C      PRANDTL NUMBER IS 0.683

C      AT A FILM TEMPERATURE OF 375K THERMAL CONDUCTIVITY OF AIR IS
C      0.000318 W/CM K, KINETIC VISCOSITY IS 0.233 CM2/SEC, AND THE
C      PRANDTL NUMBER IS 0.697

        TCAIR=0.00051
        VISK=0.624
        PRAIR=0.683
        GOTO 622
621      TWOUT=450.0
        TAMB=300.0
        TCAIR=0.000318
        VISK=0.233
        PRAIR=0.697
622      TF=(TWOUT+TAMB)/2.0
        CVEXP=1.0/TF
        GRAV=980.0

C      GRASHOF NUMBER BASED ON DIAMTER OF THE REACTOR, GRD
C      RAYLEIGH NUMBER, RA=GRD*PR

        GRD=GRAV*CVEXP*(TWOUT-TAMB)*DREA**3/VIK/VISK
        RA=GRD*PRAIR
        IF(RA.GE.1.0E9) GOTO 623
        C1=0.59
        CPOW2=0.25
        GOTO 624

623      C1=0.021
        CPOW2=0.4

C      NUSSELT NUMBER, CNU

```

```

624      CNU=C1*RA**CPOW2
          HTFREE=CNU*TCAIR/DREA

C      WALL HEAT TRANSFER COEFFICIENT
          HW=3.6*(RE/VOIDB)**0.365*TCM/DP

C      STAINLESS STEEL IS USED TO CONSTRUCT THE GASIFIER, 0.75 INCH
C      I.E. 2.0CM THICK, THERMAL CONDUCTIVITY OF STEEL IS 0.5 W/CM K

          THRW=2.0
          TCSTL=0.5
          TDREA=DREA+2.0*THRW
          TRREA=TDREA/2.0

HTCIN=1.0/(1.0/HW+RREA*ALOG(TRREA/RREA)/TCSTL+RREA/TRREA/HTFREE)
          IF(TCINSU.NE.0.0)GOTO 625
          HTCIN=HTCIN
          CIRREA=TDREA*PI

          GOTO 626

625      TDREA1=TDREA+2.0*THICKI
          TRREA1=TDREA1/2.0
          CIRREA=TDREA1*PI
          HTCIN=1.0/(1.0/HTCIN+RREA*ALOG(TRREA1/TRREA)/TCINSU)

626      CONTINUE

          A1=(R(NN)*HCOEFF-KEFF)/(R2(NN)*HCOEFF)
          A2=(R(NN)*KMCO2-DECO2)/(R2(NN)*KMCO2)
          A3=(R(NN)*KMH2O-DEH2O)/(R2(NN)*KMH2O)

C      CALCULATINGTHE CONCENTRATION AND TEMPERATURE PROFILES IN THE
C      PARTICLE USING PARTICLE MODEL

          DO 650 J=NN,21

                  CALL NEWT(X2,N2,check)
                  CALL FUNCV(N2,X2,F2)
                  IF (check) THEN
WRITE(8,*) 'Convergence problems.'
                  ENDIF
!      WRITE(*,'(1X,a5,t10,a1,t22,a1)') 'index','X2','F2'
!      DO 13 i=1,N
!      WRITE(*,'(1X,i2,2X,2f12.6)') i,X2(i),F2(i)
!13      CONTINUE

```

```

T(J)=X2(1)
CCO2(J)=X2(2)
CH2O(J)=X2(3)
CCI(J)=CCI(J)*EXP(-B30)
RR=R(J)/RP

```

650 CONTINUE

C CALCULATING FLUXES FLOWING INTO THE PARTICLE AND OUT OF  
C PARTICLE AND CONVERTED INTO FLUXES BASED ON REACTOR CROSS  
C SECTION AREA

```

FXCO2P=KMCO2*(CCO2B-CCO2(NN))
FXH2OP=KMH2O*(CH2OB-CH2O(NN))
FXCOP=FXH2OP+2.0*FXCO2P
FXH2P=FXH2OP

VOLP=R3(NN)*4.0*PI/3.0
FXPTCL=FXCMOL/(AVGCCH)/VOLP
PN=DT*FXPTCL
AP=4.0*R2(NN)*PI
APPN=AP*PN
FX1(1)=FX(1)+FXCOP*APPN
FX1(2)=FX(2)+FXH2P*APPN
FX1(3)=FX(3)-FXCO2P*APPN
FX1(4)=FX(4)-FXH2OP*APPN
FX1(5)=FX(5)
FX1(6)=FX(6)
FXCML1=FXCMOL-(FXCO2P+FXH2OP)*APPN

```

C THE LATENT HEAT OF GASES AND SOLID ARE RECORDED FOR CALCULATING  
C TEMPERATURE AT THE END PERIOD OF DT

```

CALL HLAT(TB,H)
SUM=0.0
DO 680 I11=1,6
SUM=SUM+FX(I11)*H(I11)

```

680 CONTINUE

C SPECIFIC HEAT OF CHARCOAL=12.15 J/GMOL K

```

SPCH=12.15
HCH=SPCH*(T(NN)-298.0)
FXHCH=FXCMOL*HCH

```

```

SUM=SUM+FXHCH

FX(1)=FX1(1)
FX(2)=FX1(2)
FX(3)=FX1(3)
FX(4)=FX1(4)
FXCMOL=(FXCMOL+FXCML1)/2.0
TP=T(NN)
VP=FXCMOL/AVGCCH/(1.0-VOIDB)
DTL=VP*DT
DWA=CIRREA*DTL
DTEMP=TB-TAMB
CROSSA=PI*RREA**2
HFXW=HTCIN*DTEMP
TW1=TAMB+HFXW/HTFREE
HLCC=HFXW*DWA/CROSSA/1000.0
HLORAD=SIGMA*EMSVTY*DWA*(TB**4-TAMB**4)/1000.0/CROSSA
HLOSS=HLORAD+HLCC+DREA**2*0.071*DTEMP/92**2/1000.0

CONTINUE

      X3(1)=TB*0.95

      CALL NEWTT(X3,N3,check)
      CALL FUNCVT(N3,X3,F3)
      IF (check) THEN
        WRITE(8,*) 'Convergence problems.'
      ENDIF
!      WRITE(*,'(1X,a5,t10,a1,t22,a1)') 'index','X3','F3'
!      DO 13 i=1,N
!      WRITE(*,'(1X,i2,2X,2f12.6)') i,X3(i),F3(i)
!13      CONTINUE

      TB=X3(1)
      IF(TB.LT.400.0) GOTO 910
      SPEEL=0.0

      DO 690 J=NN,21
        CR=CCI(J)/CCH
        IF(CR.GT.0.03) GOTO 700
        SPEEL=SPEEL+CCI(J)*VOL1*PI*4.0/60.0*FXPTCL

690      CONTINUE

700      X1(1)=T(J)

```

```
X1(2)=CCO2(NN)
X1(3)=CH2O(NN)
NN=J
```

```
C THE FLUX OF LOW HEATING VALUE OF THE PRODUCT GAS
```

```
HVCO=2.83E5
HVV2=2.41E5
HVCH4=8.023E5
```

```
HFLUX=HVCO*FX(1)+HVV2*FX(2)+HVCH4*FX(5)
```

```
HVFUFX=LHV*FXFUE/100.0
EFF=HFLUX/HVFUFX
```

```
EFF=100*EFF
TL=TL+DTL
IF(TL.GT.50) GOTO 1000
```

```
TFX=TMFXG-FX(4)
```

```
CO =FX(1)/TFX
H2 =FX(2)/TFX
CO2=FX(3)/TFX
CH4=FX(5)/TFX
TN2=FX(6)/TFX
```

```
710 WRITE(8,710)TT,TL,CO,H2,CO2,CH4,TN2,EFF,TB
FORMAT(F6.0,F6.1,4X,5(F5.3,1X),F5.1,1X,F5.0)
```

```
FXCMOL=FXCML1-SPEEL
IF(NN.EQ.21) GOTO 1000
```

```
C CALCULATING THE AVERAGE CHAR CONCENTRATION
```

```
SUM=0.0
```

```
DO 890 JJ=NN,21
SUM=SUM+CCI(JJ)*2.0
```

```
890 CONTINUE
```

```
SUM=SUM-CCI(NN)-CCI(21)
AVGCCH=SUM/2.0/REAL(21-NN)
```

```
900 CONTINUE
```

```

910      WRITE(8,920)
920      FORMAT(///1X,'COMPLETE GASIFICATION IS NOT FEASIBLE'///)
1000     STOP
        END

```

```

C      *****SUBROUTINE FUNCVS*****
C      SUBROUTINE FUNCVS REPRESENTS THE PYROLYSIS EQUATIONS TO
C      CALCULATE THE COMPOSITION AND TEMPERATURE OF GAS LEAVING THE
C      ZONE

```

```

SUBROUTINE FUNCVS(N1,X1,F1)
IMPLICIT NONE
INTEGER N1
REAL X1(N1),F1(N1),MC,K,WATMOL,ENGYIN,ENGYOUT,CHAR,TARS,AIR,FC,
$     HL,MWFD,HHVFD,HVCHAR,HVTAR,HVGAS,SHCHAR,TP,DP,SHTAR,
$     SHGAS,xCH4,CMOL,HMOL,OMOL,Q,MWCHAR,MWTAR,TDCOM,AH2OML,
$     TAIR,AFR,H(6)

```

```

COMMON/BK1/ TAIR,CMOL,HMOL,OMOL
$     /BK2/ xCH4,CHAR,MWFD,MC,DP,HHVFD,AFR,HL,FC

```

```

K=10**-(36.7251-(3994.7/X1(6))+(4.462E-3*X1(6))-
c      (6.72E-7*X1(6)*X1(6))-12.22*ALOG10(X1(6)))

```

```

TDCOM=160.0*DP**0.3
TP=X1(6)-TDCOM

```

```

MWTAR=12.0*1.0+1.0*1.03+16.0*0.03
MWCHAR=12.0

```

```

HVTAR=0.3491*88.8+1.1783*7.6-0.1034*3.6
HVCHAR=34.39

```

```

AIR=AFR*MWFD/29.0
AH2OML=0.0686*AIR*29.0/18.0
CHAR=(FC/100.0)*MWFD/MWCHAR
TARS=0.01*MWFD/MWTAR
WATMOL=(MC/100.0)*MWFD/18.0

```

```

F1(1)=X1(1)*X1(3)-K*X1(2)*X1(4)
F1(2)=X1(2)+X1(3)+xCH4+CHAR+TARS-CMOL
F1(3)=2.0*X1(1)+2.0*X1(4)+4.0*xCH4+1.03*TARS-
      2.0*(WATMOL+AH2OML)-HMOL
F1(4)=X1(2)+2.0*X1(3)+X1(4)-2.0*0.21*AIR-WATMOL-AH2OML

```

```

      -OMOL+0.03*TARS
F1(5)=X1(5)-0.79*AIR

ENGYIN=HHVFD*MWFD*1000.0

HVCHAR=HVCHAR*1000.0*MWCHAR*CHAR
HVTAR=HVTAR*1000.0*MWTAR*TARS
HVGAS=2.41E5*X1(1)+2.83E5*X1(2)+8.023E5*xCH4
SHCHAR=12.15*CHAR*(TP-TAIR)
SHTAR=1.568*TARS*MWTAR*(X1(6)-TAIR)

CALL HLAT(X1(6),H)

SHGAS=X1(1)*H(2)+X1(2)*H(1)+X1(3)*H(3)+X1(4)*H(4)+xCH4*H(5)+
C X1(5)*H(6)
Q=(HL/100.0)*ENGYIN

ENGYOUT=HVCHAR+HVTAR+HVGAS+SHCHAR+SHTAR+SHGAS+Q

F1(6)=ENGYIN-ENGYOUT

RETURN
END

C *****SUBROUTINE FUNCV*****
C SUBROUTINE FUNCV CONTAINS THE MASS AND ENERGY BALANCE
C EQUATIONS. THESE NON LINEAR COUPLED SIMULTANEOUS EQUATIONS ARE
C THE RESULTS OF FREDHOLM INTEGRAL CONVERSION OF THE DIFFERENTIAL
C MASS AND ENERGY EQUATIONS. THIS SUBROUTINE IS USED TO CALCULATE
C THE PROFILE CHAR PARTICLE IN THE GASIFICATION ZONE.

SUBROUTINE FUNCV(N2,X2,F2)
  IMPLICIT NONE
  INTEGER N2,J,NN
  REAL X2(3),F2(3),KEFF,KMCO2,KMH2O,DECO2,CCI(21),R(21),
$ R2(21),DT,R3(21),DLTH1,DLTH2,BB1,B1,BB2,B2,B3,B4,BB5,
$ B5,B6,RP2,RP,RP3,B7,B22,B8,TB,B10,B11,A1,A2,B12,CCO2B,
$ B20,B21,A3,CH2OB,B30,DEH2O

  COMMON/BK3/A1,A2,A3,R,R2,R3,CCI,CCO2B,CH2OB,RP,RP2,RP3,KEFF,
$ DECO2,DEH2O,DT,B30,KMCO2,KMH2O,NN,J
$ /BK4/TB

  DLTH1=1.724E5
  DLTH2=2.319E5
  IF(X2(2).LT.0.0)X2(2)=-X2(2)

```

```

IF(X2(3).LT.0.0)X2(3)=-X2(3)
BB1=X2(2)**0.7
B1=BB1*CCI(J)
BB2=X2(3)**0.7
B2=BB2*CCI(J)
B3=(B1*DLTH1+B2*DLTH2)/KEFF
B4=3.561E5
IF(ABS(X2(1)).LT.550.0)X2(1)=550.0
IF(ABS(X2(1)).GT.1700.0)X2(1)=650.0
BB5=EXP(-13065.0/X2(1))
B5=B4*BB5*B4*BB5
B6=B3*B5
R2=R*R
R3=R2*R
RP2=RP*RP
RP3=RP2*RP

B7=(R2(J)-R3(J)*A1)/3.0
B8=(R2(NN)-R2(J))/2.0-(R3(NN)-R3(J))*A1/3.0

F2(1)=TB-X2(1)-(B7+B8)*B6

B10=B1/DECO2
B11=(R2(J)-R3(J)*A2)/3.0
B12=(R2(NN)-R2(J))/2.0-(R3(NN)-R3(J))*A2/3.0

F2(2)=CCO2B-X2(2)-(B11+B12)*B5*B10

B20=B2/DEH2O
B21=(R2(J)-R3(J)*A3)/3.0
B22=(R2(NN)-R2(J))/2.0-(R3(NN)-R3(J))*A3/3.0

F2(3)=CH2OB-X2(3)-(B21+B22)*B5*B20

B30=B5*(BB1+BB2)*DT

```

```

RETURN
END

```

```

C *****SUBROUTINE FUNCVT*****
C SUBROUTINE FUNCVT REPRESENTS THE ENERGY BALANCE TO CALCULATE
C THE GAS BULK TEMPERATURE

```

```

SUBROUTINE FUNCVT(N3,X3,F3)

```

```

      IMPLICIT NONE
      INTEGER N3, I
      REAL F3(1), X3(1), H(6), FX(6), TB, SUM, TP, APPN, HCOEFF, HLOSS, HCH,
$      FXCML1, SUM1

      COMMON/BK4/ TB
$      /BK5/ FX
$      /BK6/ SUM, TP, APPN, HCOEFF, HLOSS, HCH, FXCML1

      CALL HLAT(X3(1), H)
      SUM1=0.0
      DO 5 I=1, 6
          SUM1=SUM1+H(I)*FX(I)
5      CONTINUE
      SUM1=SUM1+FXCML1*HCH

      F3=SUM-HCOEFF*(TB-TP)*APPN-SUM1-HLOSS

      RETURN
      END

C      *****SUBROUTINE HLAT*****
C      SUBROUTINE HLAT CONTAINS OF THE CORRELATED EXPRESSION OF THE
C      ENTHALPIES OF THE GASES FROM JANAF THERMODYNAMIC DATA FOR
C      TEMPERATURE RANGING FROM 200K TO 2500K

      SUBROUTINE HLAT(T, H)
      REAL H(6), T2, T3

          T2=T*T
          T3=T2*T

          H(1)= -8162.0+26.02*T+4.49E-3*T2-6.38E-7*T3
          H(2)= -8372.0+27.91*T+9.41E-4*T2+2.10E-7*T3
          H(3)= -10856.0+32.38*T+1.41E-2*T2-2.36E-6*T3
          H(4)= -8998.0+27.91*T+7.77E-3*T2-6.61E-7*T3
          H(5)= (-1937.0+04.03*T+8.24E-3*T2-1.19E-6*T3)*4.184
          H(6)= -8127.0+26.02*T+4.15E-3*T2-5.47E-7*T3

      RETURN
      END

```

## APPENDIX F

### DETERMINATION OF MASS FLOW RATE OF THE GAS

The following equation is used to determine the mass flow rate of the product gas passing through an orifice plate (Australian Standard 1993).

$$Q_m = C\varepsilon_1 \frac{\pi}{4} d^2 (2\Delta P \rho_1)^{0.5} (1 - \beta^4)^{-0.5} \quad (\text{F.1})$$

Where;

- C = Coefficient of discharge
- d = Diameter of the orifice
- $Q_m$  = Mass flow rate
- $\beta$  = Diameter ratio (d/D)
- $\Delta P$  = Pressure difference
- $\varepsilon$  = Expansion factor
- $\rho$  = Density of the gas mixture

Subscript 1 is referred to the upstream condition

The Reynolds number is dependent on mass flow rate and the mass flow rate is dependent on the coefficient of discharge. Thus the final values of coefficient of discharge and mass flow rate have to be obtained by iteration.

Reynolds number can be written as follows:

$$Re_1 = \frac{4Q_m}{\pi\mu_1 D} \quad (\text{F.2})$$

Where;

- $Re_1$  = Reynolds number

$\mu$  = Dynamic viscosity of the gas mixture

Subscript  $I$  is referred to the upstream condition

Initially a value for  $C$  is assumed and the corresponding  $Re$  number is found for a given  $\beta$  from the Tables given in Australian Standard (1993). Next using equation F.2,  $Q_m$  is calculated and finally this  $Q_m$  value is substituted in equation F1 to calculate  $C$ . If the assumed and calculated  $C$  values are different then the second iteration is started with the calculated  $C$  value. This process is repeated until the  $C$  values are equal for a given precision.

### Determination of viscosity of the gas mixture

Wilke (1950) has proposed the following equations to calculate the viscosity of a gas mixture.

$$\eta_{mix} = \sum_{i=1}^n \frac{y_i \eta_i}{\sum_{j=1}^n y_j \phi_{ij}} \quad (F.3)$$

and

$$\phi_{ij} = \frac{[1 + (\eta_i / \eta_j)^{0.5} (M_j / M_i)^{0.25}]^2}{[8(1 + M_i / M_j)]^{0.5}} \quad (F.4)$$

Where;

$M_i$  = Molecular weight of  $i$ th component

$n$  = Number of components

$y_i$  = Mole fraction of  $i$ th component

$\phi_{ij}$  = Parameter

$\eta_{mix}$  = Viscosity of the mixture

$\eta_i$  = Viscosity of the  $i$ th component

## Determination of the density of the gas mixture

Ideal gas law is used to calculate the density of the gas mixture.

$$\rho_{mix} = \frac{PM}{RT} \quad (F.5)$$

Where;

$\rho_{mix}$  = Density of the mixture

$P$  = Pressure

$M$  = Molecular weight of the mixture

$R$  = Universal gas constant

$T$  = Temperature

Molecular weight of the gas mixture is total mass of each gas component divided by total number of moles. All the calculations have been carried out in the Excel Programme.

## References

Australian Standard. (1993), *Measurement of fluid flow in closed conduits*, AS 2360.1.1, Standards Association of Australia.

Wilke, C. R. (1950), 'Diffusional properties of multicomponent gases', *Chemical Engineering Progress*, vol. 46, pp. 95-104.

## APPENDIX G

### SUMMARY OF EXPERIMENTAL RESULTS

#### Gas composition

Test No.	Chip size (in)	m/c (w.b.)	Air/fuel ratio	CO	H <sub>2</sub>	CO <sub>2</sub>	CH <sub>4</sub>	N <sub>2</sub>	Gas flow rate (m <sup>3</sup> /hr)	Gas HHV (MJ/m <sup>3</sup> )	Eff (%)
1	1.5	18.5	2.03	19.6	17.2	9.9	1.4	51.9	52.37	4.950	60.5
2		16.0	2.20	20.2	18.3	9.7	1.1	50.7	57.73	5.059	68.2
3		14.7	2.37	19.4	17.2	9.7	1.1	52.6	59.68	4.828	67.3
4*		13.6	2.54	-	-	-	-	-	-	-	-
5	2.0	16.0	1.96	18.4	17.0	10.6	1.3	52.7	53.55	4.749	57.3
6		15.2	2.12	19.7	13.2	10.8	1.3	55.0	62.07	4.448	62.2
7		14.0	2.29	18.9	12.5	8.5	1.2	59.1	62.46	4.227	59.5
8*		13.1	2.45	-	-	-	-	-	-	-	-
9	2.5	14.7	1.86	19.1	15.5	11.4	1.1	52.9	51.94	4.579	51.6
10		13.8	2.04	22.1	12.7	10.5	1.3	53.4	62.54	4.672	63.3
11*		13.0	2.20	-	-	-	-	-	-	-	-
12		12.5	2.36	19.1	13.0	10.7	1.2	56.0	63.54	4.324	64.2
13	1.5	18.5	2.53	18.0	13.0	10.5	1.5	57.0	63.54	4.293	59.1

(\* = Gas analysis were not available)

### Temperature profile

Test No.	Temperature (K) along the axis from the throat					
	4cm	8cm	12cm	16cm	20cm	24cm
1	1237	1252	1174	1155	1002	762
2	1255	1276	1188	1170	1032	826
3	1211	1321	1263	1190	1043	848
4	1335	1351	1294	1212	1063	865
5	1286	1299	1259	1236	1039	796
6	1270	1295	1225	1195	1049	839
7	1349	1381	1299	1231	1060	846
8	1377	1410	1315	1241	1093	867
9	1378	1390	1352	1272	1104	973
10	1428	1444	1412	1366	1142	971
11	1454	1493	1460	1439	1202	971
12	1399	1467	1450	1420	1206	965

## APPENDIX H

### HEAT LOSS CALCULATION FOR ENERGY BALANCE

#### Heat Losses:

The main heat losses from the gasifier wall to the surrounding are through convection and radiation modes. The heat loss by convection ( $Q_c$ ) is calculated by the basic convection equation given below.

$$Q_c = h_c A(T - T_{air})$$

Where  $h_c$  and  $A$  are convective heat transfer and surface area respectively. Holman (2002) has given the following equations to calculate the convective heat transfer for different shapes.

For vertical cylinder,

$$h_c = 1.42 \left( \frac{T - T_{air}}{l_V} \right)^{0.25}$$

For horizontal plate with heated plate facing upwards,

$$h_c = 1.32 \left( \frac{T - T_{air}}{l_H} \right)^{0.25}$$

For horizontal plate with heated plate facing downwards,

$$h_c = 0.59 \left( \frac{T - T_{air}}{l_H} \right)^{0.25}$$

Where:

$l_H$  = Horizontal dimension (Area /Perimeter)

$l_V$  = Vertical dimension

$T$  = Average experimental temperature of the section

Radiation heat loss is calculated using the basic radiation equation given below.

$$Q_r = A\varepsilon\sigma(T^4 - T_{\text{sky}}^4)$$

Where  $\varepsilon$  and  $\sigma$  are emissivity and Stefan-Boltzman constant which is  $5.699 \times 10^{-8}$  W/m<sup>2</sup>K<sup>4</sup> respectively. The emissivity of iron is taken as 0.390 (Perry et al. 1997). Duffie and Beckman (1980) have given the following correlation for  $T_{\text{sky}}$ .

$$T_{\text{sky}} = 0.0552T_{\text{air}}$$

Therefore the total heat loss (Q) is:

$$Q = Q_c + Q_r$$

## References

- Duffie, J. A. & Beckman, W. A. (1980), *Solar Engineering of Thermal Process*, John Wiley and Sons, New York.
- Holman, J. P. (2002), *Heat Transfer*, Ninth Edition, McGraw Hill, NY.
- Perry, R. H., Green, D. W. & Maloney, J. O. (1997), *Perry's Chemical Engineers Handbook*, Sixth Edition, McGraw-Hill Book Company, New York.