This report investigates the use of wood to fuel a trimburner incorporated in a combined-cycle, biomass-fired power plant. The term "trimburner" applies to a ductburner which is designed to boost the temperature of the air stream entering the gas turbine. The work presented here is part of a larger biomass energy project at Oregon State University.

Wood conversion processes capable of producing a clean synthetic fuel were investigated since direct wood combustion products are too "dirty" to be allowed to pass through the turbine blading. Of the three wood conversion processes considered (pyrolysis, gasification, methanol production), gasification was selected as the most applicable process for the trimburner concept.

Three wood-fueled trimburner systems were then developed employing an up-draft gasifier design. These sub-systems, designated A, B, and C, differ in the way the
producer gas formed in the gasifier was compressed to the trimburner operating pressure. Sub-system A compresses the producer gas after the gasifier, thus simulating the operation of an atmospheric gasifier. Sub-systems B and C compress the inlet air entering the gasifier and, therefore, simulate a pressurized gasifier system. Sub-system B accomplishes the inlet air compression via a separate compressor whereas sub-system C bleeds some compressed air from the power plant's main compressor.

The simulation of the sub-systems allowed for comparison of the designs in terms of parameters such as efficiency and gas composition. In addition, the effects of changing system variables such as wood moisture content and gasifier air/fuel equivalence ratio on the performance of the sub-systems and the overall system were evaluated. It was determined that the most efficient operation of all the trimburner sub-systems occurred at the lowest allowable operating gasifier equivalence ratio. Also, increasing the wood moisture content decreased the efficiency of the sub-systems.

In general, sub-system C exhibited superior sub-system efficiency values by about 2 percentage points over the entire range of equivalence ratios and moisture contents. Also, results for the entire plant indicate that sub-system C again maintains superiority in efficiency by about .5 percentage points. However,
results from the simulation of the sub-systems indicate a basic difference in the producer gas formed from pressurized and non-pressurized gasifiers. The atmospheric gasifier of sub-system A produces a gas higher in combustible gases but lower in temperature as compared to that from the pressurized gasifiers of sub-systems B and C. This indicates that sub-system C (and B) will be much more sensitive to the heat exchanger process used to recover the sensible energy in the dirty producer gas stream than sub-system A.

Based on the apparent superiority of sub-system C, a simple economic analysis between this sub-system and a conventional oil-fired trimburner was conducted. Results of this analysis indicated that the oil-fired trimburner would yield a more attractive payback period than the wood-fueled sub-system C trimburner. The payback periods for the oil-fired and wood-fired trimburner systems were estimated at approximately one year and four years respectively.
Analysis of a Wood-Fueled Trimburner System for Use in a
Combined-Cycle, Wood-fired Power Plant

by

Jerry D. Stephenson

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To my wife Susan, and daughter Becky, for their love and understanding throughout this endeavor.
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## TABLE OF CONTENTS

<table>
<thead>
<tr>
<th>Chapter 1 -- Introduction</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>System Description</td>
<td>1</td>
</tr>
<tr>
<td>System Constraints</td>
<td>1</td>
</tr>
<tr>
<td>System Variations</td>
<td>5</td>
</tr>
<tr>
<td>System Comparison</td>
<td>7</td>
</tr>
<tr>
<td>System 4</td>
<td>8</td>
</tr>
<tr>
<td>System 7</td>
<td>8</td>
</tr>
<tr>
<td>System Comparison</td>
<td>11</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Chapter 2 -- Wood Conversion Processes</th>
<th>15</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pyrolysis</td>
<td>15</td>
</tr>
<tr>
<td>Types of Pyrolyzers</td>
<td>17</td>
</tr>
<tr>
<td>Advantages</td>
<td>25</td>
</tr>
<tr>
<td>Disadvantages</td>
<td>26</td>
</tr>
<tr>
<td>Gasification</td>
<td>27</td>
</tr>
<tr>
<td>Modes of Gasification</td>
<td>30</td>
</tr>
<tr>
<td>Types of Gasifiers</td>
<td>31</td>
</tr>
<tr>
<td>Advantages</td>
<td>38</td>
</tr>
<tr>
<td>Disadvantages</td>
<td>39</td>
</tr>
<tr>
<td>Methanol Production</td>
<td>40</td>
</tr>
<tr>
<td>Advantages</td>
<td>43</td>
</tr>
<tr>
<td>Disadvantages</td>
<td>44</td>
</tr>
<tr>
<td>Conclusion</td>
<td>44</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Chapter 3 -- Gasifier/Trimburner Sub-system Design</th>
<th>46</th>
</tr>
</thead>
<tbody>
<tr>
<td>Design Considerations</td>
<td>46</td>
</tr>
<tr>
<td>Producer Gas Cleaning and Conditioning</td>
<td>47</td>
</tr>
<tr>
<td>Sensible Energy Recovery</td>
<td>48</td>
</tr>
<tr>
<td>Producer Gas Compression</td>
<td>49</td>
</tr>
<tr>
<td>Sub-system Description</td>
<td>49</td>
</tr>
<tr>
<td>Sub-system A</td>
<td>50</td>
</tr>
<tr>
<td>Sub-system B</td>
<td>50</td>
</tr>
<tr>
<td>Sub-system C</td>
<td>52</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Chapter 4 -- Gasifier/Trimburner Sub-system Modelling</th>
<th>55</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sub-system Component Modelling</td>
<td>55</td>
</tr>
<tr>
<td>Gasifier</td>
<td>55</td>
</tr>
<tr>
<td>Heat Exchanger</td>
<td>69</td>
</tr>
<tr>
<td>Gas Cleaning Equipment</td>
<td>71</td>
</tr>
<tr>
<td>Compressor</td>
<td>71</td>
</tr>
<tr>
<td>Trimburner</td>
<td>75</td>
</tr>
<tr>
<td>Overall Sub-system Analysis</td>
<td>80</td>
</tr>
<tr>
<td>First Law Efficiency</td>
<td>80</td>
</tr>
<tr>
<td>Second Law Efficiency</td>
<td>83</td>
</tr>
<tr>
<td>Irreversibility</td>
<td>85</td>
</tr>
<tr>
<td>Accuracy of Sub-system Model</td>
<td>86</td>
</tr>
<tr>
<td>Chapter 5 -- Results of Sub-system Simulations</td>
<td>Page</td>
</tr>
<tr>
<td>-----------------------------------------------</td>
<td>------</td>
</tr>
<tr>
<td>Results</td>
<td>94</td>
</tr>
<tr>
<td>Pyrolysis Temperature</td>
<td>94</td>
</tr>
<tr>
<td>Producer Gas Composition</td>
<td>97</td>
</tr>
<tr>
<td>Gasifier Exit Temperature</td>
<td>100</td>
</tr>
<tr>
<td>Producer Gas Stream Energy</td>
<td>104</td>
</tr>
<tr>
<td>Sub-system Compressor Power</td>
<td>110</td>
</tr>
<tr>
<td>Overall Sub-system First Law Efficiency</td>
<td>113</td>
</tr>
<tr>
<td>Overall Sub-system Second Law Efficiency</td>
<td>117</td>
</tr>
<tr>
<td>Overall Sub-system Irreversibility</td>
<td>119</td>
</tr>
<tr>
<td>Overall Biomass Plant Net Efficiency</td>
<td>123</td>
</tr>
<tr>
<td>Summary of Results</td>
<td>131</td>
</tr>
<tr>
<td>Comparision with Oil-fired Trimburner Design</td>
<td>134</td>
</tr>
<tr>
<td>Conclusions</td>
<td>138</td>
</tr>
</tbody>
</table>

References ........................................ 143

APPENDICES

A. Description of Biomass Plant Variations ........................ 146
   System 2 ........................................ 146
   System 3 ........................................ 148
   System 5 ........................................ 148
   System 6 ........................................ 148
   System 8 ........................................ 151
   Systems 9, 10, and 11 ............................ 151

B. Newton-Raphson Method ........................................ 152
   Example 6-3 ...................................... 155
   Solution ......................................... 155

C. Sample Output From Integrated Power Plant
   Simulation and Fortran Source Codes for Each
   Sub-System Design .................................. 158
   Sample Output ..................................... 159
   Sub-System A Fortran Code ........................ 171
   Sub-System B Fortran Code ........................ 182
   Sub-System C Fortran Code ........................ 193
<table>
<thead>
<tr>
<th>Figure</th>
<th>Description</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.1</td>
<td>Schematic diagram of base case power plant</td>
<td>2</td>
</tr>
<tr>
<td>1.2</td>
<td>Schematic diagram of system 4 power plant</td>
<td>9</td>
</tr>
<tr>
<td>1.3</td>
<td>Schematic diagram of system 7 power plant</td>
<td>10</td>
</tr>
<tr>
<td>1.4</td>
<td>Gross and net efficiencies of all the systems, based on higher and lower heating values</td>
<td>12</td>
</tr>
<tr>
<td>1.5</td>
<td>Gross, auxiliary, and net powers of all the systems</td>
<td>12</td>
</tr>
<tr>
<td>2.1</td>
<td>Effect of temperature on yield of pyrolysis products from wood</td>
<td>18</td>
</tr>
<tr>
<td>2.2</td>
<td>Schematic diagram of kiln pyrolyzer</td>
<td>19</td>
</tr>
<tr>
<td>2.3</td>
<td>Schematic diagram of fixed-bed pyrolyzer</td>
<td>21</td>
</tr>
<tr>
<td>2.4</td>
<td>Schematic diagram of fluidized-bed pyrolyzer</td>
<td>24</td>
</tr>
<tr>
<td>2.5</td>
<td>Schematic diagram of an operating chamber for a typical up-draft gasifier</td>
<td>29</td>
</tr>
<tr>
<td>2.6</td>
<td>Schematic diagram of an operating chamber for a typical down-draft gasifier</td>
<td>34</td>
</tr>
<tr>
<td>2.7</td>
<td>Schematic diagram of a typical fluidized-bed gasifier with associated gas cleaning equipment</td>
<td>37</td>
</tr>
<tr>
<td>2.8</td>
<td>Flow diagram of a typical high-pressure methanol production plant</td>
<td>42</td>
</tr>
<tr>
<td>3.1</td>
<td>Schematic diagram of sub-system A gasifier/trimburner design</td>
<td>51</td>
</tr>
<tr>
<td>3.2</td>
<td>Schematic diagram of sub-system B gasifier/trimburner design</td>
<td>53</td>
</tr>
<tr>
<td>3.3</td>
<td>Schematic diagram of sub-system C gasifier/trimburner design</td>
<td>54</td>
</tr>
<tr>
<td>4.1</td>
<td>Schematic diagram of flow model used to simulate up-draft gasifier operation</td>
<td>58</td>
</tr>
<tr>
<td>Figure</td>
<td>Description</td>
<td>Page</td>
</tr>
<tr>
<td>--------</td>
<td>-----------------------------------------------------------------------------------------------------------</td>
<td>------</td>
</tr>
<tr>
<td>4.2</td>
<td>Temperature dependency of equilibrium constant for the water-gas reaction</td>
<td>66</td>
</tr>
<tr>
<td>4.3</td>
<td>Flow model used to simulate operation of trimburner</td>
<td>77</td>
</tr>
<tr>
<td>4.4</td>
<td>Control volumes specified in overall sub-system analysis</td>
<td>81</td>
</tr>
<tr>
<td>4.5</td>
<td>Comparison of gasifier pyrolysis temperature between Desrosiers model and sub-system A model over range of equivalence ratios</td>
<td>91</td>
</tr>
<tr>
<td>4.6</td>
<td>Comparison of dry producer gas composition between Desrosiers model and sub-system A model over range of equivalence ratios</td>
<td>92</td>
</tr>
<tr>
<td>5.1</td>
<td>Pyrolysis temperature for each sub-system as a function of equivalence ratio</td>
<td>96</td>
</tr>
<tr>
<td>5.2</td>
<td>Dry producer gas composition as a function of equivalence ratio for sub-system A</td>
<td>96</td>
</tr>
<tr>
<td>5.3</td>
<td>Dry producer gas composition as a function of equivalence ratio for sub-system B</td>
<td>98</td>
</tr>
<tr>
<td>5.4</td>
<td>Dry producer gas composition as a function of equivalence ratio for sub-system C</td>
<td>98</td>
</tr>
<tr>
<td>5.5</td>
<td>Temperature of producer gas at gasifier exit for sub-system A as a function of equivalence ratio at four levels of wood moisture content</td>
<td>102</td>
</tr>
<tr>
<td>5.6</td>
<td>Temperature of producer gas at gasifier exit for sub-system B as a function of equivalence ratio at four levels of wood moisture content</td>
<td>102</td>
</tr>
<tr>
<td>5.7</td>
<td>Energy content of producer gas for sub-system A as a function of equivalence ratio with a wood moisture content of .05</td>
<td>106</td>
</tr>
<tr>
<td>5.8</td>
<td>Energy content of producer gas for sub-system A as a function of equivalence ratio with a wood moisture content of .50</td>
<td>106</td>
</tr>
</tbody>
</table>
Figure | Page
--- | ---
5.9 Energy content of producer gas for sub-system B as a function of equivalence ratio with a wood moisture content of .05 | 107
5.10 Energy content of producer gas for sub-system B as a function of equivalence ratio with a wood moisture content of .50 | 107
5.11 Sub-system A compressor power requirements as a function of equivalence ratio at four levels of wood moisture content | 111
5.12 Sub-system B compressor power requirements as a function of equivalence ratio at four levels of wood moisture content | 111
5.13 First law Efficiency of all three sub-systems as a function of equivalence ratio at wood moisture contents of .05 and .30 | 114
5.14 First law Efficiency of all three sub-systems as a function of equivalence ratio at wood moisture contents of .15 and .50 | 114
5.15 Second law efficiency of the sub-systems as a function of equivalence ratio at wood moisture contents of .05 and .30 | 118
5.16 Second law efficiency of the sub-systems as a function of equivalence ratio at wood moisture contents of .15 and .50 | 118
5.17 Irreversibility of the sub-systems as a function of equivalence ratio at wood moisture contents of .05 and .30 | 120
5.18 Irreversibility of the sub-systems as a function of equivalence ratio at wood moisture contents of .15 and .50 | 120
5.19 Plant net efficiency of system 7 employing the sub-systems as a function of equiv. ratio at wood moisture contents of .05 and .30 | 124
5.20 Plant net efficiency of system 7 employing the sub-systems as a function of equiv. ratio at wood moisture contents of .15 and .50 | 124
<table>
<thead>
<tr>
<th>Figure</th>
<th>Description</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>5.21</td>
<td>Gas turbine mass flow rate as a function of equivalence ratio for sub-system A at four levels of wood moisture content</td>
<td>128</td>
</tr>
<tr>
<td>5.22</td>
<td>Gas turbine mass flow rate as a function of equivalence ratio for sub-system B at four levels of wood moisture content</td>
<td>128</td>
</tr>
<tr>
<td>5.23</td>
<td>Gas turbine mass flow rate as a function of equivalence ratio for sub-system C at four levels of wood moisture content</td>
<td>129</td>
</tr>
<tr>
<td>5.24</td>
<td>Gas turbine net power as a function of equivalence ratio for sub-system A at four levels of wood moisture content</td>
<td>129</td>
</tr>
<tr>
<td>5.25</td>
<td>Gas turbine net power as a function of equivalence ratio for sub-system B at four levels of wood moisture content</td>
<td>130</td>
</tr>
<tr>
<td>5.26</td>
<td>Gas turbine net power as a function of equivalence ratio for sub-system C at four levels of wood moisture content</td>
<td>130</td>
</tr>
<tr>
<td>5.27</td>
<td>Comparison of system 7 plant net efficiency using the oil-fired and wood-fired trimburner designs at various wood moisture contents</td>
<td>135</td>
</tr>
<tr>
<td>5.28</td>
<td>Simple payback period for the oil-fired and wood-fired trimburner designs as a function of oil cost for a .50 wood moisture content</td>
<td>135</td>
</tr>
<tr>
<td>A.1</td>
<td>Schematic diagram of system 2 power plant</td>
<td>147</td>
</tr>
<tr>
<td>A.2</td>
<td>Schematic diagram of system 3 power plant</td>
<td>149</td>
</tr>
<tr>
<td>A.3</td>
<td>Schematic diagram of system 5 power plant</td>
<td>150</td>
</tr>
<tr>
<td>Table</td>
<td>Title</td>
<td>Page</td>
</tr>
<tr>
<td>-------</td>
<td>----------------------------------------------------------------------</td>
<td>------</td>
</tr>
<tr>
<td>1.1</td>
<td>Equipment components in base case design</td>
<td>3</td>
</tr>
<tr>
<td>2.1</td>
<td>Important reactions in gasification process</td>
<td>28</td>
</tr>
<tr>
<td>4.1</td>
<td>Specific heat equations for gasifier analysis</td>
<td>62</td>
</tr>
<tr>
<td>4.2</td>
<td>Properties of gases flowing through sub-system compressor</td>
<td>75</td>
</tr>
<tr>
<td>4.3</td>
<td>Casifier performance data from ENFOR survey and sub-system gasifier model</td>
<td>88</td>
</tr>
<tr>
<td>5.1</td>
<td>Factors used in trimburner economic analysis</td>
<td>137</td>
</tr>
</tbody>
</table>
ANALYSIS OF A WOOD-FUELED TRIMBURNER SYSTEM FOR USE IN A COMBINED-CYCLE, WOOD-FIRED POWER PLANT

Chapter 1

INTRODUCTION

The overall objective of this study was to investigate an efficient means of using wood fuel to boost the inlet air temperature to the gas turbine in a combined-cycle, biomass-fired power plant. Before such a task could be undertaken, a solid understanding of the plant design was required. The remainder of this chapter is provided as a background on some previous design work done on the biomass plant which justifies the need for an investigation of such a system.

SYSTEM DESCRIPTION

The previously designed base case power plant is shown in figure 1.1. Table 1.1 identifies the various system components shown in figure 1.1. The power plant is designed to burn wood fuels, obtained from sources such as forest residues and industrial mill wastes, to yield 10 MW of net output power. The power plant is intended to be semi-portable so that as fuel supplies in the area diminish, the plant can be re-located to a new site where residue wood supplies are abundant. The total power
Figure 1.1 Schematic diagram of base case power plant.
### TABLE 1.1 EQUIPMENT COMPONENTS IN BASE-CASE DESIGN
(Letters are identified on figure 1.1)

- **Dryer** - A
- **Cyclones** - B1, B2 and B3
- **Fans** - C1, C2, C3, C4 and C5
- **Storage Bins** - D
- **Screw Feeders** - E
- **Combustor** - F
- **Heat Exchanger (steam/flue gas and air/flue gas)** - G
- **Gas Turbines** - H
- **Compressor** - I
- **Gearboxes** - J1 and J2
- **Generators** - K1 and K2
- **Waste Heat Boiler** - L
- **Steam Turbine** - M
- **Condenser** - N
- **Deaerator** - O
- **Pump** - P
- **Hammer Mill** - Q
- **Air Pollution Control Devices** - R
produced is the sum of the power yielded from an indirect-fired gas turbine cycle and a conventional Rankine steam cycle. This dual cycle arrangement is called a combined-cycle system.

Two important points to note about the base case system are the heat exchanger (G), and the fuel preparation equipment (components A, B1, B2, and Q). These components are common in most indirect-fired power plants using dried wood as the fuel source. The heat exchanger transfers thermal energy from the hot combustion gases to the air and steam streams which are eventually expanded through turbines to yield power. This form of indirect firing is necessary for wood-fueled systems due to the "dirty" nature of wood combustion gases. As expected, indirect firing is less efficient than direct firing and it presents some serious limitations as well.

The fuel preparation equipment shown in figure 1.1 basically consists of a wood dryer and a hammer mill. These components dry and reduce the size of the incoming wood so that it can be burned efficiently in the combustor. The increased efficiency of the combustor burning dry, crushed wood is somewhat negated since some of the power produced must be used to operate this equipment. It can be seen then, that the performance of the power plant will depend greatly on the degree of fuel preparation.
It should be noted that figure 1.1 portrays only half of the overall base case power plant. All of the equipment and streams mentioned above are duplicated in a twin system, each producing one half of the total output power for the plant. Due to the relatively small size of the power plant, there is only one waste heat boiler for both of the twin systems. This is done by merging the two systems at C15 of figure 1.1, passing through the boiler as a single stream, and then splitting the exiting steam into two streams for each heat exchanger.

SYSTEM CONSTRAINTS

For the base case power plant described in the previous section there exist several constraining factors which must be considered when attempting to estimate the plant's performance. These constraints require restrictions to be placed on the system operation in addition to pointing out the need for possibly altering the base case system design itself.

Perhaps the most significant constraint on the given power plant design is the maximum temperature that the heat exchanger can withstand. Higher combustion gas inlet temperatures to the heat exchanger would yield both a higher temperature air stream to the gas turbine and a higher degree of superheat in the steam going to the steam
turbine. This would result in more power from both cycles. However, metallic heat exchangers cannot withstand this high inlet flue gas temperature without sustaining damage. Therefore, the maximum temperature that the steam and air stream can obtain is defined by heat exchanger constraints and not by what thermodynamics might determine to be the most efficient operating point of the gas and steam cycles.

Another important constraint on the power plant is the inlet condition of the wood. Incoming wood chip size and moisture content has a direct effect on the maximum combustion temperature obtainable, and therefore, the plant performance. Very large, wet wood chips will decrease the combustion gas temperature due to the amount of energy required to evaporate the water contained in each wood particle. This energy is lost to the system and will be reflected by a decrease in the overall efficiency of the plant.

Closely associated with the wood condition is the auxiliary work required by the fuel preparation equipment. Combustor designs specify the allowable fuel size and moisture content which in turn determine the requirements on the fuel preparation equipment. Suspension burners, for example, require very small dry wood particles for adequate performance while pile burners can handle large wood chips with up to a 50% moisture content. It can be
seen that the suspension burner will require much more auxiliary power for the fuel preparation equipment than the pile burner. The most efficient burner, however, cannot be determined without performing an adequate thermodynamic analysis between the two burners for the same power plant. It may be found that, although the suspension burner does indeed require more auxiliary power, it may yield a higher combustion gas temperature and result in a lower fuel burning rate.

The examples of constraints mentioned above, although not complete, do indicate a need for a means of varying the base case model so that some form of comparison could be made to determine the ideal plant structure. For this reason, the base case system was altered to several different systems each which closely resemble the base case, but vary enough to allow for analyzing the effects of key variables. The predicted performances of the various systems were then obtained using a thermodynamic computer model of the plant which could be varied to account for the different plant configurations. A description of the system variations and their predicted performance will be discussed in the following section.

SYSTEM VARIATION

Several variations on the base case design have been identified in previous work by Brynjolfsson, Bauer, and
Fox [Brynjolfsson, 1983; Bauer, 1984; Fox, 1984]. Two of these system variations (system 4 and 7) are important to the work of this study and are described below. Descriptions of the remaining system variations can be found in Appendix A.

System 4

System 4, as shown in figure 1.2, differs from the base case design (system 2) in that a "trimburner" is placed in the air stream just before the gas turbine. Since the metallic heat exchanger has a temperature limit of about 1500 F, the exit air temperature can only obtain a temperature of about 1450 F. The turbine specified for this plant, however, provides optimum performance at its maximum rated turbine inlet temperature of 1612 F. The object of the trimburner then, is to burn a small amount of clean fuel (oil in this case) directly in the air stream just before the gas turbine to obtain the 1612 F inlet temperature.

System 7

System 7, as shown in figure 1.3, incorporates the use of a trimburner with wet wood combustion. This system is similar to system 4 with the exception that all of the fuel preparation equipment has been removed. The wood
Figure 1.2 Schematic diagram of system 4 power plant.
FLOW DIAGRAM
BIOMASS-POWERED COMBINED-CYCLE POWER GENERATION FACILITY
SYSTEM VII

Figure 1.3 Schematic diagram of system 7 power plant.
chips entering the combustor will be of varying size and have moisture contents as high as 60%. Because of this, a different combustor which can handle the wet wood fuel is used.

SYSTEM COMPARISON

Simulations to determine the performance characteristics of each system variation were conducted. Although the results from this analysis are extensive, only those results which pertain to this study will be discussed here.

One of the important results obtained from the previous work was that of the efficiency of each of the systems. The net plant efficiency of each system considered is shown in figure 1.4. A comparison based on net efficiency shows the systems employing the trimburner concept (systems 4 and 7) to be advantageous. This result can be explained in part by the increased gas turbine efficiency when operated at the optimum inlet temperature of 1612 F. In addition, an increased percentage of power will be produced by the gas turbine cycle with the trimburner systems. This will also increase the overall plant efficiency since the gas turbine cycle is more efficient than the steam cycle.

Figure 1.4 also points out that the effect of fuel drying on the trimburner systems' efficiency is
Figure 1.4 Gross and net efficiencies of all the systems, based on higher and lower heating values.

Figure 1.5 Gross, auxiliary, and net powers of all the systems.
negligible. Since the net efficiency of systems 4 and 7 are approximately equal, there appears to be no advantage in drying the fuel if trimburners are used. This implies that system 7 may have a substantial economic advantage over system 4 because no fuel preparation equipment is required.

Another important parameter to consider when comparisons are being made is that of the power output. The gross, auxiliary, and net power out of each system considered is shown in figure 1.5. The net power out is simply the gross power minus the power required to operate the plant itself (auxiliary). Again, it can be seen that the trimburner systems yield the most advantageous performance. This result is due directly to the increased output of the gas turbine at the higher inlet temperature. System 4 has a higher gross power output than system 7 due to the usage of a relatively dry fuel. This advantage is negated, however, by the high auxiliary power requirements of system 4. This high auxiliary power is directly attributable to the power requirements of the fuel preparation equipment. The result is that the net power output for systems 4 and 7 are the same. This indicates that system 7 is more desirable than system 4 since the same power output can be achieved without requiring the fuel preparation equipment.
These preliminary results appear to indicate that the trimburner concept should be employed in the final power plant design. In addition, system 7 offers a more compact, moveable plant design without compromising efficiency or net power. Since system 7 is apparently the best design option considered, it will be investigated in further detail. The remainder of this work will be dedicated to the study of viable trimburner designs for the system 7 variation of the power plant. Because this plant is to be wood-fired, the emphasis of this study will be on the design of trimburners which will use wood as the fuel source.
Chapter 2

WOOD CONVERSION PROCESSES

The previous chapter concluded that a trimburner incorporated into the combined cycle power plant has good potential. For the power plant to be entirely wood-fired, however, the trimburner must also use wood as its fuel source. This presents a problem in that the fuel used in the trimburner must burn cleanly since the combustion products will pass through the gas turbine. Because wood burns "dirty", it then becomes necessary to consider wood conversion processes capable of converting wood to a clean synthetic fuel. The synthetic fuel could then be burn directly in the trimburner as previously modelled. In this chapter pyrolysis, gasification, and methanol production are investigated as three wood-conversion processes which may be usable in the wood-fired trimburner concept. The discussion includes a brief description of the conversion techniques as well as addressing the applicability of each process to the specific requirements of the trimburner system.

PYROLYSIS

Pyrolysis is the breakdown of any biomass substance by heat at medium temperatures (200 C to 700 C) to yield
low energy gases, tars and oils, and solid char. The pyrolysis step is necessary in all combustion and gasification processes. Combustion results when the products of pyrolysis are oxidized for total heat release. In gasification, the production of low energy gas is encouraged by breaking down the oils and char produced in pyrolysis to gaseous components. This "cracking" process is an endothermic reaction requiring high temperatures.

There are two types of pyrolysis, slow and fast. The rate at which pyrolysis occurs is a function of both temperature and particle size [Robinson, 1980]. Low temperatures and large particle size result in slow pyrolysis whereas high temperatures and small particle size promote fast pyrolysis. The increased reaction rate of fast pyrolysis is attributed to increased thermal energy and larger surface area. The two major differences between fast and slow pyrolysis are 1) the relative amounts of char, oil, and gas formed, and 2) the chamber residence time of the particles. Slow pyrolysis requires long residence times and yields larger amounts of char. If the particle remains in the chamber long enough, equilibrium conditions can be approached. In fast pyrolysis, the chamber residence time is short and the production of low energy gas is enhanced. It is this type of pyrolysis that is encouraged in the gasification
process. The effect of higher temperatures on the promotion of synthesis gas is shown in figure 2.1.

Types of Pyrolyzers

There are three common types of commercial pyrolysis units available today: kiln, fixed-bed, and fluid-bed. These units vary in the way the wood particles are fed and heated in the reaction chamber, but the chemical processes are all the same. A brief description of each pyrolyzer type is given below.

Kiln pyrolyzers, as shown in figure 2.2, make use of standard and rotary chambers for wood pyrolysis. Wood travels through the chamber at a variable speed and is pyrolyzed at temperatures varying from 400 C to 600 C [ENFOR, 1983]. Conveying grates or rotating inclined chambers are used to move the wood through the unit. The heat to pyrolyze the wood is supplied in one of three ways: 1) indirectly through the chamber wall or in metallic tubes, 2) direct contact with hot combustion gases from a separate combustor, 3) burning some of the wood in the chamber itself and using the resulting hot combustion gases. Kiln pyrolyzers have an energy conversion efficiency on the order of 60 to 70 percent [ENFOR, 1983].

Because the residence time of the wood can be varied with kiln pyrolyzers, high moisture content wood can be
Figure 2.1 Effect of temperature on yield of pyrolysis products from wood [source: Scott and Piskorz, 1982].
Figure 2.2 Schematic diagram of kiln pyrolyzer manufactured by Pillard Incorporated.
used. A penalty is suffered at higher moisture contents due to the long residence time required. Output from the unit for a given heat input will decrease as a result of the lowered throughput rate of wood. In addition, high amounts of tars and oils are produced when high moisture content wood is used. Units using indirect methods of heating have a maximum moisture content limit of 20% due to the inability of the heat to be uniformly distributed among the wood feed [Reed, 1981]. Direct heating units can handle moisture contents up to 65%, but the product output will decline dramatically [ENFOR, 1983]. Kiln pyrolyzers that burn some of the wood in the chamber for the required heat have a practical moisture content limitation of about 35% due to limit on the amount of heating energy that can be supplied by the burning wood [ENFOR, 1985].

Fixed-bed pyrolysis units use a vertical reactor chamber as shown in figure 2.3. Wood enters the top of the unit and char exits the bottom. The pyrolysis gas exits either the top or the bottom of the unit depending on the design. The heat for these units is usually supplied by either burning some of the wood in the chamber or burning the low-energy gas formed in a separate combustor and discharging the resulting flue gas into the chamber. The normal operating temperature of fixed-bed
Figure 2.3 Schematic diagram of fixed-bed pyrolyzer manufactured by Enerco.
units is 400 C to 600 C. Fixed-bed pyrolyzers have an energy conversion efficiency of about 70 to 80 percent [Reed, 1981]. These values include the sensible heat energy contained in the products which may or may not be usable depending on the application. The main point to note, however, is that the conversion efficiency is somewhat higher than that of the kiln design.

Fixed-bed pyrolyzers are classified according to the relative motion of wood feed and hot gases. Units where the wood enters the reactor top and the thermal energy enters the bottom are called up-draft pyrolyzers. When the wood and the agent supplying the thermal energy traverse the reactor in the same direction, the unit is called a downdraft pyrolyzer. When some of the wood is burned in the chamber, the pyrolyzer operates essentially like a gasifier designed to produce char.

Because of the vertical design of a fixed-bed pyrolyzer, the feed size is very important. The wood particles should be a consistant combination of chips and small blocks to prevent channeling of the hot gases through the chamber. Channelling occurs when the pile of wood inside the chamber is not of uniform density. The gases will flow through the wood in the path of least resistance, forming the channels in the wood pile. This situation most often arises when the wood feed contains large amounts of sawdust [ENFOR, 1983].
The maximum moisture content that can be tolerated for a fixed-bed pyrolyzer is on the order of 50% for up-draft units and 25% for down-draft units [ENFOR, 1983]. This difference is a result of the drying that occurs in the up-draft design. Because the hot pyrolysis gas rises upward in the up-draft design, the sensible heat can be put to use in drying the incoming wood. Conversely, the down-draft design has the pyrolysis gas flowing downward so that there is no contact between the hot gas formed and the incoming wet wood. The result is that the up-draft pyrolyzer produces a cool, dirty pyrolysis gas, whereas the down-draft unit produces a hot, clean gas but has a lower moisture content limit [Foley and Barnard, 1983].

Fluid-bed pyrolyzers employ the same concept as fluid-bed combustors but operate at a lower temperature to encourage char production [ENFOR, 1983]. Figure 2.4 displays a typical fluid-bed pyrolyzer with the associated gas cleaning equipment. The fluid material used is usually sand or air. Heat for the fluid is supplied by either burning the pyrolysis gas in a separate combustor or using a conventional fuel in an external combustor. The fluid is heated in the external combustor and then is blown into the pyrolysis chamber with the wood particles. The fluid exits the chamber with the formed char and pyrolysis gas. Separation of the pyrolysis gas, char, and
Figure 2.4 Schematic diagram of fluidized-bed pyrolyzer manufactured by Alberta Industrial Development Ltd.
heating fluid is usually accomplished by using a series of cyclones. The energy conversion efficiency for these units ranges from 75 to 85 percent, depending on operating conditions [ENFOR, 1983].

The biggest advantage of the fluid-bed pyrolyzer is the quick and uniform temperature distribution in the chamber. Because of this, uneven particle sizes can be handled without difficulty. This advantage is somewhat negated by the amount of post-process cleaning and separation that is required. In addition, the maximum moisture content that can be handled by these units is only about 25% [ENFOR, 1983].

Advantages

The major advantage of the pyrolysis process is the ability to control the form of the products. Pyrolyzers are extensively used in the area of charcoal briquet production for this reason [Robinson, 1980]. By controlling the pyrolysis temperature and chamber residence time, the products will vary from mainly solid char to low-energy gas.

Because of the low temperature required to initiate pyrolysis, many pyrolyzers are used in heat recovery systems [Miller, 1983]. Flue gas exiting combustors and boilers in power generation plants can be channeled into pyrolyzers to produce a desired gaseous or solid product.
This offers flexibility to the plant by giving it the ability to produce various fuel sources.

Disadvantages

There are several disadvantages to the use of pyrolyzers solely for gaseous fuel production. Because most of the commercially available pyrolysis units are designed for solid char production, maximizing gaseous products would be difficult. In fact, a pyrolyzer designed to yield mainly low-energy gas is called a gasifier.

Another disadvantage of most pyrolyzers is the restriction imposed on the maximum allowable wood moisture content. Many types of pyrolyzers cannot handle wood moisture contents in excess of 25%. The units that can use higher moisture content wood produce large quantities of undesirable tars and oils as well as suffering a significant decrease in throughput rate. In addition, the inlet feed must be of uniform size (with the exception of a fluidized bed unit) to insure a consistent chamber residence time. The type of product formed will depend on residence time as well as operating temperature. Larger size feed require longer residence times to produce a gaseous product. If high throughput rates are required the size of the feed must necessarily be small. This,
coupled with the moisture content restriction, indicate the need for a substantial amount of inlet feed preparation.

The production of tars and oils in the pyrolysis process present a disadvantage as well. The tars and oils formed are difficult to use in conventional combustors and present environmental problems when disposing [Tillman, et. al., 1982; Reed, 1981]. Because these products contain energy, any tars and oils produced and not used will decrease the overall conversion efficiency. A typical conversion efficiency for a pyrolyzer operating at moderate temperatures is 45% [ENFOR, 1983].

GASIFICATION

In general, pyrolyzers are not designed to enhance the production of gaseous products. Many of the restrictions of the pyrolysis units mentioned above are due to the design emphasis being on solid char production. Pyrolysis units which are designed specifically for low-energy gas production are called gasifiers. Many of the restrictions which apply to pyrolyzers do not apply to gasifiers because of several design changes.

Gasification is basically a pyrolysis process occurring at high temperatures and/or pressures. The chamber design is usually vertical so that heat flow can be utilized to dry the incoming feed. Gasification of
biomass occurs in four zones inside the chamber: oxidation, reduction, pyrolysis, and drying [Reed, 1981]. The top-to-bottom order of these zones vary depending on the type of gasifier. An example of this arrangement is shown in figure 2.5 for an up-draft gasifier design.

The important reactions which occur in the gasifier take place in the reduction and pyrolysis zones. These reactions, listed in table 2.1, represent the chemical breakdown of the biomass feed to the gaseous products [Desrosiers, 1977].

Table 2.1 IMPORTANT REACTIONS IN GASIFICATION PROCESS

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Energy Released/Consumed @ 298 K (KJ/KMOL)</th>
</tr>
</thead>
<tbody>
<tr>
<td>a) CO + H₂O = CO₂ + H₂</td>
<td>-41.2</td>
</tr>
<tr>
<td>b) C + 2H₂ = CH₄</td>
<td>-74.93</td>
</tr>
<tr>
<td>c) C + H₂O = CO + H₂</td>
<td>131.4</td>
</tr>
<tr>
<td>d) C + CO₂ = 2CO</td>
<td>172.6</td>
</tr>
<tr>
<td>e) C + O₂ = CO₂</td>
<td>-393.8</td>
</tr>
</tbody>
</table>

As table 2.1 indicates, some of the reactions occurring are endothermic and require energy. The energy for these reactions is supplied by the oxidation zone where some of the inlet feed is burned to completion. This is accomplished by supplying the gasifier with a less than
Figure 2.5 Schematic diagram of an operating chamber for a typical up-draft gasifier.
stoichiometric amount of air or oxygen. It is important to note that the reactions listed in table 2.1 do not reach equilibrium conditions. The water-shift reaction:

\[ \text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2 \]

occurs on the surface of the inlet feed chips and therefore, has the longest reaction time to obtain equilibrium. Studies indicate that this reaction is of paramount importance in the gasification process because equilibrium conditions are nearly obtained in most gasifier chambers [Graham and Huffman, 1981].

The products of the reactions occurring in the pyrolysis and reduction zones makeup the composition of the low-energy producer gas. These products exit the pyrolysis zone at high temperatures and aid in drying the incoming feed. The degree of drying which occurs depends on the chamber design.

Modes of Gasification

The gasification of biomass can occur a number of different ways. The difference in these processes lies in the manner in which the combustion oxygen is supplied to the oxidation zone. The three modes of gasification discussed here are air, oxygen, and steam gasification.

Air gasification is perhaps the most common technique used in supplying combustion oxygen. A less than
stoichiometric amount of air is fed to the oxidation zone for biomass combustion. Because of the large amount of nitrogen present in the air, the final producer gas will have a relatively low volumetric heating value.

Oxygen gasification is the feeding of pure oxygen to the oxidation zone in less than stoichiometric amounts. This yields a producer gas with relatively high heating values due to the absence of nitrogen. Because of the high cost of oxygen production, however, this process is unreasonable in all but a few isolated applications.

A third process, steam gasification, utilizes steam to obtain the combustion oxygen required. This process is advantageous in that it does not introduce any inert elements such as nitrogen into the gasifier chamber. In addition, steam gasification allows for better control of the gasifier chamber temperature. Although steam addition slightly increases the production of methane and hydrogen, it drastically increases the amount of water vapor in the products as well. This increases the amount of gas cleaning equipment required as well as reducing the amount of drying which takes place in the gasifier.

Types of Gasifiers

There are several types of gasifiers commercially available today. The three common styles of gasifiers are
the up-draft unit, and down-draft unit, and fluidized bed units. Each type of gasifier exhibits different performance characteristics due to difference in design. These gasifier types are briefly discussed below.

The up-draft, or counter-current gasifier, is the most common of the gasifiers available today. The name indicates that the combustion air is fed from underneath the unit and filters upward. As shown in figure 2.5, the up-draft gasifier has the oxidation zone on the bottom and the reacting zones above it. This simple design allows the up-draft gasifier to be scaled up to very large capacities [Foley and Barnard, 1983]. The hot combustion gas rises from the bottom zone to initiate pyrolysis and produce the low-energy gas. The hot producer gas then rises through the wet incoming feed cooling as it rises. This arrangement allows for a large amount of drying to take place in the drying zone. Up-draft gasifiers can operate with inlet feeds having a moisture content as high as 50% w.b. [Reed, 1981]. The producer gas exiting the up-draft units has relatively low temperatures of 100-200 C and high water vapor concentrations [ENFOR, 1985].

Although pyrolysis does occur at high temperatures in the up-draft gasifier, some tars and oils are still formed. These products are generally in vapor form when exiting the gasifier. These tars and oils entrained in the producer gas are what stimulate the need for gas
cleaning equipment [Gallagher, 1979; Garland and Gluckman, 1979]. The amount of tar and oil exiting up-draft gasifiers is on the order of 3% by volume of the exiting gas and contain 8-10% of the energy in the inlet feed [ENFOR, 1983]. A typical overall conversion efficiency (biomass-to-producer gas) of an up-draft gasifier is approximately 70% [Sofer and Zaborsky, 1981; ENFOR, 1983].

Down-draft, or co-current gasifiers are similar to the up-draft units with two subtle differences. As shown in figure 2.6, the order of the gasifier zones is different with the reduction zone being on the bottom. In addition, the combustion air is fed near the top and travels downward to the oxidation zone. Although pyrolysis of the feed still occurs above the oxidation zone, the flow of the gases is such that the pyrolysis products are pulled back down through the hotter oxidation and reduction zones and exit the bottom of the unit. This type of arrangement results in different performance characteristics than the up-draft unit.

The main advantage of the down-draft unit is the reduction of tars in the product gas [Synthetic Fuels Assoc., 1982; Reed, 1981; ENFOR, 1983]. This is primarily a result of the producer gas and tars flowing through the hotter combustion and reduction zones. The high temperatures in these zones cause the tars to "crack" or
Figure 2.6  Schematic diagram of an operating chamber for a typical down-draft gasifier.
completely burn, yielding a clean producer gas virtually free of tars and oils. Reduction of the tars also occur to produce a higher content of hydrogen and hydrocarbons in the exiting producer gas. This reduction and oxidation of the tars and oils enable the reactions in table 2.1 to nearly obtain equilibrium conditions. Also, the energy contained in the tars is no longer lost and thus efficiency is higher than that of up-draft units. Typical values for the conversion efficiency of downdraft gasifiers are on the order of 85% [Reed, 1981; ENFOR, 1983].

Because the hot producer gas does not flow upward through the wet inlet feed, very high exit temperatures can be obtained with downdraft gasifiers. The drawback, however, is that very little drying occurs in the gasifier chamber. For this reason, the maximum moisture content of the inlet feed to a downdraft unit is limited to 15-20% [Reed, 1981; ENFOR, 1983]. In addition, downdraft gasifiers are very sensitive to changes in the moisture content of the inlet feed.

The major disadvantage of the down-draft gasifier is the difficulty in scaling them up to sizes larger than 300 kW output [Foley and Barnard, 1983]. This problem is due primarily to the configuration of the down-draft chamber. Because the reduction of tars in the producer gas depends on the gas passing through the high temperature combustion
zone, it is critical that no local "cold spots" exist in the oxidation zone. The presence of cold spots will not only retard combustion, but will cause the unit to gum up with condensed tars until operation will cease altogether. For this reason, the chambers are tapered to a throat at the oxidation zone which allows for even dispersion of air coming in through side tuyeres. The even dispersion of air insures that no oxygen starved spots will be present and, therefore, combustion can occur. As the capacity requirements of the unit increases, however, the chamber dimensions (including the throat) must also increase accordingly. This increased area at the throat makes it difficult for the combustion air to disperse to the inner portions of the combustion zone. The result is cold spots forming where there is insufficient air to promote combustion and the unit fails to operate.

Fluidized bed gasification units are the subject of much research in the area of biomass conversion. Figure 2.7 displays a typical fluidized bed gasifier with associated gas cleaning and combustion equipment. These units are very similar to the fluidized bed pyrolyzers but operate at much higher temperatures. The operating fluid is generally sand or aggregate which suspends the wood and provides the uniform temperature field. Although the fluidized bed unit can handle varying size feed chips, it
Figure 2.7 Schematic diagram of a typical fluidized-bed gasifier with associated gas cleaning equipment.
is still sensitive to moisture content. The maximum moisture content that fluidized bed gasifiers can handle is approximately 25-30% [Reed, 1981]. An extensive amount of gas cleaning equipment is required for these units to separate the producer gas and the operating fluid.

Because of the quick and uniform heating that is obtainable with fluidized bed units, high throughput rates can be achieved. This results in a smaller unit size than the fixed bed (up-draft and down-draft) units for a given capacity. The high heating rates also give the fluidized bed gasifier the ability to quickly alter the output of producer gas to accommodate load variations. Uniform heating at moderate throughput rates have yielded conversion efficiencies in excess of 90% for several small bench-scale prototypes [ACS Symposium, 1980].

Since the field of fluidized bed combustion and gasification is relatively new, it is difficult to determine various performance parameters for full size units. Also, separate combustion units are required to heat the moving fluid, thus increasing cost. As research in this field increases, it is anticipated that fluidized bed gasification will become a more attractive option when considering biomass conversion processes.

Advantages

As can be ascertained from the above discussion, the
relative advantages of gasification vary depending on the type of gasifier considered. In general, the gasification process is designed to maximize producer gas output and, therefore, should be considered when gaseous conversion is required. In addition, the gasification process has a higher conversion efficiency when producer gas is deemed the desired product.

Depending on the type of gasifier selected, less feed preparation is required in gasification units than in pyrolysis units. Biomass of varying size and up to 50% moisture content can be effectively handled by some types of gasifiers. This is advantageous in reducing system auxiliary power required as well as decreasing system complexity.

Disadvantages

The major disadvantage of all gasification units is the requirement of gas cleaning equipment to yield the producer gas in a usable form. Although producer gas is the cleanest fuel obtainable from biomass conversion, it still requires additional cleaning to be combustible in most turbomachinery applications [Gallagher, 1979; Garland and Gluckman, 1979].

Other various disadvantages of the gasification processes exist depending on the type of gasifier
considered. The main points to note here are the inability to scale up down-draft units and the relative infancy of the fluidized bed gasification field.

METHANOL PRODUCTION

Methanol, CH₃OH, is a primary alcohol with an average heat of combustion of 23.85 MJ/kg. The concept of using methanol as a fuel is not a new one. In the early 1900's, methanol was produced by distillation of wood and used as city gas [Chermisinoff, et. al., 1980]. Today, a large portion of the methanol produced is converted to formaldehyde and there is little emphasis on large scale methanol production for fuel considerations.

There are two primary processes that can be used to produce methanol [Robinson, 1980]. The most common practice today is to produce methanol from the slow oxidation of natural gas as shown below:

\[ 2\text{CH}_4 + \text{O}_2 \longrightarrow 2\text{CH}_3\text{OH} \]

A second technique used to produce methanol utilizes producer gas from biomass distillation processes to initiate the following gas feedstock reactions:

\[ \text{CO}_2 + 3\text{H}_2 \longrightarrow \text{CH}_3\text{OH} + \text{H}_2\text{O} \]
\[ \text{CO} + 2\text{H}_2 \longrightarrow \text{CH}_3\text{OH} \]
The distillation process most often used is gasification since the desired reactants (CO$_2$, H$_2$, & CO) naturally exist in the producer gas exiting the gasifier. The production of methanol is therefore a continuation, or secondary conversion process.

Although conceptually simple, the actual production of methanol from biomass is somewhat more difficult. There are currently two industrial processes which are used to create methanol from biomass producer gas [Cheremisinoff, et. al., 1980; Robinson, 1980]. The high-pressure process has existed for many years and is well defined. Overall conversion efficiency from biomass to methanol for the high pressure process is approximately 55% based on heating values of the inlet biomass and exit methanol [Robinson, 1980]. The low-pressure process is relatively new and is still somewhat experimental. Overall conversion efficiency of this process has been measured to be 30-40% [Robinson, 1980]. Because the high-pressure process is more efficient and is widely accepted in industry, it will be the process discussed here.

A flow diagram of a typical high-pressure methanol production plant is shown in figure 2.8 [Robinson, 1980]. The producer gas is fed into the plant after being cleaned and cooled. The gas is compressed to about 7 atm and treated to remove nitrogen, excess carbon dioxide, and any
Figure 2.8 Flow diagram of a typical high-pressure methanol production plant.
hydrocarbons present. After further compressing the gas to about 27 atm, it is fed into the shift reactor and stoichiometrically shifted to the appropriate ratio of hydrogen to carbon monoxide. This is done by using steam injection and a metallic catalyst. Excess carbon dioxide formed in the shift reactor is removed and the gas is again compressed to approximately 170 atm. Compression of the gas to high pressures favors the production of methanol in the gas feedstock reactions and is the main reason why the conversion efficiency is higher for the high-pressure process. Finally, crude methanol is formed from hydrogen and carbon monoxide in the methanol reactor. Depending on the application, the crude methanol is cleaned and refined. The average exit temperature of methanol from a high-pressure process is approximately 100-300 degrees centigrade, depending on the amount of gas cleaning required.

Advantages

The advantage of producing methanol for use as a fuel in the trimburner system is the consistancy of the composition supplied to the air stream burner. The pure methanol fuel will have a higher energy content than the raw producer gas and will therefore, have lower volumetric flow rates. The combustion efficiency of methanol will
also be higher than that of producer gas due to the absence of nitrogen and other inerts in the incoming fuel.

Disadvantages

The obvious disadvantages of producing methanol for use in the trimburner system are the amount of equipment required (and the associated capital cost) and the low conversion efficiency. As seen in figure 2.8, the amount of hardware required to produce methanol far exceeds that required for pyrolysis or gasification. To use the high-pressure methanol production process an entire gasification system to produce the required raw producer gas would be needed as well. In addition, high auxiliary power requirements exist for methanol production due to the amount of gas compression which is done. The high cost of producing methanol, coupled with the relatively low conversion efficiency, yield this alternative unpractical in most instances.

CONCLUSION

In reviewing the three conversion processes considered in this study, it appears that biomass gasification is the apparent best choice for the trimburner system. The high cost and low conversion efficiency of methanol production negate any advantages that might be realized with this option. The amount of
non usable product created in addition to the severe moisture content restriction yield the pyrolysis process unpractical for the trimburner system considered. Gasification, on the other hand, yields a product which, once cleaned, satisfies the constraints imposed by the trimburner system quite well. Also, the anticipated high moisture content and non-uniformity of the incoming biomass feed can still be used by some of the types of gasifiers discussed.

With gasification selected as the best conversion process to use in the trimburner system, the next question was to determine what type of gasifier is best suited to this application. When the condition of the as received biomass is considered, the choice becomes obvious. Both the fluidized bed and the down-draft gasifiers have maximum allowable inlet moisture contents which are far below the anticipated values of the as received biomass to the plant. In addition, the down-draft unit cannot be scaled up to capacities required by the trimburner system. The ability of the up-draft gasifier to handle high moisture content feeds and scale up to large capacities make the use of these types of units much more practical in the trimburner application. The decision therefore, was to use an up-draft gasifier to convert the as received biomass to a clean fuel for combustion in the trimburner of the biomass power plant.
Chapter 3

GASIFIER/TRIMBURNER SUB-SYSTEM DESIGN

In was concluded in the previous chapter that biomass gasification appeared to be the most practical conversion process for the trimburner application in systems 5 and 7. In addition, the decision to use an up-draft gasifier design was made based on the advantage these units possessed in scaling up and handling high moisture feeds. With a conversion process chosen, the next step was to design a trimburner/gasifier sub-system to incorporate into system 7 of the biomass plant. Three designs which satisfy the objectives of the trimburner/gasifier sub-system are formulated and discussed in this chapter.

DESIGN CONSIDERATIONS

In designing possible gasifier/trimburner sub-systems, certain considerations had to be addressed. These considerations ultimately determine the specific design of the sub-system. The major concerns in the design of the trimburner/gasifier sub-system were: 1) producer gas cleaning and conditioning, 2) compression of the producer gas to facilitate flow into the trimburner, and 3) utilization of the sensible heat energy contained in the raw producer gas exiting the gasifier.
Producer Gas Cleaning and Conditioning

A significant amount of producer gas conditioning was required to yield a clean burning gaseous fuel which could pass through the gas turbine blading. The detailed analysis of this gas conditioning was not the scope of this study and will not be presented. Instead, a brief overview of possible gas conditioning equipment was conducted and selection was made based on simplifying assumptions. It was assumed that the removal of all the tars and oils entrained in the producer gas would yield a sufficiently clean gaseous fuel. The validity of this assumption is based on the fact that only the clean combustible gases hydrogen, carbon monoxide, and methane are present in the tar-free producer gas. Since condensation of the tars and oils begin at approximately 100 C, it was further assumed that lowering the producer gas temperature to 25 C will remove all tar and oil from the gas flow.

The removal of the entrained tars and oils can be accomplished a number of ways. Various scrubbers and precipitation devices could be employed to remove the oil mists from the flow. The gas cleaning system proposed for the trimburner/gasifier sub-system was a wet scrubber combination of a venturi scrubber in series with a dry plate scrubber [Reed, 1981]. The venturi scrubber is
assumed to remove all the oil and tar mists entrained in the producer gas flow. The tar-free gas then enters the dry plate scrubber where all water vapor is assumed to be removed. The resulting producer gas flow exiting the gas cleaning equipment is assumed to be tar-free and moisture free at 25 C. Pressure drops through the cleaning equipment were neglected so that the gas pressure exiting the cleaning equipment was assumed to be equal to the gas pressure in the gasifier.

Sensible Energy Recovery

The biggest disadvantage of cleaning the raw producer gas exiting the gasifier is the loss of the sensible energy contained in the flow. Invariably, passing the producer gas through the cleaning equipment results in a necessary reduction in the gas temperature to condense the tars and oils. This temperature reduction signifies a loss in energy in the sub-system. In an attempt to utilize the sensible energy contained in the raw producer gas, a heat exchanger was incorporated into the sub-system at the gasifier exit [Gallagher, 1979; Garland and Gluckman, 1979]. The purpose of this heat exchanger was to transfer heat energy from the raw producer gas to the clean, conditioned producer gas exiting the cleaning equipment. In reality, this arrangement may not be feasible due to the unavoidable gumming of the heat
exchanger that could result as the tars and oils in the raw gas stream cool and begin to condense out. The assumption of the use of such a heat exchanger, however, is useful in determining the effect of utilizing the sensible energy in the raw producer gas. A possible configuration for this heat exchanger might be a simple shell-and-tube arrangement with the raw gas flowing in the shell and the clean producer gas flowing through the tubes.

Producer Gas Compression

Another important consideration in the sub-system design was that of compressing the producer gas. Because the air stream in which the producer gas is to be combusted is pressurized, compression of the fuel gas is essential. The operating pressure of the air stream entering the trimburner component is approximately 129 psia. Therefore, the pressure of the producer gas stream must be at least 129 psia for flow into the trimburner to occur.

SUB-SYSTEM DESCRIPTION

A general gasifier/trimburner sub-system design was developed based on the design considerations discussed above. To incorporate the various ways in which the
producer gas could be compressed, three variations on the basic sub-system design were investigated. These sub-system variations, designated A, B, and C, are described below.

Sub-System A

Sub-system A, as shown in figure 3.1, incorporates compression of the producer gas after it has been cleaned and cooled. An up-draft gasifier operating at atmospheric pressure converts biomass to raw producer gas. The raw gas flows through a heat exchanger where it transfers sensible energy to the clean producer gas entering the trimburner. The raw gas is then cleaned by the wet scrubber equipment and compressed in the producer gas compressor. The clean, compressed gas flows back through the heat exchanger to claim the sensible energy of the raw producer gas exiting the gasifier. The heated, high pressure producer gas is then combusted in the trimburner to boost the temperature of the air stream entering the gas turbine.

Sub-System B

The effect of operating a pressurized up-draft gasifier is considered with sub-system B. Sub-system B is similar to sub-system A with the exception that the required compression occurs in the gasifier. The raw
Figure 3.1 Schematic diagram of sub-system A gasifier/trimburner design.
producer gas exiting the gasifier, therefore, is also compressed. Pressure losses through the heat exchanger and cleaning equipment are neglected so that the clean producer gas entering the trimburner is assumed to be at the gasifier pressure. A schematic of sub-system B is shown in figure 3.2.

Sub-System C

The concept of a pressurized gasifier is again considered with sub-system C. However, this design requires no additional compressor to pressurize the product gas. As shown in figure 3.3, pressurized air is bled from the main compressor and used as the combustion air in the pressurized gasifier. The remainder of sub-system C is identical to that of sub-system B. It should be noted that the total mass flow rate of air flowing through the gas turbine in this sub-system will be less than the flow rates in sub-systems A and B. This is an unavoidable result of bleeding some of the compressed air from the main compressor to operate the pressurized gasifier.
Figure 3.2 Schematic diagram of sub-system B gasifier/trimburner design.
Figure 3.3 Schematic diagram of sub-system C gasifier/trimburner design.
Chapter 4

GASIFIER/TRIMBURNER SUB-SYSTEM MODELLING

To determine the performance of the three sub-systems described in the previous chapter, a computer model of each sub-system was developed. The development of these models was based on thermodynamic relations which govern the operation of the components that make up each sub-system. This chapter presents an analytical discussion of the development of these component models. It should be noted that the sub-system models are based on theoretical thermodynamic concepts and, therefore, represent an idealized situation.

SUB-SYSTEM COMPONENT MODELLING

The three gasifier/trimburner sub-systems modelled were similar in that each sub-system contains a gasifier, heat exchanger, gas cleaning equipment, and a trimburner. The only variation in modelling occurs in the location of a gas compressor. For this reason, general models for each component of the sub-system were developed and then integrated together differently to simulate the three sub-system variations.

Gasifier

Simulation of the up-draft gasifier was simplified by
making certain generalizing assumptions. These assumptions were made in an effort to simplify the complex nature of the gasification process while maintaining accurate results. This approach was considered valid in that the scope of this study was not to formulate a detailed gasification model, but to consider the merits of a wood-fueled trimburner as applied to the given biomass plant.

In representing the idealized case, it was assumed that all of the wood not consumed in the combustion zone was converted to raw producer gas. The formation of gasifier tars and oils is a complex function of reaction rate, transient temperature gradients, and non-equilibrium reactor conditions. The assumption of total gaseous conversion then, can be seen to simplify the problem dramatically.

In reality it is known that the solid char formed during pyrolysis is generally the material which is combusted in the oxidation zone. Since the model does not account for char production, it was assumed that wood was the material combusted in the combustion zone. To facilitate this assumption the wood was divided into two separate streams. One stream contains wood to be combusted to the combustion zone. The remainder of the wood is assumed to gasify completely in the pyrolysis zone. This assumption appears valid in that it accurately
accounts for the wood that is unavailable for gaseous conversion due to char combustion in the oxidation zone.

For the purpose of modelling, the up-draft gasifier was divided into two sections as shown in figure 4.1. The drying section simulates the drying of the incoming wet wood by the rising hot producer gas. The oxidation/pyrolysis section simulates the combustion of a small amount of wood and subsequent pyrolysis that occurs from the heat released. Known inputs to the gasifier are assumed to be: air fuel ratio (equivalence ratio), ambient air temperature, wood moisture content (wet basis), heating value of dry wood, and inlet wood and air composition. The desired outputs from the model are raw producer gas composition, temperature, and flow rate.

The chemical equation used to describe the pyrolysis of wood is given as:

\[
\text{CH}_{1.4055} + (\bar{\Phi})(1.055)\text{O}_2 + (\bar{\Phi})(1.055)(3.76)\text{N}_2 \rightarrow (c)\text{H}_2 + (d)\text{H}_2\text{O} + (e)\text{CO} + (f)\text{CO}_2 + (g)\text{CH}_4.
\]

where

\[
\begin{align*}
\text{CH}_{1.4055} & = \text{chemical composition of wood} \\
\bar{\Phi} & = \text{equivalence ratio: } (\text{actual air/fuel ratio})/(\text{stoic. air/fuel ratio}) \\
c,d,e,f,g & = \text{unknown composition of raw producer gas} \\
1.055 & = \text{Moles of } \text{O}_2 \text{ required for stoichiometric combustion}
\end{align*}
\]
Figure 4.1 Schematic diagram of flow model used to simulate up-draft gasifier operation.
The chemical basis of 1 mole of dry inlet wood will be used throughout the analysis for the up-draft gasifier.

The composition of the raw gas is a function of the five pyrolysis chemical reactions shown in chapter 2. As mentioned earlier, these reactions do not progress to equilibrium conditions in the gasifier chamber. The water-gas reaction:

$$\text{CO} + \text{H}_2\text{O} \rightarrow \text{H}_2 + \text{CO}_2$$

does, however, progress to 90-95% of the equilibrium condition in most gasification processes [Graham and Huffman, 1981]. This observation is due to the fact that this reaction occurs on the surface of the wood and therefore, reacts continuously for the longest period of time. In addition, the equilibrium constant for this reaction is significantly higher than the other reactions at the low temperatures at which gasification occurs. Because of this, the assumption was made that the water-gas reaction is the only reaction of significance during the pyrolysis process.

The pyrolysis equation above defines five unknowns for the composition of the raw producer gas. To solve for these unknowns, a set of five simultaneous equations is required. One equation is provided by the chemical
equilibrium of the water-gas reaction in terms of molar composition:

\[(c)(f)/(e)(d) = K_{eq} \text{ (at specified temp.)}\]

where

\[K_{eq} = \text{Equilibrium constant for water-gas reaction}\]

The remaining four equations were obtained by considering mass and energy balances for the oxidation/pyrolysis section. The mass balance yields three significant equations for the different elements:

(2) Hydrogen: \[1.4 = 2(c) - 2(d) - 4(g)\]

(3) Carbon: \[1.0 = (e) - (f) - (g)\]

(4) Oxygen: \[.59 - (2)(1.055)(X) = (d) - (e) - 2(f)\]

If the reacting chamber of the gasifier is assumed to be adiabatic, the energy balance for the oxidation/pyrolysis section can be derived:

\[\sum_{\text{react.}} n_r (h_f^r) + \int_{\text{Tamb.}}^{\text{Tent.}} C_{p,\text{react}} \, dT = \sum_{\text{prod.}} n_p (h_f^p) + \int_{\text{Tamb.}}^{\text{Tpyr.}} C_{p,\text{prod}} \, dT\]
where

\[ \eta_r = \text{number of moles of reactant} \]
\[ \eta_p = \text{number of moles of product} \]
\[ h_r^f = \text{standard heat of formation of reactant} \]
\[ C_{p,\text{react}} = \text{molar specific heat of reactant} \]
\[ C_{p,\text{prod}} = \text{molar specific heat of product} \]
\[ T_{\text{amb.}} = \text{ambient temperature} \]
\[ T_{\text{pyr.}} = \text{temperature at which pyrolysis occurs} \]
\[ T_{\text{ent.}} = \text{temperature of reactants entering gasifier} \]

Expanding the energy balance yields:

\[ (5) \quad \Delta h_{\text{wood}} = (c) \int_{T_{\text{ent}}}^{T_{\text{pyr}}} C_{p,H_2} dT + (a) \left[ h_{H_2O}^f + \int_{T_{\text{ent}}}^{T_{\text{pyr}}} C_{p,H_2O} dT \right] \]
\[ + (e) \left[ h_{CO}^f + \int_{T_{\text{ent}}}^{T_{\text{pyr}}} C_{p,CO} dT \right] + (f) \left[ h_{CO_2}^f + \int_{T_{\text{ent}}}^{T_{\text{pyr}}} C_{p,CO_2} dT \right] \]
\[ + (g) \left[ h_{CH_4}^f + \int_{T_{\text{ent}}}^{T_{\text{pyr}}} C_{p,CH_4} dT \right] + (\phi)(0.55 \times 3.76) \left[ \int_{T_{\text{ent}}}^{T_{\text{pyr}}} C_{p,H_2} dT \right] \]

For the energy balance above, it was assumed that the wood enters the oxidation/pyrolysis section at essentially ambient conditions. The correlations used to describe the temperature dependancy of the specific heats for the substances in the above equation are given in Table 4.1.

The only unknown yet to be resolved in the oxidation/pyrolysis section was that of the chamber
TABLE 4.1 SPECIFIC HEAT EQUATIONS FOR GASIFIER ANALYSIS  
(Van Wylen, Sontag, 1978)

<table>
<thead>
<tr>
<th>SUBSTANCE</th>
<th>SPECIFIC HEAT EQUATION (Btu/mole °R)</th>
<th>% ERROR</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrogen</td>
<td>13.505 - 167.96(Z⁻₀.７₅) + 278.44(Z⁻¹) - 134.01(Z⁻¹.₅)</td>
<td>.60</td>
</tr>
<tr>
<td>Water Vapor</td>
<td>34.190 - 43.868(Z⁰.₂₅) + 19.778(Z⁰.₅) - 0.88407</td>
<td>.43</td>
</tr>
<tr>
<td>Carbon Monoxide</td>
<td>16.526 - 0.16841(Z⁰.₇₅) - 47.985(Z⁻⁰.₅) + 42.246(Z⁻⁻⁰.₇₅)</td>
<td>.42</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>- 0.89286 + 7.2967(Z⁰.₅) - 0.98074(Z) + 0.0057835(Z²)</td>
<td>.19</td>
</tr>
<tr>
<td>Methane</td>
<td>- 160.82 + 105.10(Z⁰.₂₅) - 5.9452(Z⁰.₇₅) + 77.408(Z⁻⁰.₅)</td>
<td>.15</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>9.3355 - 122.56(Z⁻¹.₅) + 256.38(Z⁻²) - 196.08(Z⁻³)</td>
<td>.43</td>
</tr>
<tr>
<td>Wood*</td>
<td>6.075 + 0.007354(Tf + Ti - 64.0)</td>
<td></td>
</tr>
</tbody>
</table>

Where:  
Z = T / 180  T is substance temperature in °R  
Ti = Initial temperature of wood in °F  
Tf = Final temperature of wood in °F  

* experimental correlation from Beall, 1968.
operating temperature. Because of the temperature
dependency of the equilibrium constant in equation (1) and
the product enthalpies in equation (5), it was necessary
to determine this value. To simplify the gasifier
analysis, it was assumed that this temperature remained
constant during the pyrolysis reaction. The pyrolysis
temperature depends on the temperature of the gaseous
combustion products filtering up from the oxidation zone.
For this model, it was assumed that some of the wood
combusts stoichiometrically with the amount of air
introduced into the gasifier as shown:

\[
(\Phi) \text{CH}_14\text{O}_{59} + (\Phi)(1.055)\text{O}_2 + (\Phi)(1.055)(3.76)\text{N}_2 \\
(\Phi)\text{CO}_2 + (\Phi)(.7)\text{H}_2\text{O} + (\Phi)(1.055)(3.76)\text{N}_2
\]

It can be seen that the amount of wood which is consumed
in the oxidation process is a function of the specified
operating equivalence ratio (\(\Phi\)). As the equivalence ratio
approaches 1.0 more wood is burned and less wood is
available for gasification. Modelling in this manner
allows for simulation which follows trends observed in
actual gasification units.

The exit temperature of the combustion products is
determined by considering an energy balance for the wood
combustion reaction above:
By iterating on the exit temperature of the products, the solution to the above energy equation can be obtained. This temperature is called the adiabatic flame temperature of the wood since the ideal case in which no energy losses occur is assumed.

As the hot combustion products (CO$_2$, H$_2$O, N$_2$) at the adiabatic flame temperature rise through the remaining un-combusted cool wood, thermal energy transfer occurs. This transfer of energy continues until both the wood and gas are at the same equilibrium temperature. It is this temperature at which pyrolysis is assumed to occur. The heat transfer between the hot gas and wood can be expressed in equation form as:

\[
(1 - \bar{\Phi}) \int_{T_{\text{amb}}}^{T_{\text{pyr.}}} C_{p_{\text{wood}}} dT = (\bar{\Phi}) \int_{T_{\text{pyr.}}}^{T_{\text{flame}}} C_{p_{\text{CO}_2}} dT + (.7)(\bar{\Phi}) \int_{T_{\text{pyr.}}}^{T_{\text{flame}}} C_{p_{H_2O}} dT
\]

\[
+ (\bar{\Phi} \times 1.055)(3.76) \int_{T_{\text{pyr.}}}^{T_{\text{flame}}} C_{p_{N_2}} dT
\]

where

\( T_{\text{FLAME}} \) = flame temperature of gas combustion products
where

\[ C_{p_{wood}} = \text{molar specific heat of wood} \]

Again, because of the temperature dependancy of the specific heats, iteration is required to determine the pyrolysis temperature. With the pyrolysis temperature determined, the chemical equilibrium constant in equation (1) can be determined. This was done by curve fitting equilibrium constant data for the water-gas reaction over the anticipated range of temperatures. The most accurate curve fit was found by dividing the overall temperature range into four smaller sections and fitting second order polynomials to each section. The temperature dependancy of the equilibrium constant for the water-gas reaction and subsequent curve fitting equations are given in figure 4.2.

Solution to the five simultaneous non-linear equations given above was obtained by employing a Newton-Raphson iteration scheme. The use of this scheme is common for solving systems of non-linear equations. A detailed description of this technique can be found in Appendix B.

After determining the composition of the hot raw producer gas in the oxidation/pyrolysis section, the final step in the gasifier simulation was to model the drying section. This was done by again performing an energy
Figure 4.2 Temperature dependency of equilibrium constant for the water-gas reaction.
balance around the drying section. As shown in figure 4.1, inputs to the control volume are the inlet wet wood and the raw producer gas at the pyrolysis temperature determined earlier. Exiting the drying section are the dried wood and the cooled raw producer gas. One mole of inlet wood was again used as the basis so that a continuity of mass flow would be maintained through the drying and oxidation/pyrolysis sections. Formulation of the energy equation for the drying section was similar to a phase change heat exchanger. The total energy required to dry the wood is a sum of the energy required to raise the wood and water to the saturation temperature, and the latent heat of evaporation of the water at that temperature. The saturation temperature is a function of the operating pressure of the gasifier. Delineating the individual energy requirements yielded the following analysis:

heating wood and water to evap. temperature,

\[ Q_{\text{wood}} = (1 \text{ mole})(C_{\text{p,wood}})(T_{\text{sat}} - T_{\text{amb}}) \]
\[ Q_{\text{water}} = (n_{\text{water}})(C_{\text{p,water}})(T_{\text{sat}} - T_{\text{amb}}) \]
evaporating water contained in wood and heating to gasifier exit temperature,

\[ Q_{\text{evap}} = (n_{\text{water}}) \Delta H_{fg} - \int_{T_{\text{sat}}}^{T_{\text{exit}}} c_{H_2O} dT \]

where

- \( T_{\text{sat}} \) = saturation temperature of water at gasifier operating pressure
- \( n_{\text{water}} \) = moles of water contained in wet wood entering gasifier
- \( \Delta H_{fg} \) = molar latent heat of vaporization of water at saturation temperature
- \( T_{\text{exit}} \) = temperature of raw producer gas exiting gasifier

The above analysis assumes a free behavior of the water contained in the wood. The total energy required to dry the wood is:

\[ Q_{\text{total}} = Q_{\text{wood}} + Q_{\text{water}} + Q_{\text{evap}} \]

It is this amount of energy that must be obtained from the hot raw producer gas filtering up through the wet wood.
Summing the energy demands for both wood and gas flows yields the overall energy balance:

\[ Q_{\text{total}} = (c) \int_{T_{\text{exit}}}^{T_{\text{pyr}}} C_{P_{H_2}} \, dT + (d) \int_{T_{\text{exit}}}^{T_{\text{pyr}}} C_{P_{H_2}O} \, dT + (e) \int_{T_{\text{exit}}}^{T_{\text{pyr}}} C_{PCO} \, dT + (f) \int_{T_{\text{exit}}}^{T_{\text{pyr}}} C_{P_{CO_2}} \, dT + (g) \int_{T_{\text{exit}}}^{T_{\text{pyr}}} C_{P_{CH_4}} \, dT + (\Phi \times 1.055)(3.76) \int_{T_{\text{exit}}}^{T_{\text{pyr}}} C_{P_{N_2}} \, dT \]

The final step in the drying section was to add the water vapor evaporated from the wood to the raw producer gas. With this done, the gasifier model was completed. The significant result obtained from this model is the raw producer gas composition and temperature exiting the gasifier.

Heat Exchanger

The purpose of incorporating a heat exchanger in the sub-system design is to reclaim any sensible energy contained in the raw producer gas exiting the gasifier. This energy is usually lost in the gas cleaning process as the gas is cooled and cleaned.

The inlet flows to the heat exchanger consist of the hot, raw producer gas from the gasifier and the cool, clean producer gas from either the sub-system compressor.
(sub-system A) or the gas cleaning equipment (sub-systems B & C). The mass flow rate of the clean gas to be heated will always be less than the flow rate of the raw gas because of the removal of water vapor and tars in the gas cleaning equipment. The composition of the gas flows are identical with the exception of the water vapor present in the raw gas. Because of these conditions, it is theoretically possible to design a heat exchanger of infinite size which could yield hot clean producer gas at the same temperature as the raw producer gas exiting the gasifier.

Consideration of the above information led to the development of the simplified heat exchanger model. This model initially compares the temperature of the raw gas exiting the gasifier to the cool, clean gas exiting either the gas compressor (sub-system A) or the gas cleaning equipment (sub-systems B & C). If the raw gas temperature is greater than the cool clean gas, then the heat exchanger is activated and the cool gas is increased to within the allowable terminal temperature difference of the hot raw gas. For the heat exchanger used in this model, a 10 degrees F. terminal temperature difference is specified. If the temperature of the raw gas does not exceed the temperature of the clean gas, then the clean gas stream by-passes the heat exchanger and enters the trimburner.
Gas Cleaning Equipment

As previously mentioned, a detailed model of the gas cleaning equipment is not the objective of this study. Rather, the need for such equipment is defined and some basic considerations made. In modelling the sub-system designs, however, it is necessary to determine the effect that this equipment will have on the temperature and composition of the gas flowing through it. The requirement on the gas cleaning model, then, is to alter the gas stream to the condition it would be if it had passed through such equipment.

The objectives of the gas cleaning equipment, as stated earlier, are to remove all tar and water vapor from the gas flow. This is usually done by cooling the gas to nearly ambient conditions and allowing condensation of the various particulates and water to occur. The clean gas exiting the cleaning equipment then is nearly void of water vapor and very nearly at ambient temperature. The simplified model developed simulates the existence of this equipment by reducing the gas stream temperature to 298 K and setting the mass of water in the stream equal to zero.

Compressor

The purpose of the gas compressor model was to simulate the compression of a composite mixture of gases.
The desired output from such a simulation is the power required to operate the compressor, and the average exit temperature of the composite gas.

Gas compression in the sub-systems is assumed to be a reversible polytropic process. In addition, the gas is considered to be ideal so that independent behavior of the gaseous components can be assumed. For a polytropic process:

\[ P_1 V_1^n = P_2 V_2^n \]

and

\[ \frac{T_2}{T_1} = \left( \frac{P_2}{P_1} \right)^{\frac{n-1}{n}} \]

where

- \text{STATE 1} = inlet pressure, volume, and temperature to the compressor
- \text{STATE 2} = exit pressure, volume, and temperature from the compressor
- \( n \) = a constant (equal to \( k = \frac{C_p}{C_v} \) for an isentropic process)

This equation yields the exit temperatures of the individual components of the gas if the compression ratio and inlet temperatures are known. Because isentropic compression is assumed, \( n \) is replaced by the ratio of specific heats.
The isentropic work required to compress each gaseous component in the gas stream can be determined from:

$$W_2 = \int_1^2 P \, dV$$

For the isentropic process the work rate is:

$$\dot{W_R}_2 = \dot{m}R(T_2 - T_1)/\gamma - k)$$

where

- $\dot{W_R}_2$ = work required by compressor for gaseous component $i$
- $R$ = gas constant for gaseous component $i$
- $\dot{m}$ = mass flow rate of gaseous component $i$

The total power required by the compressor is calculated by summing the power required by each gas component in the stream.

$$WR_{\text{isent}} = \text{Power} = \sum_i \dot{W_R}_2 \text{ component } i$$

In conforming with previous work done on the biomass plant, an 82% compressor efficiency and 90% motor
efficiency is assumed. Therefore, the total power required to compress the gas is:

\[ \text{Power} = W_{R_{\text{sen}}} / (0.82)(0.90) \]

The determination of the average stream temperature of the gas exiting the compressor is required for subsequent calculations in either the heat exchanger (sub-system A) or the gasifier (sub-system B). This temperature is calculated by using an energy-weighted averaging scheme. For the gaseous stream exiting the compressor:

\[ n_T C_{p_{\text{mean}}}(T_{\text{mean}} - T_{\text{ref}}) = \sum n_i C_{p_i}(T_{Z_i} - T_{\text{ref}}) \]

where

- \( n_T \) = total number of moles in compressor exit stream
- \( C_{p_{\text{mean}}} \) = mean specific heat of compressor exit stream
- \( T_{\text{mean}} \) = average temperature of compressor exit stream
- \( T_{\text{ref}} \) = reference temperature used to define enthalpy scale
- \( n_i \) = number of moles of gaseous component \( i \) in compressor exit stream
- \( C_{p_i} \) = specific heat of gaseous component \( i \) (assumed constant over temperature range)
$T_{zi} =$compressor exit temperature of gaseous component $i$

Solving for $T_{mean}$ yields:

$$T_{mean} = \left[ \sum_i C_{pi}(T_{exiti} - T_{ref})/C_{p_{mean}} \right] + T_{ref}$$

Values pertinent to the compressor analysis are given in Table 4.2 for the gas constituents contained in the producer gas stream.

**TABLE 4.2 PROPERTIES OF GASES FLOWING THROUGH SUB-SYSTEM COMPRESSOR (Van Wylen, Sonntag, 1978)**

<table>
<thead>
<tr>
<th>GAS CONSTANT (R) (ft-lbf/lbm-R)</th>
<th>*SPECIFIC HEAT (Cp) (Btu/lbm-R)</th>
<th>*SPECIFIC HEAT RATIO (k)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Oxygen</td>
<td>48.28</td>
<td>.219</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>55.15</td>
<td>.248</td>
</tr>
<tr>
<td>Methane</td>
<td>96.35</td>
<td>.532</td>
</tr>
<tr>
<td>Carbon Monoxide</td>
<td>55.16</td>
<td>.249</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>35.10</td>
<td>.203</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>766.40</td>
<td>3.430</td>
</tr>
</tbody>
</table>

* ( $Cp$ and $k$ are evaluated at 80 degrees F. )

Trimburner

The trimburner used in the sub-system design is
modelled as a duct burner which combusts the clean producer gas directly in the pressurized air stream. The known inputs for the trimburner are the air stream mass flow rate, inlet temperature, and desired exit temperature. In addition, the composition and temperature of the clean producer gas to be burned in the trimburner is also known. The output of the trimburner model include the mass flow rate of the producer gas required to boost the air stream temperature and the trimburner exit conditions.

Because the action in the trimburner is a combustion process, all assumptions previously defined for combustion in the main combustor model were adopted here. These assumptions define a 2% radiative heat loss and a 99% carbon conversion efficiency for the combustion process.

To model the trimburner, a control volume was drawn around the component to isolate the inlet and exit flows. The control volume and resulting combustion reaction used for the trimburner are shown in figure 4.3. An energy balance for the trimburner was conducted to yield:

$$\sum_{j} \hat{n}_{I_j}(h_{f_j}) + \int_{T_{ref}}^{T_x} \dot{c}_{p_j} dT + \sum_{j} \hat{n}_{\Pi_j}(h_{f_j}) + \int_{T_{ref}}^{T_x} \dot{c}_{p_j} dT$$

$$= \sum_{j} \hat{n}_{\Pi_j}(h_{f_j}) - \int_{T_{ref}}^{T_x} \dot{c}_{p_j} dT$$
INLET PRODUCER GAS
(H₂, CO, CO₂, CH₄, N₂)

TRIMBURNER

STOICHIOMETRIC COMBUSTION

EXIT STREAM (TO GAS TURBINE)
(CO₂, H₂O, O₂, N₂, CO)

\[ (x)H₂ + (x)CO + (x)CO₂ + (x)CH₄ + \left(\text{mass}_{\text{H₂O}} \times 2\right)O₂ + \left[\text{mass}_{\text{N₂}} \times 0.79\right]N₂ \]

\[ \rightarrow [(x)+(x)+(x)]CO₂ + [(x)+(2)+(x)]H₂O + \left[\text{mass}_{\text{H₂O}} \times 2\right] - (1/2)(x)+(1/2)(x) - 2(x)O₂ \]

\[ + \left[\text{mass}_{\text{N₂}} \times 0.76\right]N₂ \]

Figure 4.3 Flow model used to simulate operation of trimburner.
where

\[ n_{ij} = \text{molar rate of gas component } j \text{ in stream I} \]
\[ T_1 = \text{temperature of stream I} \]
\[ n_{II,j} = \text{molar rate of gas component } j \text{ in stream II} \]
\[ T_2 = \text{temperature of stream II} \]
\[ n_{III,j} = \text{molar rate of gas component } j \text{ in stream III} \]
\[ T_3 = \text{temperature of stream III} \]

expanding the summation of streams I, II, and III gives:

\[
\begin{align*}
&= \left( x(c) \int_{T_{ref}}^{T_1} C_{p,H_2} \, dT + (x)(e) \int_{T_{ref}}^{T_2} C_{p,CO} \, dT \right) + (x)(f) \int_{T_{ref}}^{T_3} C_{p,CO_2} \, dT \\
&\quad - (x)(g) \int_{T_{ref}}^{T_1} C_{p,CH_4} \, dT - (x)(h) \int_{T_{ref}}^{T_2} C_{p,CH_4} \, dT - (x)(i) \int_{T_{ref}}^{T_3} C_{p,CH_4} \, dT \\
&\quad + \left( m_{w_{air}} X.79 \right) \int_{T_{ref}}^{T_1} C_{p,H_2} \, dT - (m_{w_{air}} X.79) \int_{T_{ref}}^{T_2} C_{p,CO} \, dT \\
&\quad - (m_{w_{air}} X.79) \int_{T_{ref}}^{T_3} C_{p,CO_2} \, dT \\
&\quad + \left( m_{w_{air}} X.79 \right) \int_{T_{ref}}^{T_1} C_{p,CH_4} \, dT - (m_{w_{air}} X.79) \int_{T_{ref}}^{T_2} C_{p,CH_4} \, dT \\
&\quad + (m_{w_{air}} X.79) \int_{T_{ref}}^{T_3} C_{p,CH_4} \, dT \\
&\quad + \left( m_{w_{air}} X.79 \right) \int_{T_{ref}}^{T_1} C_{p,CO_2} \, dT + (m_{w_{air}} X.79) \int_{T_{ref}}^{T_2} C_{p,CO_2} \, dT + (m_{w_{air}} X.79) \int_{T_{ref}}^{T_3} C_{p,CO_2} \, dT \\
&= \left[ \left( x(c) + (x)(e) + (x)(g) \right) \int_{T_{ref}}^{T_1} C_{p,CO} \, dT \right] + \left[ \left( x(f) + (2)(x)(g) \right) \int_{T_{ref}}^{T_3} C_{p,CO_2} \, dT \right] \\
&\quad + \left[ \left( m_{w_{air}} X.79 \right) - (5)(x)(e) - (5)(x)(g) \right] \int_{T_{ref}}^{T_1} C_{p,CH_4} \, dT \\
&\quad - \left[ \left( m_{w_{air}} X.79 \right) - (2)(x)(g) \right] \int_{T_{ref}}^{T_2} C_{p,CH_4} \, dT \\
&\quad + \left( m_{w_{air}} X.79 \right) \int_{T_{ref}}^{T_3} C_{p,CH_4} \, dT \\
&\quad + \left( m_{w_{air}} X.79 \right) \int_{T_{ref}}^{T_1} C_{p,CO_2} \, dT \right]
\end{align*}
\]

As shown above and in figure 4.3, the variable X represents the molar rate at which the producer gas must be supplied to obtain the necessary heating in the trimburner. It is this molar rate of the producer gas (X)
that is desired in the solution of the energy balance. The molar rate is obtained readily by solving the above equation for the variable $X$. Since the composition of the producer gas was calculated from the basis of 1 mole of dry wood to the gasifier, determining the value of $X$ specifies the mass rate of wet and dry wood to the gasifier as well.

$$
\dot{m}_{\text{dry wood}} = (x)(22.84) \\
\dot{m}_{\text{wet wood}} = (x)(22.84)/(1.0 - \text{M.C.})
$$

where

- $\dot{m}_{\text{dry wood}}$ = mass rate of dry wood into gasifier
- $\dot{m}_{\text{wet wood}}$ = mass rate of wet wood into gasifier
- M.C. = wood moisture content (wet basis)

The mass flow rate of air and combustion products exiting the trimburner can be obtained from a mass balance consideration of the control volume in figure 4.3.

$$
\sum \dot{m}_i + \sum \dot{m}_x = \sum \dot{m}_m
$$

With the molar supply rate of producer gas ($X$) known, the composition of the stream exiting the trimburner can be
calculated from the combustion reaction given in figure 4.3.

OVERALL SUB-SYSTEM ANALYSIS

The three gasifier/trimburner sub-system designs previously described were constructed from the component models developed in the preceding section. To determine the performance of these sub-systems, it is necessary to evaluate certain parameters which pertain to the entire sub-system. Comparisons of these parameters can then be made to determine the most advantageous sub-system for a given set of operating conditions. These overall sub-system parameters include:

(1) first law thermodynamic efficiency.
(2) second law thermodynamic efficiency.
(3) irreversibility.

This section presents the derivation of these parameters based on the three gasifier/trimburner sub-system designs.

First Law Efficiency

To derive an expression for any type of system parameter a control volume defining the system boundary is required. In working with the gasifier/trimburner sub-systems, control volumes as shown in figure 4.4 were used.
Figure 4.4 Control volumes specified in overall sub-system analysis.
System efficiency based on the first law of thermodynamics is generally defined as:

\[ \eta = \left( \frac{\text{energy sought}}{\text{energy that costs}} \right) \times 100 \]

Applying this definition to the sub-system control volumes in figure 4.4 yields:

\[ \eta = \frac{(m_{\text{trim \ out}})(h_{\text{trim \ out}}) - (m_{\text{trim \ in}})(h_{\text{trim \ in}})}{(m_{\text{wood}})(HHV_{\text{wood}}) + WR_{\text{comp}}} \]

where

- \( h_{\text{trim \ out}} \) = enthalpy of gas stream exiting trimburner
- \( h_{\text{trim \ in}} \) = enthalpy of compressed air stream entering trimburner
- \( m_{\text{trim \ out}} \) = mass flow rate of gases exiting trimburner
- \( m_{\text{trim \ in}} \) = mass flow rate of compressed air entering trimburner
- \( HHV_{\text{wood}} \) = higher heating value of dry wood entering gasifier
- \( m_{\text{wood}} \) = mass flow rate of dry wood entering gasifier
- \( WR_{\text{comp}} \) = work rate of the producer gas compressor (sub-system A), gasifier compressor (sub-system B), or 0 (sub-system C)
Second Law Efficiency

The second law efficiency, or effectiveness, is defined by [Moran, 1982] as:

\[ \epsilon = \frac{\text{Availability out in product}}{\text{Availability in}} \]

Availability is defined as the potential of a substance to perform work in a given thermodynamic environment. For the analysis of the sub-systems, only the flow availability need be considered for the gaseous streams. In terms of thermodynamic properties, flow availability is expressed as:

\[ A_f = (h - h_0) - T_0(s - s_0) \]

where

- \( A_f \) = flow availability
- \( h, s \) = enthalpy and entropy of substance considered
- \( T_0 \) = dead state temperature (usually environment temperature)
- \( h_0, s_0 \) = enthalpy and entropy of substance at dead state conditions

In the gasifier/trimburner designs, the availability out in the product shall be defined as the increase in
availability that occurs as the compressed air stream passes through the trimburner. Therefore:

\[
\text{Availability out in product} = A_{f_{\text{trim out}}} - A_{f_{\text{trim in}}}
\]
\[
= m_{\text{out}} \left[ (h_{\text{trim out}} - h_0) - T_0 (s_{\text{trim out}} - s_0) \right]
\]
\[
- m_{\text{in}} \left[ (h_{\text{trim in}} - h_0) - T_0 (s_{\text{trim in}} - s_0) \right]
\]

Simplifying the above equation yields:

\[
A_p = (m_{\text{out}})(h_{\text{out}}) - (m_{\text{in}})(h_{\text{in}}) - T_0 \left[ (m_{\text{out}})(s_{\text{out}}) - (m_{\text{in}})(s_{\text{in}}) \right]
\]

where

- \( A_p \) = availability out in product
- \( m_{\text{out}} \) = mass flow rate of gas stream exiting trimburner
- \( h_{\text{out}} \) = enthalpy of gas stream exiting trimburner
- \( s_{\text{out}} \) = entropy of gas stream exiting trimburner
- \( m_{\text{in}} \) = mass flow rate of compressed air entering trimburner
- \( h_{\text{in}} \) = enthalpy of compressed air entering trimburner
- \( s_{\text{in}} \) = entropy of compressed air entering trimburner

The total availability into the gasifier/trimburner sub-systems is the sum of all the availability entering across the control volume boundary. This includes:

(1) the chemical availability of the dry wood fuel entering the gasifier (assumed equal to the higher heating value of the dry wood).
(2) the availability of the electrical power required to operate the producer gas compressor (sub-system A) or the gasifier compressor (sub-system B). The availability here is equal to the electrical power.

(3) the availability of the air entering the gasifier.

(4) the availability of the water contained in the wood entering the gasifier.

With the inlet availability streams identified, the second law efficiency can be derived:

$$
\eta = \frac{(m_{out}X_{out}) - (m_{in}X_{in}) - T_0 [(m_{out})(s_{out}) - (m_{in})(s_{in})]}{(m_{wood})(HHV_{wood}) + WR_{comp} + A_{gasifier} + A_{water}}
$$

where

- $A_{gasifier}$ = availability of air entering gasifier
- $A_{water}$ = availability of water contained in wood entering gasifier

Irreversibility

Availability, unlike energy, is not conserved. Availability is destroyed by irreversible processes which occur in the system. This destruction of availability is called irreversibility and is proportional to the production of entropy in a given system. The
irreversibility of a system can be determined by applying the availability rate equation for a control volume:

$$\frac{dA_{cv}}{dt} = \int_S (1 - \frac{T_{c1}}{T_a}) q_s \, dS - (\dot{W}_{cx} - p \frac{dV}{dt}) + \sum \dot{m}_i A_{fi} - \sum \dot{m}_e A_{fe} = \dot{I}$$

where

$$\int_S (1 - \frac{T_{c1}}{T_a}) q_s \, dS = \text{rate of availability which occurs due to heat transfer across the control surface (designated by S)}$$

$$(\dot{W}_{cx} - p \frac{dV}{dt}) = \text{work that crosses the control volume boundary or is done by control volume expansion or contraction}$$

$$\sum \dot{m}_i A_{fi} = \text{sum of flow availability rate entering control volume from inlet streams}$$

$$\sum \dot{m}_e A_{fe} = \text{sum of flow availability rates exiting the control volume from exit streams}$$

$$\dot{I} = \text{irreversibility rate for control volume}$$

For the steady-state control volumes shown in figure 4.4, the availability rate equation reduces to:

$$\dot{I} = -\dot{W}_{cx} + \sum \dot{m}_i A_{fi} - \sum \dot{m}_e A_{fe}$$

ACCURACY OF SUB-SYSTEM MODEL

In an effort to determine the accuracy of the output obtained from sub-system models, comparisons with commercially available components were sought. The only component in the sub-system models which required a close investigation was the up-draft gasifier because of the
modelling assumptions made. The other components in the sub-system are modelled from well defined thermodynamic concepts and will, therefore, yield predictable results. This section evaluates the validity of the up-draft gasifier model through comparision with both commercially available units and previously developed gasifier models.

To evaluate the accuracy of the sub-system gasifier model, information from an ongoing study by ENFOR [1983] was utilized. In phase I of the ENFOR study, several up-draft gasifier manufacturers were surveyed concerning the performance of their products. The information received on the up-draft units was then averaged and compiled in tabular form to represent the expected performance of a typical up-draft gasifier. Runs of the atmospheric gasifier model in sub-system A and the pressurized unit in sub-systems B and C were then made at the same operating conditions as the data obtained by ENFOR. The output from these runs were averaged and compared with the results contained in the ENFOR study. The results of this comparision can be found in table 4.3. As table 4.3 indicates, the sub-system gasifier models are in fair agreement with the data obtained from the commercial gasifiers. The values for the sub-system gasifier models tend to lie on the upper limit of the manufacturers data. This can be understood by realizing that the sub-system gasifier model represents an idealized case in which all
Table 4.3 GASIFIER PERFORMANCE DATA FROM ENFOR SURVEY AND SUB-SYSTEM GASIFIER MODEL

<table>
<thead>
<tr>
<th>PARAMETER</th>
<th>ENFOR DATA SURVEY</th>
<th>SUB-SYSTEM GASIFIER MODEL</th>
</tr>
</thead>
<tbody>
<tr>
<td>Moisture Content (wet basis)</td>
<td>50 %</td>
<td>50% (specified)</td>
</tr>
<tr>
<td>Equivalence Ratio</td>
<td>.30</td>
<td>.30 (specified)</td>
</tr>
<tr>
<td>Producer Gas Exit Temperature (F)</td>
<td>167 - 248</td>
<td>241</td>
</tr>
<tr>
<td>Dry Producer Gas Composition (vol %)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Nitrogen</td>
<td>53</td>
<td>47.8</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>9</td>
<td>8.0</td>
</tr>
<tr>
<td>Carbon Monoxide</td>
<td>24.5</td>
<td>27.8</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>11</td>
<td>12.4</td>
</tr>
<tr>
<td>Methane</td>
<td>2.5</td>
<td>4.0</td>
</tr>
<tr>
<td>Higher Heating Value of Dry Producer Gas</td>
<td>121 - 161</td>
<td>161.8</td>
</tr>
<tr>
<td>(Btu/cubic ft.)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Conversion Efficiency (%)</td>
<td>70-90</td>
<td>88.0</td>
</tr>
</tbody>
</table>
the carbon in the wood is converted and no thermal losses are assumed. Because of this, the mole fractions of the desirable gas products CO, H₂, and CH₄ tend to be slightly higher as shown in table 4.3. Consequently, the higher heating value of the producer gas and the conversion efficiency are also higher for the sub-system model.

Although the comparison between the sub-system gasifier models and the manufacturers data is useful, it does not provide information about the accuracy of the models at varying operating conditions. Because manufacturers typically operate their products at conditions which optimize producer gas production (equivalence ratio = .3), obtaining off-design performance data is difficult. To check the accuracy of the sub-system gasifier model at various equivalence ratios, it was necessary to use previous gasification research studies for comparison. In a study by Desrosiers [1977], an exhaustive model was developed to simulate the performance of an atmospheric pressure, up-draft gasifier. By comparing the results of that study to computer runs of the atmospheric pressure gasifier modelled in sub-system A, performance trends over a range of equivalence ratios can be evaluated. It is important to note that this comparison is between two theoretical gasifier models, and neither model is totally predictive of an actual
gasification unit in operation. However, comparisons of this nature are useful in that the accuracy of performance trends can be reinforced if good agreement between the two models exist over the same operating range.

The key parameters to analyze when comparing the two models over the range of equivalence ratios are the composition and pyrolysis temperature. These parameters ultimately determine all the performance characteristics of a gasification system. Figure 4.5 shows the gasifier pyrolysis temperature as a function of equivalence ratio for both Desrosiers model and the sub-system A model. The correlation here is quite good between the two models over the entire range of equivalence ratio.

Figure 4.6 displays the mole fraction of each component in the dry producer gas as a function of equivalence ratio for both models. The trend of each gas as the equivalence ratio increases is similar for both Desrosiers model and the sub-system A model. Desrosiers model tends to predict slightly higher amounts of CO and H₂ in the producer gas, while the sub-system A model predicts a slightly higher mole fraction of CH₄. Like the sub-system gasifier model, Desrosiers model assumes complete conversion of carbon (above equivalence ratio=.275) and no thermal losses. It is expected, therefore, that the values of CO and H₂ in both models are slightly higher than in actual up-draft gasifiers. The
Figure 4.5 Comparision of gasifier pyrolysis temperature between Desrosiers model and sub-system A model over range of equivalence ratios.
Figure 4.6 Comparision of dry producer gas composition between Desrosiers model and sub-system A model over range of equivalence ratios.
approximately 5% difference in mole fraction of CO and H₂ between the two models may be accountable by the fact that Desrosiers assumed that all the gasification reactions listed in table 2.1 proceeded to equilibrium conditions.

In evaluating the accuracy of the sub-system gasifier models, it was determined that the results obtained are somewhat optimistic because of the idealized assumptions made in modelling. The sub-system gasifier results are, however, within the upper limits of the typical values expected in the study by ENFOR. Therefore, it appears that the sub-system gasifier model is accurate. In addition, comparison of the sub-system A gasifier model with a gasifier model developed by Desrosiers [1977] indicate that the sub-system model exhibits expected trends in composition and temperature variation as the equivalence ratio is changed.
Chapter 5

RESULTS OF SUB-SYSTEMS SIMULATIONS

The sub-system models developed in the previous chapter were integrated into a main biomass power plant computer program developed in previous work [Brynjolfsson, 1983; Bauer, 1984; Fox, 1984]. The integrated program was used to analyze the performance of each sub-system design and evaluate the overall biomass plant when using the three gasifier/trimburner sub-systems. The results and evaluations obtained from this analysis are presented in this chapter. A sample of the output from the integrated program as well as the fortran source codes for each sub-system can be found in Appendix C.

RESULTS

The presentation of the simulation results is separated into several sections pertaining to each of the critical parameters which were investigated.

Pyrolysis Temperature

The temperature at which pyrolysis occurs in the gasifier is an important parameter in determining the overall performance of the sub-systems. As discussed in chapter 4, the pyrolysis temperature determines the equilibrium constant in the chemical equilibrium equation
(iteration equation #1) as well as the enthalpies used in the heat balance equation (iteration equation #5). It is these equations which determine the composition of the producer gas formed.

The reason the pyrolysis temperature is the only parameter which influences the gas composition can be shown by considering the water-gas dissociation equation:

\[
\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2
\]

Since the water contained in the inlet wood is assumed to be removed in the drying section, the wood moisture content will have no effect on this dissociation reaction. In addition, the number of moles of reactants equals the number of moles of products in this reaction. This eliminates any pressure dependancy and temperature is left as the only variable of concern. Thus, the importance of the gasifier pyrolysis temperature is apparent.

Figure 5.1 shows the pyrolysis temperature of each sub-system as a function of equivalence ratio. The pyrolysis temperature is independant of the wood moisture content because of the gasifier modelling scheme used. The curves shown in figure 5.1, therefore, apply to any inlet wood moisture content. From figure 5.1 it is apparent that the pyrolysis temperature curves of sub-systems B and C are virtually the same. In addition, the
Figure 5.1 Pyrolysis temperature for each sub-system as a function of equivalence ratio.

Figure 5.2 Dry producer gas composition as a function of equivalence ratio for sub-system A.
sub-system A pyrolysis temperature curve is consistently about 300 F less than that of the sub-systems B and C curves over the range of equivalence ratio. These two tendencies are due to the condition of the inlet air to the gasifiers. In sub-system A (atmospheric gasifier), ambient air at 60 F is used in the gasifier. In sub-systems B and C (pressurized gasifier) the gasifier inlet air has been compressed to approximately 9 atmospheres. This compression results in a temperature increase in the gasifier inlet air for these sub-systems. The temperature of the air entering the gasifier in sub-systems B and C is about 550 F. The higher temperature of the gasifier inlet air in these sub-systems result in hotter combustion products in the gasifier combustion zone, and ultimately, higher pyrolysis temperatures as figure 5.1 indicates.

Producer Gas Composition

The dry producer gas composition as a function of equivalence ratio is shown in figures 5.2, 5.3, and 5.4 for sub-systems A, B and C respectively. The gas composition is independant of moisture content and is a function of pyrolysis temperature only. The compositions in figures 5.2, 5.3, and 5.4 therefore, apply to any moisture content of the inlet wood. From these figures it can be seen that again, sub-systems B and C have identical gas compositions over the entire range of equivalence
Figure 5.3  Dry producer gas composition as a function of equivalence ratio for sub-system B.

Figure 5.4  Dry producer gas composition as a function of equivalence ratio for sub-system C.
ratio. Sub-system A, however, has a noticeably different gas composition over the same operating range. This trend is directly a result of the difference in pyrolysis temperatures for the sub-systems. The dry producer gas compositions for sub-systems B and C tend to be the same because the pyrolysis temperatures of these two cases are equal. The lower pyrolysis temperature in sub-system A enhances the production of hydrogen and carbon monoxide in the gas yield. This effect is a result of the water-gas dissociation equation which favors hydrogen production at lower temperatures. The slight increase in carbon monoxide production in sub-system A may be attributable to a lower temperature in the combustion zone of the gasifier.

Figures 5.2, 5.3, and 5.4 also indicate that the producer gas yield from sub-system A maintains a higher percent of combustible gases over the entire range of equivalence ratio than the other sub-systems. The mole fractions of the main combustibles, CO and H₂, decrease rapidly in sub-systems B and C until they vanish altogether at an equivalence ratio of .7. In contrast, gas yields from sub-system A contain appreciable amounts of CO and H₂ even at equivalence ratios as high as .95. This effect is again most likely a result of the elevated pyrolysis temperature of sub-systems B and C which tends
to drive the process more toward combustion than gasification.

The effect of higher pyrolysis temperature on methane production can be seen in figures 5.2, 5.3, and 5.4. Although the mole fraction of CH₄ is small over the operating range for all three sub-systems, a definite trend is apparent. Sub-systems B and C, with the higher pyrolysis temperature, yield higher amounts of CH₄ than does sub-system A. This trend is consistent with trends noticed in methanol production plants where elevated temperatures and pressures are used to promote higher CH₄ yields from biomass distillation processes for conversion to methanol.

From figures 5.1 through 5.4 it is evident that the performance of the gasifiers in sub-systems B and C are very similar. For this reason, all subsequent analysis pertaining to gasifier output will be restricted to the investigation of sub-systems A and B. It will therefore be implied that the performance of the gasifier in sub-system C is essentially the same as that of the gasifier in sub-system B in all discussions that follow.

Gasifier Exit Temperature

As discussed in chapter 3, the sub-systems have been designed to utilize the sensible energy contained in the producer gas stream exiting the gasifier. A good measure
of the amount of sensible energy contained in the gasifier exit stream is the producer gas exit temperature. For this reason, the gas exit temperature as a function of equivalence ratio has been plotted and is shown in figures 5.5 and 5.6 for sub-systems A and B respectively.

As figures 5.5 and 5.6 indicate, the temperature of the gas exiting the gasifier decreases for both sub-systems as the moisture content increases. This trend is due to the decrease in producer gas temperature which must occur as the demand for wood drying in the gasifier drying section increase. As the moisture content increases, more energy is removed from the rising producer gas to accommodate the necessary water evaporation. In fact, the operational limits of the gasifier are defined by the gasifier exit temperature. The temperature of the exiting gas must be at least equal to the saturation temperature of the water at the gasifier operating pressure for total drying to occur. If the exit temperature is below this limit, then total drying of the inlet wood is impossible and the gasifier will fail to operate.

This effect can be seen in figures 5.5 and 5.6. The atmospheric gasifier exit temperature must be higher than 212 F. As shown in the figure, equivalence ratios of less than approximately .31 would result in exit temperatures below this limit and the gasifier would fail. The
Figure 5.5 Temperature of producer gas at gasifier exit for sub-system A as a function of equiv. ratio at four levels of wood moisture content.

Figure 5.6 Temperature of producer gas at gasifier exit for sub-system B as a function of equivalence ratio at four levels of wood moisture content.
pressurized gasifier of sub-system B (figure 5.6), reaches this operational limit at a lower equivalence ratio of approximately .29. This is due solely to the fact that the higher inlet air temperature of the pressurized gasifier yields higher gas temperatures and, therefore, a greater potential for wood drying.

Another point to note about the gasifier exit temperature is the increase exhibited as the equivalence ratio is increased. As figures 5.5 and 5.6 display, the producer gas exit temperature increases with equivalence ratio for all moisture contents. This result is attributed to the increased combustion that occurs in the gasifier as more air is fed into the unit. Higher equivalence ratios (for the range indicated) result in more of the wood being burned and higher temperature combustion products. These products in turn produce a higher temperature producer gas as shown in figures 5.5 and 5.6. This trend of increasing temperature is analogous to the increase in the adiabatic flame temperature of a combustion process as the amount of air is increased to the stoichiometric quantity (equivalence ratio=1.0).

The trend of slightly higher gas exit temperatures for sub-system B is seen to continue over the entire range of equivalence ratio. This is again due to the increased temperature of the air entering the gasifier. The effect
of this is that the producer gas exiting the gasifiers of sub-systems B and C will always be about 250 F hotter than the gas exiting the gasifier in sub-system A under the same conditions. This indicates that the sensible energy contained in the producer gas streams of sub-systems B and C will always be greater than the sensible energy contained in the sub-system A gas stream.

Producer Gas Stream Energy

From the previous discussions concerning the producer gas stream composition and exit temperature, it seems prudent to investigate the amount of energy contained in the producer gas stream exiting the gasifier. The total energy in the producer gas stream is defined as the sum of the chemical and sensible energy present. Chemical energy is defined as the energy released during combustion of the combustible gases (CO, H₂, and CH₄) present in the exit stream. This energy is equal to the higher heating value of each combustible gas multiplied by the amount of that gas present in the stream. Sensible energy is the energy that is present in the producer gas due to the high temperature of the gas stream. This energy is equal to the sum of the enthalpies of all the gas constituents at the gasifier exit temperature. In the sub-system designs considered, the sensible energy is recoverable by the use of the heat exchanger located at the gasifier exit.
Figures 5.7 and 5.8 show the chemical, sensible, and total energy as a function of equivalence ratio for sub-system A at moisture contents of .05 and .50 respectively. Figures 5.9 and 5.10 display the same values for sub-system B (and sub-system C) at the same operating conditions. The energy values have been calculated based on a per pound of wood gasified basis. This is done to alleviate the effect of increased wood flow rates on the results obtained.

The most obvious observation to make when viewing figures 5.7 through 5.10 is the relationship of chemical to sensible energy as the equivalence ratio is increased. The chemical energy is maximized at low equivalence ratios where the production of the combustible gases is greatest. As the equivalence ratio increases, the chemical energy steadily decreases. This is due to the decreased formation of combustibles and increased production of combustion products CO₂ and H₂O. The significance of this observation is that as the equivalence ratio increases, more of the wood is burned and less is gasified. The sensible energy on the other hand, is small at low equivalence ratios and steadily increases to a maximum value at the highest equivalence ratio. This is directly due to the trend noticed in figures 5.5 and 5.6 where the gas exit temperature increases with equivalence ratio.
Figure 5.7  Energy content of producer gas for sub-system A as a function of equivalence ratio with a wood moisture content of .05.

Figure 5.8  Energy content of producer gas for sub-system A as a function of equivalence ratio for a wood moisture content of .50.
Figure 5.9  Energy content of producer gas for sub-system B as a function of equivalence ratio for a wood moisture content of .05.

Figure 5.10  Energy content of producer gas for sub-system B as a function of equivalence ratio for a wood moisture content of .50.
Therefore, it appears that the plots of chemical and sensible energy in figures 5.7 through 5.10 actually represent another form of interpreting the information contained in the gas composition curves (chemical) and the gasifier exit temperature curves (sensible).

Another point to note is the nature of the total energy curve in figures 5.7 through 5.10. These curves are relatively flat, indicating a constant total energy content in the producer gas over the entire range of equivalence ratios. This result demonstrates that all of the energy entering the gasifier in the form of chemical energy in the wood has been accounted for in the gasifier products. This fact is reinforced by viewing the .05 moisture content curves of figures 5.7 and 5.9. Here, because of the low moisture content, very little energy loss occurs and the total energy in the producer gas is nearly equal to the heating value of the inlet wood (9000 Btu/lb).

The slight dip associated with the total energy curves of figures 5.7 through 5.10 is due to the nature of the gasifier modelling. In the pyrolysis section of the gasifier, some water vapor is formed during the wood conversion process. The amount of water formed increases from .07 moles at low equivalence ratios to .7 moles at the highest equivalence ratio considered. There is a latent heat of vaporization associated with this water
vapor formed which has not been included in the accounting of the chemical and sensible energy in the exiting products. As the equivalence ratio increases, the amount of latent energy will increase due to the formation of more water vapor. The result is that the total energy curve will dip slightly with increasing equivalence ratio due to the increase in the amount of latent energy. This energy, however, is not significant in the overall analysis since, at the most extreme point, the latent energy makes up only 3.7% of the total energy in the gas stream.

The effect of increasing moisture content in the inlet wood can also be seen in figures 5.7 through 5.10. Since the producer gas composition is independent of the moisture content, increasing the percent of water in the inlet wood will only affect the sensible energy. By comparing sub-system A at .05 moisture content (figure 5.7) with the same sub-system at .50 moisture content (figure 5.8), this trend becomes apparent. In both figures 5.7 and 5.8, the chemical energy curve remains constant (in the gasifier's operating range). The sensible energy in the .50 moisture content case, however, is less than it is in the .05 moisture content case. This decrease is due to the loss in sensible energy which must occur in the gasifier drying section to evaporate the
water in the wet inlet wood. The result is that the total energy curve in figure 5.8 is a consistent 1000 Btu less than the total energy curve in figure 5.7 over the same operating range. This amount of energy coincides very closely with the latent heat of vaporization for the increased amount of water in the wood for the two cases.

When comparing sub-system A and B by using figures 5.7 through 5.10, two points are most notable. First, because of the lower pyrolysis temperature, sub-system A has a higher chemical energy content than sub-system B over the entire range of equivalence ratios. This advantage, however, is offset by the lower sensible energy content of the gas in sub-system A. The second point to note is that increasing the moisture content of the inlet wood has a larger adverse effect on sub-system B than on sub-system A. This is primarily due to the fact that the energy in the producer gas of sub-system B is more reliant upon the sensible form of energy rather than the chemical form. Therefore, since drying the wet wood taxes only the sensible energy in the stream, sub-systems B and C will suffer the greatest from an increase in wood moisture content.

Sub-System Compressor Power

The power required to operate the sub-system compressors are shown in figures 5.11 and 5.12 as a
Figure 5.11 Sub-system A compressor power requirements as a function of equivalence ratio at four levels of wood moisture content.

Figure 5.12 Sub-system B compressor power requirements as a function of equivalence ratio at four levels of wood moisture content.
function of equivalence ratio. Figure 5.11 displays the sub-system A compressor power requirement over the range of equivalence ratios. Figure 5.12 displays the sub-system B compressor power requirement over the same range. Since there is no additional compressor used in sub-system C, no plot is presented.

From figures 5.11 and 5.12 it can be seen that the compressor power increases with both increasing wood moisture content and increasing equivalence ratio. In both cases the power increase is due to an increase in mass flow rate through the sub-system compressors. Increasing the wood moisture content increases the amount of wood that needs to be gasified to heat the air stream, thus increasing the producer gas flow rate. Increasing the equivalence ratio increases the wood and air flow rate to the sub-system which again, increases the flow rates through the compressors.

An important point to note about figures 5.11 and 5.12 is the large difference in compressor power required for sub-systems A and B. The compressor in sub-system A requires approximately 5 times more power to operate than the compressor in sub-system B. This large differential in power is the result of two major differences between the flows passing through each compressor. In sub-system A, the compressor must compress a higher flow rate than
the compressor in sub-system B. This is due to the fact that the sub-system A compressor is compressing producer gas (which is the sum of wood and air flow) whereas the sub-system B compressor is compressing only the gasifier inlet air stream. In addition, the average specific volume of the gases flowing through the compressor in sub-system A is greater than the specific volume of the air flowing through the sub-system B compressor. More energy is required to compress gases with high specific volumes. The result of these two differences is that much more power is required to compress the producer gas stream (sub-system A), as opposed to compressing the gasifier inlet air stream (sub-system B).

Overall Sub-System First Law Efficiency

The first law efficiency as defined in chapter 4 is shown as a function of equivalence ratio for all three sub-systems in figures 5.13 and 5.14. Figure 5.13 displays the efficiency curves for each sub-system at moisture contents of .05 and .30. Figure 5.14 displays the sub-systems efficiency curves at moisture contents of .15 and .50.

The effect of moisture content on efficiency can readily be seen in figures 5.13 and 5.14. As expected, the efficiency of all three sub-systems decreases with increasing moisture content. Increasing the moisture
Figure 5.13 First law Efficiency of all three sub-systems as a function of equivalence ratio at wood moisture contents of .05 and .30.

Figure 5.14 First law Efficiency of all three sub-systems as a function of equivalence ratio at wood moisture contents of .15 and .50.
content from .05 to .50 decreases the first law efficiency in all three sub-systems by about 10%. This decrease is a result of the sensible energy lost to drying the wet wood in the gasifier.

Investigation of figures 5.13 and 5.14 also reveal that the efficiency in all three sub-systems decrease with increasing equivalence ratio. Furthermore, this decrease is more extreme at higher moisture contents. This trend is due to the sensible energy losses which occur in the gas cleaning equipment. As the equivalence ratio increases, the mass flow rate differential between the raw producer gas containing water vapor and the dry clean producer gas increases. The wet raw gas exiting the gasifier passes through the heat exchanger and transfers energy to the cool dry gas flowing from the cleaning equipment. As the mass differential between these two streams increases, the temperature drop of the wet raw gas as it passes through the heat exchanger will decrease. This results in higher raw gas temperatures and flow rates entering the cleaning equipment as the equivalence ratio increases. The cleaning equipment then reduces the temperature and flow rate of the gas by cooling it down to 77 F and removing all the water present. This represents an energy loss to the system which will increase as the equivalence ratio increases. In addition, this effect
will be more dramatic at higher moisture contents because of the additional water vapor present in the gas stream which will be removed as the stream flows through the cleaning equipment.

With figures 5.13 and 5.14, it is possible to draw more concrete conclusions concerning the performance of the three sub-systems. From figures 5.13 and 5.14 it appears that sub-systems B and C have superior efficiencies at moisture contents less than .50 and equivalence ratios less than .8. At equivalence ratios above .8 the sub-system efficiency of sub-system A rises above that of sub-systems B and C. The amount of combustible gases in sub-systems B and C declines rapidly above equivalence ratios of .7. Since sub-system A contains a significant amount of combustible products even at high equivalence ratios, its efficiency will be higher than that of sub-systems B and C. At .50 moisture content all three sub-systems have essentially the same efficiency. Therefore, it appears that there is no real advantage to choosing one sub-system as opposed to another when basing the evaluation on first law efficiency.

Perhaps the most important observation to note concerning figures 5.13 and 5.14 is that the most efficient operation of all three sub-systems occurs at the lowest operable equivalence ratio. This leads to the
conclusion that the sub-system should be operated at low equivalence ratios to achieve optimum performance.

Overall Sub-System Second Law Efficiency

The second law efficiency as defined in chapter 4 is presented in figures 5.15 and 5.16 for each sub-system as a function of equivalence ratio. Again, the figures are divided to display the different moisture contents considered. Figure 5.15 shows the second law efficiency of the sub-systems at moisture contents of .05 and .30, whereas figure 5.16 shows the second law efficiency of the sub-systems at moisture contents of .15 and .50.

The trends observed in figures 5.15 and 5.16 are very similar to the trends observed for the first law efficiency. The decrease in second law efficiency with increasing moisture content is due to the same reasons as discussed for the first law efficiency. An increase in moisture content to the sub-systems increases the availability into the sub-system, but does not aid in the net availability change across the trimburner. More importantly, an increase in moisture content reduces the total energy in the producer gas stream. This requires additional wood to be gasified to accomplish the same heating in the trimburner. The result of these two effects is that the total availability into the sub-systems increases with moisture content while the
Figure 5.15 Second law efficiency of the sub-systems as a function of equivalence ratio at wood moisture contents of .05 and .30.

Figure 5.16 Second law efficiency of the sub-systems as a function of equivalence ratio at wood moisture contents of .15 and .50.
availability out in the products remains essentially unchanged. This will, therefore, reduce the second law efficiency.

A decrease in second law efficiency with increasing equivalence ratio is also observed in figures 5.15 and 5.16. This is due to the increase in availability into the sub-system which occurs with increasing equivalence ratio. As more air is introduced to the sub-system, more wood is consumed which means an increase in the inlet availability. Much of this increased inlet availability, however, is uselessly removed in the gas cleaning equipment. In addition, the work rate of the sub-system compressor increases with increasing equivalence ratio. This represents an additional increase in inlet availability. These trends all tend to decrease the second law efficiency due to the increase in inlet availability to achieve essentially the same net availability increase across the trimburner.

Overall Sub-System Irreversibility

The sub-system irreversibility, as discussed in chapter 4, is presented in figures 5.17 and 5.18 as a function of equivalence ratio. Figure 5.17 shows the sub-system irreversibility at moisture contents of .05 and .3. Figure 5.18 is a plot of the sub-system irreversibility at moisture contents of .15 and .50.
Figure 5.17 Irreversibility of the sub-systems as a function of equivalence ratio at wood moisture contents of .05 and .30.

Figure 5.18 Irreversibility of the sub-systems as a function of equivalence ratio at wood moisture contents of .15 and .50.
From figures 5.17 and 5.18 it can be seen that the irreversibility of the sub-systems increases with an increase in moisture content. This is due to the increased inlet mass wood rate which is needed to accomplish the required heating in the trimburner. The increase in inlet wood rate increases the total inlet availability more than the increase in total exit availability. This has the effect of increasing the sub-system irreversibility as defined in chapter 4.

Figures 5.17 and 5.18 also indicate that the sub-system irreversibility increases with increasing equivalence ratio. This is due to a combination of reasons. First, an increase in equivalence ratio increases the availability entering the sub-system via increased flow rates of gasifier inlet air. Also, an increase in equivalence ratio increases the mass rate of wood into the sub-system, thus increasing the inlet availability. Finally, the work rate of the product gas compressor increases with increasing equivalence ratios, increasing the irreversibility. The cumulative effect of these trends result in a net increase in the irreversibility of the sub-systems.

The increase in irreversibility with increasing moisture contents and equivalence ratios can also be explained from a practical point of view. As either the
moisture content or equivalence ratio increases, the mass rate of producer gas entering the trimburner also increases. This increases the amount of mixing which occurs in the trimburner. Since stream mixing is a highly irreversible process, the sub-system irreversibility should increase with the increased trimburner mixing.

This effect becomes more pronounced at higher equivalence ratios where there is little combustion occurring in the trimburner. At the higher equivalence ratios, the trimburner accomplishes the required air stream heating by mixing in high flow rates of very hot gases from the gasifier rather than by combustion. This increased mixing with high flow rates accounts again for increased sub-system irreversibility with increasing equivalence ratio.

Since sub-systems B and C rely more on the sensible energy mixing (especially at high equivalence ratios) to accomplish the required trimburner heating, the irreversibility due to mixing should be greater for these sub-systems. This is indeed the case as shown in figures 5.17 and 5.18 where the irreversibility of sub-systems B and C is higher than the irreversibility of sub-system A over the entire range of equivalence ratios. The dramatic increase in irreversibility of sub-systems B and C at the higher equivalence ratios is due in part to the high gas
flow rates required to provide the trimburner heating via stream mixing.

Sub-system A has a less pronounced increase in the irreversibility at the higher equivalence ratios because of the lower producer gas stream flow rates. Because the producer gas in sub-system A has a chemical energy content, the required mass flow rate is less. This reduces the sub-system irreversibility by decreasing the availability losses in both the heat exchanger and trimburner.

Overall Biomass Plant Net Efficiency

The net first law efficiency values for the entire biomass power plant (system 7) incorporated with the three sub-systems are shown in figures 5.19 and 5.20 as a function of sub-system equivalence ratio. The net efficiency as defined here is based on the higher heating value of the wood. The term "net" as used here is defined as the gross power generated in the power plant minus the power required to run all the auxiliary components (compressors, pumps, fans, etc.). Figure 5.19 shows the overall plant net efficiency as a function of equivalence ratio at wood moisture contents of .05 and .30. Figure 5.20 shows the net efficiency at wood moisture contents of .15 and .50.
Figure 5.19  Plant net efficiency of system 7 employing the sub-systems as a function of equiv. ratio at wood moisture contents of .05 and .30.

Figure 5.20  Plant net efficiency of system 7 employing the sub-systems as a function of equiv. ratio at wood moisture contents of .15 and .50.
As can be seen in figures 5.19 and 5.20, the plant net efficiency decreases with increasing wood moisture content. This expected result is due to the higher energy loss associated with the combustion and gasification of wet wood. As the wood moisture content increases from .05 to .50, the plant net efficiency decreases about 3 percentage points for all three sub-systems.

The plant net efficiency is also seen to decrease slightly with increasing sub-system equivalence ratio. This result is due solely to the performance of the sub-systems. As discussed earlier, the sub-system efficiency decreases with increasing equivalence ratio. This effect alone is the cause of the slight dip in plant net efficiency over the range of increasing equivalence ratio.

The most important point to note when observing figures 5.19 and 5.20 is the relative net efficiency of the power plant when the three different sub-systems are employed. It appears that the incorporation of sub-system C yields the highest plant net efficiency of the three wood conversion sub-systems considered in this study. In addition, the net efficiency of the plant with sub-system C is higher over the entire range of equivalence ratios. This result is not evident when only the sub-system efficiencies are considered. Figures 5.13 and 5.14 do not indicate the same efficiency trends as the plant net efficiency curve of figures 5.19 and 5.20.
The reason that sub-system C appears more desirable in the plant net efficiency curves (figures 5.19 and 5.20) than in the sub-system efficiency curves (figures 5.13 and 5.14), is due to the definition of efficiency used in each case. In the sub-system efficiency definition, the work rate of the sub-system compressors (sub-systems A and B) was specified as an energy input and, therefore, appeared in the denominator of the efficiency equation. This type of definition penalizes sub-system A and B less for the energy required to run the compressors. In the plant net definition, the sub-system compressor power is specified as an auxiliary power requirement and is subtracted from the gross power produced by the plant. This calculation occurs in the numerator of the net efficiency definition. This definition penalizes sub-systems A and B (especially sub-system A) more severely for the energy consumed by the sub-system compressors. The result is that sub-systems A and B (particularly sub-system A with the high compressor power requirement) will have lower plant net efficiencies. The effect is that sub-system C (with no sub-system compressor power requirement) will appear more desirable in the plant net efficiency curves than in the sub-system efficiency curves.

Another point to note concerning the plant net efficiency curves in figures 5.19 and 5.20 is the
increasing differential of efficiencies between sub-systems C (and B) and sub-system A as the equivalence ratio increases. Again, this result is not observed in the sub-system efficiency curves of figures 5.13 and 5.14. This effect is due not only to the definitions as discussed above, but to the gas turbine inlet flow rate as well. Sub-systems C (and B) have higher producer gas flow rates entering the trimburner than sub-system A. This trend is even more pronounced at high equivalence ratios. The higher trimburner inlet gas rates result in higher flow rates entering the gas turbine. This is shown graphically in figures 5.21, 5.22, and 5.23 for sub-systems A, B, and C respectively. As can be seen in figures 5.21 through 5.23, the inlet mass flow rate to the gas turbine increases with equivalence ratio for all three sub-systems. However, the increase in gas turbine inlet mass rate is much greater for sub-systems B and C. The higher gas turbine mass rates in sub-systems B and C will produce more net power out of the gas turbine/compressor set than the lower mass rate of sub-system A. This effect is shown in figures 5.24, 5.25, and 5.26 which show the net power out of the gas turbine/compressor set as a function of equivalence ratio for sub-systems A, B, and C respectively. Thus, the plant net efficiency with sub-systems B and C incorporated will be disproportionately higher than the plant net efficiency with sub-system A.
Figure 5.21  Gas turbine mass flow rate as a function of equivalence ratio for sub-system A at four levels of wood moisture content.

Figure 5.22  Gas turbine mass flow rate as a function of equivalence ratio for sub-system B at four levels of wood moisture content.
Figure 5.23  Gas turbine mass flow rate as a function of equivalence ratio for sub-system C at four levels of wood moisture content.

Figure 5.24  Gas turbine net power as a function of equivalence ratio for sub-system A at four levels of wood moisture content.
Figure 5.25 Gas turbine net power as a function of equivalence ratio for sub-system B at four levels of wood moisture content.

Figure 5.26 Gas turbine net power as a function of equivalence ratio for sub-system C at four levels of wood moisture content.
used as the equivalence ratio is increased.

The tendency for higher gas turbine flow rates as the equivalence ratio increases may present a problem in the blading of the gas turbine. The increased mass flow rates through the gas turbine will eventually cause the gas to achieve supersonic speeds in the blading. At this point, the flow through the turbine blades becomes choked and the mass flow rate cannot increase any further. For this reason, a 5% increase in mass flow rate over the base flow rate (mass rate through compressor=36.8 lb/s) has been specified as the maximum allowable mass rate increase that the gas turbine can tolerate. This limit is shown graphically in figures 5.21, 5.22, and 5.23 for sub-systems A, B, and C respectively. The effect of this limit is that the increased power generated in the gas turbine due to high flow rates at high equivalence ratios is unobtainable. This is not a major concern, however, because all the efficiency curves indicate that the maximum plant performance occurs at low equivalence ratios anyway.

SUMMARY OF RESULTS

The results obtained from the computer runs of the sub-systems revealed several important characteristics of each gasifier/trimburner sub-system. A brief summary of
the important conclusions made concerning the three sub-systems is provided here.

Perhaps the most important conclusion made concerning the sub-systems was that of determining the optimum operating conditions. From the results of the computer runs, it was determined that the most efficient operation occurred at the lowest possible equivalence ratio for all three sub-systems. At the low equivalence ratios, chemical energy in the producer gas is maximized. Therefore, for the sub-system designs investigated, the production of chemical energy in the producer gas is more desirable than the production of sensible energy.

One major difference observed in the three sub-systems was that of the type of energy contained in the producer gas. Sub-system A yields a producer gas which is lower in temperature (sensible energy), but higher in percent of combustibles (chemical energy). Sub-systems B and C yield a higher temperature producer gas which has a lower percent of combustible products. This results in sub-systems B and C having higher producer gas flow rates to achieve the same heating requirement in the trimburner. In addition, it was observed that the higher flow rates associated with sub-systems B and C increase the irreversibility.

The location of the sub-system B compressor was found to be much more advantageous than the location of the sub-
system A compressor. It was determined that it takes about 5 times more energy to compress the producer gas via compression of the producer gas rather than compressing the inlet air to the gasifier. The high auxiliary power requirement of sub-system A ultimately yielded that this design is the least favorable when the overall power plant efficiency was used as the deciding criteria. When operating under the assumption that nearly all the sensible energy in the raw producer gas is recoverable, sub-system C appears to be the best choice of gasifier/trimburner systems.

Two important operating constraints were also identified from the computer runs. At a moisture content of .50, the equivalence ratio must be increased slightly to assure that total drying will occur in the gasifier. If the inlet wood is not dried completely, the gasifier will fail to operate. This limit is less restrictive for sub-systems B and C (equivalence ratio limit=.29) than for sub-system A (equivalence ratio=.31) due to the higher producer gas temperature associated with the pressurized gasifier sub-systems. The second operating constraint occurred in the high equivalence ratio range. To avoid a choked flow condition in the gas turbine, the total mass flow rate entering the turbine had to be kept below a maximum limit. This condition defined an upper limit on
the equivalence ratio since the mass flow rate to the turbine increases with increasing equivalence ratios.

COMPARISON WITH OIL-FIRED TRIMBURNER DESIGN

With the most effective wood-fired trimburner system identified (sub-system C), comparisons to conventional trimburner designs can be made. The trimburner design used in previous biomass studies was an oil-fired duct burner installed in the gas turbine inlet air stream. Therefore, this type of trimburner design will be used as the basis of this comparison.

The plant net efficiency for system 7 of the biomass plant with the oil-fired trimburner and wood-fired sub-system C design is shown in figure 5.27. The net efficiency used for sub-system C is at the optimum equivalence ratio for each wood moisture content considered. As shown in figure 5.27, the difference in net efficiency between the two trimburner designs is very small. The net efficiency of the plant with the oil-fired trimburner is slightly higher at high wood moisture contents because the performance of this design is independant of the inlet wood condition. Since the net efficiency of the two trimburner designs is nearly the same, other factors such as economics should be considered.
Figure 5.27 Comparison of system 7 plant net efficiency using the oil-fired and wood-fired trimburner designs at various wood moisture contents.

Figure 5.28 Simple payback period for the oil-fired and wood-fired trimburner designs as a function of oil cost for a .50 wood moisture content.
In an attempt to take a quick look at the scale of economics involved with these two trimburner designs, some simple evaluations were made. A total capital investment for each trimburner design was obtained based on the approximate component costs listed in table 5.1. Using this in conjunction with the current fuel prices given in table 5.1, simple payback periods were calculated.

The results of the simplified economic analysis discussed above indicated that the payback period for the oil-fired trimburner design was .86 years, whereas the payback period for sub-system C was 3.75 years. These payback periods were based on the net gain in electrical output when each of the trimburner designs were incorporated into the plant. Factors such as increasing interest rates, useable life, salvage value, and federal tax credits were neglected. Based on this analysis, it appears that the oil-fired trimburner design is much more desirable than the wood-fired sub-system C design because of a much shorter payback period.

To determine at what point the wood-fired trimburner design may become desirable, a plot displaying the payback period as a function of increasing oil prices was constructed. Figure 5.28 shows this plot for both the wood-fired sub-system C design and the conventional oil-fired design at an inlet wood moisture content of .50. Since the payback period for the wood-fired trimburner
### Table 5.1 FACTORS USED IN TRIMBURNER ECONOMIC ANALYSIS

<table>
<thead>
<tr>
<th>Component</th>
<th>Sub-system Design</th>
<th>Capital Cost ($)</th>
<th>Conventional Design</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gasifier</td>
<td>$ 800,000</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas Cleaning Equipment</td>
<td>200,000</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Producer Gas Heat Exchanger</td>
<td>200,000</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Duct Burner</td>
<td>30,000</td>
<td></td>
<td>60,000</td>
</tr>
<tr>
<td>Oil Storage and Feed System</td>
<td></td>
<td></td>
<td>40,000</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>$ 1,230,000</td>
<td>$ 100,000</td>
<td></td>
</tr>
</tbody>
</table>

**Associated Fuel Costs**

- **Wood** $ 20 Oven Dry Ton
- **# 2 Oil** $ 23.656 42 - gallon Barrel
- **Electricity** $ .06 Kilowatt - hour
design is independent of oil cost, its curve is constant at 3.75 years. The payback period for the oil-fired design, however, increases with increasing oil costs until it intersects the constant payback period curve of the wood-fired system. This intersection represents the point at which the cost of oil has increased to the point where both systems have equal payback periods of 3.75 years. Oil prices any higher than this intersection value will result in the wood-fired trimburner design being more economically attractive. For the simple analysis conducted here, this "breakeven" point occurs at an oil cost of 32.61 dollars per 42-gallon barrel. Current oil cost is 23.66 dollars per 42-gallon barrel. The obvious economic disadvantage of the wood-fired trimburner design may be somewhat offset by the increased availability of wood fuel over oil in remote plant site applications.

CONCLUSIONS

The objective of this study was to investigate an efficient means of boosting the temperature of the air entering the gas turbine in a wood-fired, combined cycle power plant. Because of the nature of the power plant considered, wood was considered the only fuel which could be used to boost the gas turbine air stream. Of the several biomass plant variations identified in previous
studies of this plant, only system 7 was investigated here due to its superiority in plant efficiency and mobility.

Wood conversion techniques were investigated in an effort to convert the wood to a synthesis fuel product which would burn cleanly enough to allow direct combustion in the gas turbine inlet air stream. Three conversion techniques, pyrolysis, gasification, and methanol production, were investigated and evaluated based on applicability to the trimburner concept. Of the three conversion techniques analyzed, only gasification appeared to yield a synthesis fuel (producer gas) which satisfied all the requirements of the trimburner design. In addition, an up-draft gasifier design was specified as the best model to use because of its ability to scale up to large sizes and handle high moisture content wood.

With the up-draft gasifier chosen as the wood conversion technique of choice, three gasifier/trimburner designs were developed. These designs, labeled sub-systems A, B, and C, were then modelled and simulated. The designs varied in the way the required compression of the producer gas entering the trimburner was accomplished. Sub-system A used an atmospheric gasifier and pressurized the producer gas after it exited the gasifier via a separate compressor. Sub-system B employed a small compressor to compress the air entering the gasifier; therefore simulating a pressurized gasifier system. Sub-
system C was similar to sub-system B with the exception that the pressurized inlet air to the gasifier was obtained by bleeding compressed air from the exit of the power plant's main compressor. All three sub-systems assumed that the sensible energy in the producer gas could be fully recovered through the use of a heat exchanger located at the exit of the gasifier.

With the sub-systems developed, computer runs were made to evaluate the relative performance of the three gasifier/trimburner designs. The results of these runs indicated that the incorporation of sub-system C with the biomass system 7 variation yielded the highest plant net efficiency by about .5 percentage points. It should be emphasized, however, that the superiority of sub-system C is based on the assumption that nearly all of the sensible energy in the producer gas is recoverable in the sub-system heat exchanger. If the sensible energy cannot be recovered, results indicate that sub-system A would be more desirable because of the higher chemical energy content in the synthesis gas produced by the atmospheric gasifier.

It was also determined that the optimum performance of all three sub-systems was obtained by operating the gasifiers at low equivalence ratios. The lowest equivalence ratio obtainable for each sub-system is
dependant upon the moisture content of the wood. At wood moisture contents less than .50, the equivalence ratio could be held as low as .275 (below this point, 100% carbon conversion is not possible) in all three sub-systems. At wood moisture contents exceeding .50, the equivalence ratio has to be increased to facilitate complete drying of the incoming wood and insure gasifier operation. Upper limits on operating equivalence ratio was dependant upon preventing choked flow conditions in the gas turbine due to high producer gas flow rates.

A final analysis was conducted concerning the comparision of the best wood-fired trimburner design (sub-system C) to the conventional oil-fired trimburner design evaluated in previous studies. It was determined that the net efficiencies of the system 7 plant with each type of trimburner design employed were approximately equal. Because of this, the comparision of the two trimburner designs was based on economic considerations. An economic analysis to determine simple payback periods for both the wood-fired and oil-fired trimburner designs was conducted. The results of that analysis indicate that the payback period for the oil-fired trimburner (.86 years) was much more attractive than the payback period for the wood-fired trimburner design (3.75 years). This conclusion was due mainly to the high capital investment cost associated with wood-fired trimburner design (sub-
system C). For the wood-fired trimburner design to become economically feasible, it was estimated that the cost of oil would have to increase from its current cost of 23.66 dollars per barrel to 32.61 dollars per barrel. This represents an increase in oil prices of 38%.
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APPENDICES
APPENDIX A

DESCRIPTION OF BIOMASS PLANT VARIATIONS

Several variations on the base case plant design discussed in chapter one were investigated in previous studies. The base case configuration was described in chapter one, and will, therefore, not be repeated here. In addition, systems 4 and 7 are omitted from this discussion since they were addressed in chapter one.

System 2

System 2, as shown in figure A.1, is identical to the base case with the exception of the combustion gas mixing stream (c16). This stream was used to provide high temperature gas to the dryer to dry the wood. It was found that removing this stream yielded a higher plant efficiency as well as a more compact power plant design. Because the mixing stream (c16) is no longer available for fuel drying, the flue gas exiting the boiler for system 2 must be at a higher temperature to meet the drying requirements. Therefore, the flue gas temperature exiting the boiler is no longer constant (as in system 1), but will vary depending on the amount of fuel drying required.

Because system 2 displays obvious advantages over system 1, it was chosen as the new base case system. All
Figure A.1 Schematic diagram of system 2 power plant.
subsequent systems considered will then have the system 2 configuration as a starting point.

System 3

System 3 considers the effect of steam injection on the overall system. Figure A.2 shows that a small amount of steam is extracted from the steam turbine and injected into the compressed air stream entering the air/gas heat exchanger. The steam is injected into the air stream to increase the mass flow rate through the gas turbine, thus increasing the output power.

System 5

System 5, shown in figure A.3, represents wet fuel combustion. Most of the fuel preparation equipment used in the previous systems has been removed. This means that wood with moisture contents as high as 60% will be entering the combustor. Because of this, a different combustor which can handle the wet fuel is used.

System 6

System 6 represents the use of a ceramic heat exchanger in place of the metallic heat exchanger used in the previous systems. The equipment diagram looks the same as for system 2. Because of the ceramic materials used in the heat exchanger, air stream temperatures can
Figure A.2 Schematic diagram of system 3 power plant.
Figure A.3 Schematic diagram of system 5 power plant.
now exceed 1450 F. For this particular system, an air stream (Ta03) temperature of 1612 F has been specified.

System 8

System 8 is a combination of the wet fuel combustion of system 5 and the ceramic heat exchanger of system 6.

Systems 9, 10, and 11

Systems 9, 10, and 11 are the same as systems 7, 6, and 8 respectively except that the air temperature into the turbine is 1800 F.
APPENDIX B

NEWTON-RAPHSON METHOD

Solutions to a system of simultaneous non-linear equations are solved most generally by either successive substitution or the Newton-Raphson method. The latter technique is used in this study and a brief description is provided here [Stoecker, 1980].

Suppose that three non-linear equations are to be solved for three unknown variables $X_1$, $X_2$, and $X_3$ from:

$$f_1(X_1, X_2, X_3) = 0$$
$$f_2(X_1, X_2, X_3) = 0$$
$$f_3(X_1, X_2, X_3) = 0$$

The scheme for solving these equations is an iterative one which follows eight general steps.

1. Rewrite the equations so that all the terms are on one side (set the equation equal to zero as above).
2. Assume initial guesses at the values of the unknowns $X_{i1}$, $X_{i2}$, and $X_{i3}$.
3. Calculate the value of the functions $f_1$, $f_2$, and $f_3$ using the initial guesses of the unknowns ($X_{i1}$, $X_{i2}$, and $X_{i3}$).
4. Compute the partial derivatives of the functions ($f_1$, $f_2$, $f_3$) with respect to all the unknown variables ($X_1$, $X_2$, $X_3$).
5. Use a Taylor-series expansion to approximate the functions. This establishes a set of simultaneous linear equations as shown below.

\[ f_1(x_{1i}, x_{2i}, x_{3i}) = f_1(x_1, x_2, x_3) + \frac{\partial f_1(x_{1i}, x_{2i}, x_{3i})}{\partial x_1} (x_{1i} - x_1) + \frac{\partial f_1(x_{1i}, x_{2i}, x_{3i})}{\partial x_2} (x_{2i} - x_2) + \frac{\partial f_1(x_{1i}, x_{2i}, x_{3i})}{\partial x_3} (x_{3i} - x_3) \]

where

\[ x_1, x_2, x_3 = \text{solution set of the system} \]

This yields a linear system of equations which can be put in matrix form as:

\[
\begin{bmatrix}
\frac{\partial f_1}{\partial x_1} & \frac{\partial f_1}{\partial x_2} & \frac{\partial f_1}{\partial x_3} \\
\frac{\partial f_2}{\partial x_1} & \frac{\partial f_2}{\partial x_2} & \frac{\partial f_2}{\partial x_3} \\
\frac{\partial f_3}{\partial x_1} & \frac{\partial f_3}{\partial x_2} & \frac{\partial f_3}{\partial x_3}
\end{bmatrix}
\begin{bmatrix}
(x_{1i} - x_1)
(x_{2i} - x_2)
(x_{3i} - x_3)
\end{bmatrix}
= 
\begin{bmatrix}
f_1 \\
f_2 \\
f_3
\end{bmatrix}
\]

6. The above matrix is then solved using a linear equation to yield the values of

\[
(x_{1i} - x_1) \\
(x_{2i} - x_2) \\
(x_{3i} - x_3)
\]
These values are actually a measure of the difference between the initial guesses and the actual solution.

7. Update the initial guesses using the solution set obtained in step 6.

\[
\begin{align*}
X_{i\text{new}} &= X_{i\text{old}} - (X_i - X) \\
X_{2i\text{new}} &= X_{2i\text{old}} - (X_{2i} - X_2) \\
X_{3i\text{new}} &= X_{3i\text{old}} - (X_{3i} - X_3)
\end{align*}
\]

8. Test for convergence by checking to see if the updated guesses \(X_{i\text{new}}, X_{2i\text{new}}, X_{3i\text{new}}\) are within the convergence criteria of the old guesses \(X_{i\text{old}}, X_{2i\text{old}}, X_{3i\text{old}}\). If this is the case, then \(X_{i\text{new}}, X_{2i\text{new}}, \) and \(X_{3i\text{new}}\) is the solution set to the system. If the convergence test is not satisfied, return to step 3 using the updated values of the guesses and proceed through another iteration.

This procedure, like successive substitution, is dependant upon the accuracy of the initial guesses. If the initial guesses are within some radius of convergence around the solution set, then a solution will quickly be obtained. Initial guesses outside the radius of convergence will result in divergence of the results obtained and no solution will be found.
To clarify the process described above, an example employing the Newton-Raphson scheme is provided [Stoecker, 1980, example 6-3].

Example 6-3 Solve Example 6-1 by the Newton-Raphson method.

SOLUTION

In example 6-1, four non-linear simultaneous equations were identified as shown below. The unknowns to be determined in this problem are \( d_p, w_1, w_2, \) and \( w \).

step 1.

\[
\begin{align*}
 f_1 &= d_p - 7.2w^2 - 392.28 = 0 \\
 f_2 &= d_p - 810 + 25w_1 + 3.75w_2^2 = 0 \\
 f_3 &= d_p - 900 + 65w_2 + 30w_2^2 = 0 \\
 f_4 &= w - w_1 - w_2 = 0
\end{align*}
\]

step 2. Choose trial values of the variables, which are here selected as \( d_p=750, w_1=3, w_2=1.5, \) and \( w=5 \).

step 3. Calculate the magnitudes of the f's at the temporary values of the variables, \( f_1=177.7, f_2=48.75, f_3=15.0, \) and \( f_4=0.5 \).

step 4. The partial derivatives are shown in Table 6-4.
Table 6-4

<table>
<thead>
<tr>
<th></th>
<th>( \frac{\partial}{\partial dp} )</th>
<th>( \frac{\partial}{\partial w_1} )</th>
<th>( \frac{\partial}{\partial w_2} )</th>
<th>( \frac{\partial}{\partial w} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>( \delta f_1 )</td>
<td>1</td>
<td>0</td>
<td>0</td>
<td>-14.4w</td>
</tr>
<tr>
<td>( \delta f_2 )</td>
<td>1</td>
<td>25 + 7.5w_1</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>( \delta f_3 )</td>
<td>1</td>
<td>0</td>
<td>65 + 60w_2</td>
<td>0</td>
</tr>
<tr>
<td>( \delta f_4 )</td>
<td>0</td>
<td>-1</td>
<td>-1</td>
<td>1</td>
</tr>
</tbody>
</table>

Step 5. Substituting the temporary values of the variables into the equations for the partial derivatives forms a set of linear simultaneous equations to be solved for the corrections to \( x \):

\[
\begin{bmatrix}
1.0 & 0.0 & 0.0 & -72.0 \\
1.0 & 47.5 & 0.0 & 0.0 \\
1.0 & 0.0 & 155.0 & 0.0 \\
0.0 & -1.0 & -1.0 & 1.0 \\
\end{bmatrix}
\begin{bmatrix}
\Delta x_1 \\
\Delta x_2 \\
\Delta x_3 \\
\Delta x_4 \\
\end{bmatrix}
= \begin{bmatrix}
177.7 \\
48.75 \\
15.0 \\
0.50 \\
\end{bmatrix}
\]

where

\( \Delta x_1 = x_{1i} - x_1 \), \( \Delta x_3 = x_{3i} - x_3 \)
\( \Delta x_2 = x_{2i} - x_2 \), \( \Delta x_4 = x_{4i} - x_4 \)
step 6. Solution of the simultaneous equations

$$\Delta x_1 = 98.84, \quad \Delta x_2 = -1.055, \quad \Delta x_3 = -0.541, \quad \Delta x_4 = -1.096$$

Table 6-5

<table>
<thead>
<tr>
<th>After iteration</th>
<th>f1</th>
<th>f2</th>
<th>f3</th>
<th>f4</th>
<th>dp</th>
<th>w1</th>
<th>w2</th>
<th>w</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>-8.64</td>
<td>4.17</td>
<td>8.78</td>
<td>0.00</td>
<td>651.2</td>
<td>4.06</td>
<td>2.04</td>
<td>6.10</td>
</tr>
<tr>
<td>2</td>
<td>-0.08</td>
<td>0.02</td>
<td>0.06</td>
<td>0.00</td>
<td>650.5</td>
<td>3.99</td>
<td>2.00</td>
<td>5.99</td>
</tr>
<tr>
<td>3</td>
<td>0.000</td>
<td>0.00</td>
<td>0.00</td>
<td>0.00</td>
<td>650.5</td>
<td>3.99</td>
<td>2.00</td>
<td>5.99</td>
</tr>
</tbody>
</table>

step 7. The corrected values of the variables are

These values of the variables are returned to step 3 for the next iteration. The values of the f's and the variables resulting from continued iterations are shown in Table 6-5.

The calculations converged satisfactorily after three iterations.
APPENDIX C

SAMPLE OUTPUT FROM INTEGRATED POWER PLANT SIMULATION AND FORTRAN SOURCE CODES FOR EACH SUB-SYSTEM DESIGN

This appendix is provided to display the type of output which is obtained from the power plant simulation program integrated with the sub-system simulations. In addition, the fortran source codes used to model each sub-system are provided. The source codes for each sub-system are identified as follows:

\[ \begin{align*}
TBN1 &= \text{Sub-system A wood-fired trimburner model} \\
TBN2 &= \text{Sub-system B wood-fired trimburner model} \\
TBN3 &= \text{Sub-system C wood-fired trimburner model}
\end{align*} \]
--- COMPRESSOR ---

Conditions at:

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Exit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure</td>
<td>14.70</td>
<td>132.42</td>
</tr>
<tr>
<td>Temperature</td>
<td>59.00</td>
<td>595.52 F</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>36.68</td>
<td>36.68 lb/s</td>
</tr>
<tr>
<td>Exergy</td>
<td>0.00</td>
<td>4644.47 kW</td>
</tr>
</tbody>
</table>

Inlet pressure loss | 4.00 in. h2o |
Bleeding air         | 4.00 %       |
Efficiency           | 83.20 %      |
Power to drive       | 5310.08 kW   |
Irreversibility      | 665.62 kW    |
TRIMBURNER SUB-SYSTEM

--GASIFIER SECTION--

Gasifier Pressure = 14.6960000 psi
Gasifier Flame Temperature = 1400.0910000 F
Product Gas Exit Temperature = 948.6503000 F
Inlet Wet Wood Mass Rate = 3.202306E-001 lbs/s
Inlet Dry Wood Mass Rate = 2.721960E-001 lbs/s
HHV of Gas before cyclone = 2704.9490000 Btu/lb
HHV of Gas after cyclone = 2953.0480000 Btu/lb
Mass Rate of Product Gas = 7.338354E-001 lbs/s
Composition of Product Gas:

CO: 8.091748E-001 moles/mole of dry wood gasified
CO2: 1.433862E-001 moles/mole of dry wood gasified
H2: 5.308191E-001 moles/mole of dry wood gasified
H2O: 8.984525E-002 moles/mole of dry wood gasified
CH4: 4.743905E-002 moles/mole of dry wood gasified
N2: 1.0812480 moles/mole of dry wood gasified
Ar: 1.288200E-002 moles/mole of dry wood gasified

--TRIMBURNER SECTION--

Mass Rate of Air Entering = 36.6792000 lbs/s
Mass Rate of Air Exiting = 37.4130400 lbs/s
Product Gas Inlet Temp. = 939.6503000 F
Inlet Air Temperature = 1450.0000000 F
Exit Air Temperature = 1612.0000000 F
Work Rate of Gas Compressor = 107.6704000 KW
Overall Sub-system Efficiency = 80.4449800 %
Overall Irreversibility = 1051.5960000 Kw
Second Law Efficiency = 60.0880900 %
GAS TURBINE

Conditions at:

<table>
<thead>
<tr>
<th>Inlet</th>
<th>Exit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure</td>
<td>128.85</td>
</tr>
<tr>
<td>Temperature</td>
<td>1612.00</td>
</tr>
<tr>
<td>Mass flow</td>
<td>37.41</td>
</tr>
<tr>
<td>Exergy</td>
<td>11841.39</td>
</tr>
</tbody>
</table>

Mass flow rate 37.41 lb/s
Cooling flow 1.53 lb/s
Power output 8391.79 kW
Irreversibility 468.03 kW

COMBUSTOR

Air-Fuel ratio (dry to dry) 17.0000
Fraction radiation loss .0200
Fraction unburned carbon .0200
Fraction of carbon burned to CO .0100
Excess air level = 151.7421 percent
Adiabatic flame temperature = 2141.8 deg F
Irreversibility rate = 7802.0 kW
Second Law Efficiency = 67.79 Percent

MIXER after combustor
Irreversibility = 963.44 kW

Cyclone #3
Pressure loss = 5.000

Splitter: after combustor
----- STEAM-GAS HEAT EXCHANGER -----

Flow conditions:

<table>
<thead>
<tr>
<th></th>
<th>flue gas</th>
<th>steam</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature at inlet, F</td>
<td>1675.000</td>
<td>700.891</td>
</tr>
<tr>
<td>Pressure at inlet, psia</td>
<td>15.057</td>
<td>640.000</td>
</tr>
<tr>
<td>Massflow, lbm/sec</td>
<td>57.540</td>
<td>9.458</td>
</tr>
<tr>
<td>Temperature at exit, F</td>
<td>1616.370</td>
<td>900.000</td>
</tr>
<tr>
<td>Pressure at exit, psia</td>
<td>15.054</td>
<td>626.225</td>
</tr>
</tbody>
</table>

| Cp - values, btu/lbm F | .298  | .534  |
| Viscosity, lbm/s ft    | .30573E-04 | .17247E-04 |
| Conductivity, btu/h ft R | .43271E-01 | .36057E-01 |
| Exergy at inlet, KW    | 15988.280 | 5243.733 |
| Exergy at exit, KW     | 15187.670 | 5855.762 |

Specifications of Heat Exchanger:

| Tube arrangement: | in-line |
| Spacing of tubes normal to flow, Sn/d: | 2.000 |
| Spacing of tubes parallel to flow, Sp/d: | 2.000 |
| Number of tubes normal to flow: | 60.000 |
| Number of tubes parallel to flow: | 1.000 |
| Number of U's: | 1.000 |
| Outside diameter of pipes: | .750 inches |
| Inside diameter of pipes | .560 inches |
| Thermal conductivity of the tubes: | 20.000 btu/h/ft/F |
| Roughness factor inside the tubes (k/D): | .002 |
| Length of the tubes: | 12.792 ft |
| Height of the heat exchanger: | 12.917 ft |
| Width of the heat exchanger: | 7.500 ft |
| Total heat transfer area: | 301.397 sqft |
| Maximum velocity: | 53.246 ft/sec |

Heat Transfer Data:

| "C" - to calculate Nusselt Number: | .254 |
| "X" - to calculate Nusselt Number: | .632 |
| Safety factor on the shell side: | .900 |
| Correction factor due to cross flow: | .999 |
| Correction factor due to few pipes: | .797 |
| H out - heat transfer coefficient: | 16.608 btu/h/sqft/F |
| H in - heat transfer coefficient: | 350.629 btu/h/sqft/F |
| Fouling in heat exchanger: | .005 |
| U - Overall heat transfer coefficient | 14.391 btu/h/sqft/F |
| UA - value: | 4337.344 btu/h/F |
| Log-Mean-Temperature-Difference: | 843.290 F |
| Maximum wall temperature: | 935.048 F |
| Average wall temperature: | 928.670 F |
| Total heat transfer: | 1061.792 kW |
| Irreversibility: | 188.588 kW |

Pressure drops:

| Pressure drop inside pipes | 13.775 psi |
| Pressure drop pr. ft of pipe | .538 psi/ft |
| Pressure drop on gas side | .080 in W.C. |
----- AIR-GAS HEAT EXCHANGER -----

Flow conditions:
- Temperature at inlet, F: flue gas 1616.370, air 595.520
- Pressure at inlet, psia: flue gas 15.054, air 132.418
- Mass flow, lbm/sec: flue gas 57.540, air 36.679
- Temperature at exit, F: flue gas 1118.579, air 1450.000
- Pressure at exit, psia: flue gas 14.820, air 129.847

Cp values, btu/lbm F: flue gas .291, air .266
Viscosity, lbm/s ft: flue gas .27997E-04, air .24998E-04
Conductivity, btu/h ft R: flue gas .39138E-01, air .33031E-01
Exergy at inlet, KW: flue gas 15137.670, air 4644.467
Exergy at exit, KW: flue gas 9870.336, air 10257.930

Specifications of Heat Exchanger:
- Tube arrangement: staggered
  - Spacing of tubes normal to flow, Sn/d: 1.700
  - Spacing of tubes parallel to flow, Sp/d: 1.500
  - Number of tubes normal to flow: 72.000
  - Number of tubes parallel to flow: 8.000
  - Number of U's: 3.000
  - Outside diameter of pipes: 1.000 inches
  - Inside diameter of pipes: .810 inches
  - Thermal conductivity of the tubes: 20,000 btu/h/ft/F
  - Roughness factor inside the tubes (k/D): .002
  - Length of the tubes: 7.564 ft
  - Height of the heat exchanger: 7.689 ft
  - Width of the heat exchanger: 10.200 ft
  - Total heat transfer area: 6844.123 sqft
  - Maximum velocity: 68.869 ft/sec

Heat Transfer Data:
- "C" - to calculate Nusselt Number: .507
- "X" - to calculate Nusselt Number: .564
- Safety factor on the shell side: .900
- Correction factor due to cross flow: 1.000
- Correction factor due to few pipes: 1.000
- H out - heat transfer coefficient: 24.531 btu/h/sqft/F
- H in - heat transfer coefficient: 49.754 btu/h/sqft/F
- Fouling in heat exchanger: .005
- U - Overall heat transfer coefficient: 14.081 btu/h/sqft/F
- UA - value: 96371.950 btu/h/F
- Log-Mean-Temperature-Difference: 311.339 F
- Maximum wall temperature: 1504.940 F
- Average wall temperature: 1136.594 F
- Total heat transfer: 8791.349 kW
- Irreversibility: 703.370 kW

Pressure Drop:
- Pressure drop inside pipes: 2.572 psi
- Pressure drop pr. ft of pipe: .057 psi/ft
- Pressure drop on gas side: 6.493 in W.C.
<table>
<thead>
<tr>
<th>Component</th>
<th>Irreversibility</th>
<th>Pressure Loss</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mixer to Dryer</td>
<td>0.00 kW</td>
<td>0.00</td>
</tr>
<tr>
<td>Air pollution</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pressure loss</td>
<td>0.00</td>
<td></td>
</tr>
<tr>
<td>Cyclones #1, #2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pressure loss</td>
<td>0.00</td>
<td></td>
</tr>
<tr>
<td>Bag house</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pressure loss</td>
<td>0.00</td>
<td></td>
</tr>
<tr>
<td>Waste Heat Boiler</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

### Flow Conditions:

<table>
<thead>
<tr>
<th></th>
<th>Flue Gas</th>
<th>Steam</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature at inlet, F</td>
<td>1118.579</td>
<td>115.223</td>
</tr>
<tr>
<td>Pressure at inlet, psia</td>
<td>14.820</td>
<td>650.00</td>
</tr>
<tr>
<td>Massflow, lbm/sec</td>
<td>57.540</td>
<td>640.00</td>
</tr>
<tr>
<td>Temperature at exit, F</td>
<td>700.891</td>
<td>9.458</td>
</tr>
<tr>
<td>Pressure at exit, psia</td>
<td>14.531</td>
<td>640.00</td>
</tr>
</tbody>
</table>

### Heat Transfer Data:

<p>| | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>dT at pinch point</td>
<td>107.108  F</td>
<td>494.721  F</td>
</tr>
<tr>
<td>Saturation temperature</td>
<td>1002.660</td>
<td>6923.844</td>
</tr>
<tr>
<td>Superheater surface area</td>
<td>9323.790</td>
<td>640.00</td>
</tr>
<tr>
<td>Boiler surface area</td>
<td>10250.290</td>
<td>1766.882</td>
</tr>
<tr>
<td>Economizer surface area</td>
<td>12561.320</td>
<td>1766.882</td>
</tr>
</tbody>
</table>

### Pressure Drops:

<table>
<thead>
<tr>
<th>Pressure drop on steam side (approx)</th>
<th>10,000 psi</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure drop on gas side</td>
<td>8,000 in W.C.</td>
</tr>
</tbody>
</table>
### STEAM TURBINE

Conditions at:

<table>
<thead>
<tr>
<th>Inlet</th>
<th>Exit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure</td>
<td>626.23</td>
</tr>
<tr>
<td>Temperature</td>
<td>900.00</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>9.46</td>
</tr>
<tr>
<td>Exergy</td>
<td>5661.33</td>
</tr>
<tr>
<td>Efficiency</td>
<td>80.00 %</td>
</tr>
<tr>
<td>Exit quality</td>
<td>93.80 %</td>
</tr>
<tr>
<td>Power output</td>
<td>3984.281 kW</td>
</tr>
<tr>
<td>Irrevers.</td>
<td>898.657 kW</td>
</tr>
</tbody>
</table>

### CONDENSER

<table>
<thead>
<tr>
<th></th>
<th>1.50 psia</th>
</tr>
</thead>
<tbody>
<tr>
<td>Quality of steam</td>
<td>.00 %</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>9.46 lbm/s</td>
</tr>
<tr>
<td>Cooling load</td>
<td>.00 Kw</td>
</tr>
<tr>
<td>Coolant flow rate</td>
<td>.00 lbm/s</td>
</tr>
<tr>
<td>Irreversibility</td>
<td>.00 Kw</td>
</tr>
</tbody>
</table>

### PUMP

<table>
<thead>
<tr>
<th></th>
<th>9.458 lbm/s</th>
</tr>
</thead>
<tbody>
<tr>
<td>Inlet pressure</td>
<td>1.500 psia</td>
</tr>
<tr>
<td>Exit pressure</td>
<td>650.000 psia</td>
</tr>
<tr>
<td>Efficiency</td>
<td>68.000 %</td>
</tr>
<tr>
<td>Power required</td>
<td>29.822 kW</td>
</tr>
<tr>
<td>Irreversibility</td>
<td>9.543 kW</td>
</tr>
</tbody>
</table>

### Air pollution dev 3

Pressure loss = 5.000

### Splitter: after fan #3

Pressure loss = .000

### Air pollution dev 2

Pressure loss = .000

### Fan: FAN #1

Conditions at:

<table>
<thead>
<tr>
<th>Inlet</th>
<th>Exit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure</td>
<td>.70</td>
</tr>
<tr>
<td>Temperature</td>
<td>350.00</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>41.57</td>
</tr>
<tr>
<td>Exergy</td>
<td>.00</td>
</tr>
<tr>
<td>Power required</td>
<td>.00 kW</td>
</tr>
<tr>
<td>Irreversibility</td>
<td>.00 Kw</td>
</tr>
<tr>
<td>Delta P</td>
<td>.00 in. h2o</td>
</tr>
<tr>
<td>Flow rate</td>
<td>50802.79 cu ft/min</td>
</tr>
</tbody>
</table>
### Fan: FAN #3

<table>
<thead>
<tr>
<th>Conditions at:</th>
<th>Inlet</th>
<th>Exit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure</td>
<td>9.57</td>
<td>0.70</td>
</tr>
<tr>
<td>Temperature</td>
<td>350.00</td>
<td>350.00°F</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>57.54</td>
<td>57.54 lb/s</td>
</tr>
<tr>
<td>Exergy</td>
<td>0.00</td>
<td>0.00 kW</td>
</tr>
<tr>
<td>Power required</td>
<td>136.56 kW</td>
<td></td>
</tr>
<tr>
<td>Irreversibility</td>
<td>50.53 kW</td>
<td></td>
</tr>
<tr>
<td>Delta P</td>
<td>10.27 in. H2O</td>
<td></td>
</tr>
<tr>
<td>Flow rate</td>
<td>71223.38 cu ft/min</td>
<td></td>
</tr>
</tbody>
</table>

### Fan: FAN #4

<table>
<thead>
<tr>
<th>Conditions at:</th>
<th>Inlet</th>
<th>Exit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure</td>
<td>0.70</td>
<td>15.00</td>
</tr>
<tr>
<td>Temperature</td>
<td>350.00</td>
<td>350.00°F</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>15.97</td>
<td>15.97 lb/s</td>
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<tr>
<td>Exergy</td>
<td>0.00</td>
<td>0.00 kW</td>
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<tr>
<td>Power required</td>
<td>51.20 kW</td>
<td></td>
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<tr>
<td>Irreversibility</td>
<td>18.95 kW</td>
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<td>Delta P</td>
<td>14.30 in. H2O</td>
<td></td>
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<td>Flow rate</td>
<td>19186.01 cu ft/min</td>
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### Fan: FAN #5

<table>
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<th>Conditions at:</th>
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<tr>
<td>Pressure</td>
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<td>0.70</td>
</tr>
<tr>
<td>Temperature</td>
<td>350.00</td>
<td>350.00°F</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>41.57</td>
<td>41.57 lb/s</td>
</tr>
<tr>
<td>Exergy</td>
<td>0.00</td>
<td>0.00 kW</td>
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<tr>
<td>Power required</td>
<td>0.00 kW</td>
<td></td>
</tr>
<tr>
<td>Irreversibility</td>
<td>0.00 kW</td>
<td></td>
</tr>
<tr>
<td>Delta P</td>
<td>0.00 in. H2O</td>
<td></td>
</tr>
<tr>
<td>Flow rate</td>
<td>50802.79 cu ft/min</td>
<td></td>
</tr>
</tbody>
</table>

### Generator/Gear box: Gas-Turbine

| Irreversibility| 256.40 kW |

### Generator/Gear box: Steam-Turbine

| Irreversibility| 331.49 kW |

### Plant Summary

<table>
<thead>
<tr>
<th>Auxiliary Power (kW)</th>
<th>Generator Power (kW)</th>
</tr>
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<tbody>
<tr>
<td>Pump</td>
<td>29.82</td>
</tr>
<tr>
<td>Cooling twr</td>
<td>45.00</td>
</tr>
<tr>
<td>Fan #1</td>
<td>0.00</td>
</tr>
<tr>
<td>Fan #3</td>
<td>136.56</td>
</tr>
<tr>
<td>Fan #4</td>
<td>51.20</td>
</tr>
<tr>
<td>Fan #5</td>
<td>0.00</td>
</tr>
<tr>
<td>Conveyor</td>
<td>12.00</td>
</tr>
<tr>
<td>Hammermill</td>
<td>0.00</td>
</tr>
<tr>
<td>Screw</td>
<td>10.00</td>
</tr>
<tr>
<td>Dryer</td>
<td>0.00</td>
</tr>
<tr>
<td>TOTAL...</td>
<td>392.26</td>
</tr>
</tbody>
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<table>
<thead>
<tr>
<th>Plant Efficiency (%)</th>
<th>HHV</th>
<th>LHV1</th>
<th>LHV2</th>
<th>LHV3</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gross</td>
<td>27.5</td>
<td>29.5</td>
<td>30.5</td>
<td>30.5</td>
</tr>
<tr>
<td>Net</td>
<td>25.8</td>
<td>27.7</td>
<td>28.5</td>
<td>28.5</td>
</tr>
</tbody>
</table>
5 iterations, specified tolerance: .0100, actual tolerance: .0000

--- Constants ---

<table>
<thead>
<tr>
<th>AFRat</th>
<th>17.0000 lb air/lb fuel</th>
<th>Pg</th>
<th>14.6960 psia</th>
</tr>
</thead>
<tbody>
<tr>
<td>ch1</td>
<td>2.540</td>
<td>Prinj</td>
<td>2.0000 % injection</td>
</tr>
<tr>
<td>ch2</td>
<td>5.070</td>
<td>Prmp</td>
<td>9.1000 press. ratio</td>
</tr>
<tr>
<td>cmpbl1</td>
<td>4.0000 %</td>
<td>Ps1</td>
<td>650.0000 psia</td>
</tr>
<tr>
<td>dPap1</td>
<td>0.0000 in. h2o</td>
<td>Ps31</td>
<td>1.5000 psia</td>
</tr>
<tr>
<td>dPap2</td>
<td>0.0000 in. h2o</td>
<td>Ps39</td>
<td>1.5000 psia</td>
</tr>
<tr>
<td>dPap3</td>
<td>5.0000 in. h2o</td>
<td>Ps402</td>
<td>1.5000 psia</td>
</tr>
<tr>
<td>dBBag</td>
<td>0.0000 in. h2o</td>
<td>radlos</td>
<td>0.0200 frac. energy in</td>
</tr>
<tr>
<td>dPB1r0g</td>
<td>8.0000 in. h2o</td>
<td>radtm</td>
<td>0.0200 frac. HMVoil</td>
</tr>
<tr>
<td>dPB1rw</td>
<td>10.0000 psia</td>
<td>rufhx1</td>
<td>0.0200</td>
</tr>
<tr>
<td>dPCmb</td>
<td>11.0000 in. h2o</td>
<td>rufhx2</td>
<td>0.0200</td>
</tr>
<tr>
<td>dPCmp</td>
<td>4.0000 in. h2o</td>
<td>SnDh1</td>
<td>2.0000</td>
</tr>
<tr>
<td>dPCyc1</td>
<td>0.0000 in. h2o</td>
<td>SnDh2</td>
<td>1.7000</td>
</tr>
<tr>
<td>dPCyc2</td>
<td>5.0000 in. h2o</td>
<td>SpDh1</td>
<td>2.0000</td>
</tr>
<tr>
<td>dPdry</td>
<td>0.0000 in. h2o</td>
<td>SpDh2</td>
<td>1.5000</td>
</tr>
<tr>
<td>dPtgb</td>
<td>2.0000 in. h2o</td>
<td>Ta01</td>
<td>59.0000 F</td>
</tr>
<tr>
<td>dPr1m</td>
<td>1.0000 psi</td>
<td>Ta031</td>
<td>1450.0000 F</td>
</tr>
<tr>
<td>drylos</td>
<td>0.5000 fract ht loss</td>
<td>Ta032</td>
<td>1612.0000 F</td>
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<tr>
<td>EFFcmp</td>
<td>83.2000 %</td>
<td>Tc112</td>
<td>1675.0000 F</td>
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<tr>
<td>EFFfn1</td>
<td>63.0000 %</td>
<td>Tc21</td>
<td>350.0000 F</td>
</tr>
<tr>
<td>EFFfn2</td>
<td>63.0000 %</td>
<td>Tds</td>
<td>59.0000 F</td>
</tr>
<tr>
<td>EFFfn4</td>
<td>63.0000 %</td>
<td>Te61</td>
<td>350.0000 F</td>
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<tr>
<td>EFFfn5</td>
<td>63.0000 %</td>
<td>Ts38</td>
<td>900.0000 F</td>
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<tr>
<td>EFFgb1</td>
<td>98.0000 %</td>
<td>Tsuper</td>
<td>207.0000 F</td>
</tr>
<tr>
<td>EFFgb2</td>
<td>95.0000 %</td>
<td>typhx1</td>
<td>1.0000 1 = in-line</td>
</tr>
<tr>
<td>EFFgn1</td>
<td>95.0000 %</td>
<td>typhx2</td>
<td>2.0000 2 = staggered</td>
</tr>
<tr>
<td>EFFgn2</td>
<td>95.0000 %</td>
<td>Ub1r</td>
<td>10.0000 btu/hr-ft2-F</td>
</tr>
<tr>
<td>EFFgb3</td>
<td>91.2000 %</td>
<td>Ulecon</td>
<td>12.0000 btu/hr-ft2-F</td>
</tr>
<tr>
<td>EFFcmp</td>
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<td>unbcbn</td>
<td>0.2000 lb/1b</td>
</tr>
<tr>
<td>EFFcnb</td>
<td>80.0000 %</td>
<td>Usuper</td>
<td>8.0000 btu/hr-ft2-F</td>
</tr>
<tr>
<td>EDrat</td>
<td>0.2750 equiv. ratio</td>
<td>Wa01</td>
<td>0.0070 lb h2o/lb dry a</td>
</tr>
<tr>
<td>fouhx1</td>
<td>0.050 fouling factor</td>
<td>Wb51</td>
<td>1.5000 lb h2o/lb dry fuel</td>
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<tr>
<td>fouhx2</td>
<td>0.050 fouling factor</td>
<td>Wb52</td>
<td>1.5000 lb h2o/lb fuel</td>
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<tr>
<td>fshx1</td>
<td>0.9000 factor safety</td>
<td>WRCnv</td>
<td>12.0000 kW</td>
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<tr>
<td>fshx2</td>
<td>0.9000 factor safety</td>
<td>WRCtr</td>
<td>45.0000 kW</td>
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<tr>
<td>HHV</td>
<td>8880.0000 btu/lb</td>
<td>WRscw</td>
<td>10.0000 kW</td>
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<tr>
<td>HHVoil</td>
<td>3194.4050 btu/lb</td>
<td>WRCml</td>
<td>0.0000 kW</td>
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<tr>
<td>IDhx1</td>
<td>0.5000 inches</td>
<td>Xash</td>
<td>0.0050 lb ash/lb dry f</td>
</tr>
<tr>
<td>IDhx2</td>
<td>0.8100 inches</td>
<td>Xash01</td>
<td>0.0000 lb ash/lb dry oil</td>
</tr>
<tr>
<td>Kpxh1</td>
<td>20.0000 btu/hr-ft-F</td>
<td>XH2</td>
<td>0.6300 lb H2/lb dry fu</td>
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<td>Kpxh2</td>
<td>20.0000 btu/hr-ft-F</td>
<td>XH2oil</td>
<td>1.2700 lb H2/lb oil</td>
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<td>XH2oil</td>
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<td>nphx2</td>
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<td>XH2oil</td>
<td>1.2700 lb H2/lb oil</td>
</tr>
<tr>
<td>nuh2</td>
<td>3.0000 no. of U's</td>
<td>X02</td>
<td>0.4050 lb O2/lb dry fu</td>
</tr>
<tr>
<td>nuh2</td>
<td>3.0000 no. of U's</td>
<td>X02oil</td>
<td>0.0015 lb O2/lb oil</td>
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<tr>
<td>ODhx1</td>
<td>0.7500 inches</td>
<td>YdscD</td>
<td>0.0003</td>
</tr>
<tr>
<td>ODhx1</td>
<td>1.0000 inches</td>
<td>YdscD</td>
<td>0.0003</td>
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<tr>
<td>Pa01</td>
<td>14.6960 psia</td>
<td>Ydsw</td>
<td>0.0111</td>
</tr>
<tr>
<td>Pa04</td>
<td>26.0000 in. h2o</td>
<td>Ydsn2</td>
<td>0.2072</td>
</tr>
<tr>
<td>Pcc22</td>
<td>7.0000 in. h2o</td>
<td>Ydsn2</td>
<td>0.7722</td>
</tr>
<tr>
<td>Pds</td>
<td>14.6960 psia</td>
<td>Ydsn2</td>
<td>0.7722</td>
</tr>
</tbody>
</table>

--- Additional Constants ---

<p>| HHV       | 8880.0000 btu/lb       | WRscw  | 10.0000 kW |
| HHVoil    | 3194.4050 btu/lb       | WRCml  | 0.0000 kW |
| IDhx1     | 0.5000 inches          | Xash   | 0.0050 lb ash/lb dry f |
| IDhx2     | 0.8100 inches          | Xash01 | 0.0000 lb ash/lb dry oil |
| Kpxh1     | 20.0000 btu/hr-ft-F    | XH2    | 0.6300 lb H2/lb dry fu |
| Kpxh2     | 20.0000 btu/hr-ft-F    | XH2oil | 1.2700 lb H2/lb oil |
| LHVoil    | 2631.7500 btu/lb       | XH2oil | 1.2700 lb H2/lb oil |
| MRC16m    | 999999.0000 lb/s       | XH2oil | 1.2700 lb H2/lb oil |
| MRC23m    | 1.0000 lb/s           | XH2oil | 1.2700 lb H2/lb oil |
| MRst      | 38.7500 lb/s          | XH2oil | 1.2700 lb H2/lb oil |
| nphx1     | 60.0000 pipes normal   | XH2oil | 1.2700 lb H2/lb oil |
| nphx2     | 72.0000 pipes normal   | XH2oil | 1.2700 lb H2/lb oil |
| nphx2     | 60.0000 pipes parallel | XH2oil | 1.2700 lb H2/lb oil |
| nphx2     | 60.0000 pipes parallel | XH2oil | 1.2700 lb H2/lb oil |
| nuh2      | 3.0000 no. of U's      | X02    | 0.4050 lb O2/lb dry fu |
| nuh2      | 3.0000 no. of U's      | X02oil | 0.0015 lb O2/lb oil |
| ODhx1     | 0.7500 inches          | YdscD  | 0.0003 |
| ODhx1     | 1.0000 inches          | YdscD  | 0.0003 |
| Pa01      | 14.6960 psia           | Ydsw   | 0.0111 |
| Pa04      | 26.0000 in. h2o        | Ydsn2  | 0.2072 |
| Pcc22     | 7.0000 in. h2o         | Ydsn2  | 0.7722 |
| Pds       | 14.6960 psia           | Ydsn2  | 0.7722 |</p>
<table>
<thead>
<tr>
<th>Variables</th>
<th>Value</th>
<th>Variables</th>
<th>Value</th>
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</thead>
<tbody>
<tr>
<td>MRc16</td>
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<td>Ps36</td>
<td>640.0000 psia</td>
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<tr>
<td>MRc25</td>
<td>41.5666 lb/s</td>
<td>Tc26</td>
<td>350.0000 F</td>
</tr>
<tr>
<td>MRs31</td>
<td>9.4583 lb/s</td>
<td>Ts36</td>
<td>700.8911 F</td>
</tr>
<tr>
<td>Pa031</td>
<td>129.8465 psia</td>
<td>Ts39</td>
<td>115.2230 F</td>
</tr>
<tr>
<td>Pc211</td>
<td>-4.5735 in. h2o</td>
<td>Ts402</td>
<td>115.2230 F</td>
</tr>
<tr>
<td>Pc29</td>
<td>0.7000 in. h2o</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Description</td>
<td>Value</td>
<td>Unit</td>
<td></td>
</tr>
<tr>
<td>-------------------------------------</td>
<td>--------</td>
<td>------------------</td>
<td></td>
</tr>
<tr>
<td>Ablr 17250.2900 square feet</td>
<td>PC27</td>
<td>0.7000 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td></td>
</tr>
<tr>
<td>Ahx1 301.3970 square feet</td>
<td>PE62</td>
<td>0.7000 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
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</tr>
<tr>
<td>Ahx2 6844.1230 square feet</td>
<td>PE65</td>
<td>0.7000 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td></td>
</tr>
<tr>
<td>CFMn1 5080.7900 cu ft/min</td>
<td>PS38</td>
<td>626.2251 psia</td>
<td></td>
</tr>
<tr>
<td>CFMn3 7122.3800 cu ft/min</td>
<td>OCmbl</td>
<td>19950.6100 btu/s</td>
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<tr>
<td>CFMn4 19186.0000 cu ft/min</td>
<td>OCnd</td>
<td>0.0000 btu/s</td>
<td></td>
</tr>
<tr>
<td>CFMn5 5080.7900 cu ft/min</td>
<td>TA021</td>
<td>595.5196 F</td>
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</tr>
<tr>
<td>dPfn3 0.0000 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>TA022</td>
<td>595.5196 F</td>
<td></td>
</tr>
<tr>
<td>dPfn4 10.2735 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>TA04</td>
<td>823.5859 F</td>
<td></td>
</tr>
<tr>
<td>dPfn5 14.3000 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>TC111</td>
<td>2141.7540 F</td>
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</tr>
<tr>
<td>dPfn6 0.0000 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>TC13</td>
<td>1616.3790 F</td>
<td></td>
</tr>
<tr>
<td>dPfn1 0.0000 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>TC14</td>
<td>1118.5790 F</td>
<td></td>
</tr>
<tr>
<td>dPfn1w 13.7749 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>TP02</td>
<td>948.6503 F</td>
<td></td>
</tr>
<tr>
<td>dPfn2a 2.5719 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>Vhx1g</td>
<td>53.2458 ft/s</td>
<td></td>
</tr>
<tr>
<td>dPfn2g 6.4930 in. h&lt;sub&gt;2&lt;/sub&gt;o</td>
<td>Vhx2g</td>
<td>68.8690 ft/s</td>
<td></td>
</tr>
<tr>
<td>EFHg 27.5048 % gross, HHV</td>
<td>WRaux</td>
<td>392.2557 kW</td>
<td></td>
</tr>
<tr>
<td>EFHin 25.6594 % net, HHV</td>
<td>WRcmp</td>
<td>5310.0850 kW</td>
<td></td>
</tr>
<tr>
<td>EFILg 29.5058 % gross, LHV1</td>
<td>WRdry</td>
<td>0.0000 kW</td>
<td></td>
</tr>
<tr>
<td>EFFL3g 30.3042 % gross, LHV3</td>
<td>WRfn1</td>
<td>0.0000 kW</td>
<td></td>
</tr>
<tr>
<td>EFtrm 80.4450 % net, EFtrm</td>
<td>WRfn3</td>
<td>136.5595 kW</td>
<td></td>
</tr>
<tr>
<td>IRBr1 1766.8820 kW</td>
<td>WRfn4</td>
<td>51.2039 kW</td>
<td></td>
</tr>
<tr>
<td>IRCmb 7802.0500 kW</td>
<td>WRfn5</td>
<td>0.0000 kW</td>
<td></td>
</tr>
<tr>
<td>IRCmp 665.6182 kW</td>
<td>WRgn1</td>
<td>2825.3120 kW</td>
<td></td>
</tr>
<tr>
<td>IRCnd 0.0000 kW</td>
<td>WRgn2</td>
<td>3652.7890 kW</td>
<td></td>
</tr>
<tr>
<td>IRDrv 0.0000 kW</td>
<td>WRgrs</td>
<td>6478.1010 kW</td>
<td></td>
</tr>
<tr>
<td>IRfn1 0.0000 kW</td>
<td>WRgtc</td>
<td>8391.7950 kW</td>
<td></td>
</tr>
<tr>
<td>IRFn3 50.5270 kW</td>
<td>WRgpc</td>
<td>107.6704 kW</td>
<td></td>
</tr>
<tr>
<td>IRF4n 18.9454 kW</td>
<td>WRmp</td>
<td>29.8217 kW</td>
<td></td>
</tr>
<tr>
<td>TRFn5 0.0000 kW</td>
<td>WRstb1</td>
<td>3984.2810 kW (1st stage)</td>
<td></td>
</tr>
<tr>
<td>IRRgn1 256.3982 kW</td>
<td>WRstb2</td>
<td>0.0000 kW (2nd stage)</td>
<td></td>
</tr>
<tr>
<td>IRRgn2 331.4922 kW</td>
<td>Xs39</td>
<td>0.9380 lb/lb quality</td>
<td></td>
</tr>
<tr>
<td>IRRgb 468.0253 kW</td>
<td>Xs402</td>
<td>0.0000 lb/lb quality</td>
<td></td>
</tr>
<tr>
<td>IRRhx1 188.5879 kW</td>
<td>YA01CD</td>
<td>0.0003 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>IRRhx2 703.3701 kW</td>
<td>YA01W</td>
<td>0.0111 mol fraction H20</td>
<td></td>
</tr>
<tr>
<td>IRRj1 0.0000 kW</td>
<td>YA012</td>
<td>0.2072 mol fraction Q2</td>
<td></td>
</tr>
<tr>
<td>IRRmcd 963.4374 kW (mix ex cmb)</td>
<td>YA01N2</td>
<td>0.7722 mol fraction N2</td>
<td></td>
</tr>
<tr>
<td>IRpexd 0.0000 kW (mix in dry)</td>
<td>YA01Ar</td>
<td>0.0092 mol fraction Ar</td>
<td></td>
</tr>
<tr>
<td>IRpm 9.5400 kW</td>
<td>YA01CD</td>
<td>0.0009 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>IRst1 898.6559 kW (1st stage)</td>
<td>YA02CD</td>
<td>0.0003 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>IRst2 0.0000 kW (2nd stage)</td>
<td>YA02W</td>
<td>0.0111 mol fraction H20</td>
<td></td>
</tr>
<tr>
<td>IRRtrm 1051.5960 kW</td>
<td>YA02Q2</td>
<td>0.2072 mol fraction Q2</td>
<td></td>
</tr>
<tr>
<td>lendry 0.0000 feet</td>
<td>YA02N2</td>
<td>0.7722 mol fraction N2</td>
<td></td>
</tr>
<tr>
<td>LHV1 8203.2200 BTU/lb -h2comb</td>
<td>YA02Ar</td>
<td>0.0092 mol fraction Ar</td>
<td></td>
</tr>
<tr>
<td>LHV2 7987.0940 BTU/lb -h2wood</td>
<td>YA02CD</td>
<td>0.0000 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>Ma01 28.8445 lb/mol</td>
<td>YA03CD</td>
<td>0.0095 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>Ma02 0.0000 lb/mol</td>
<td>YA03W</td>
<td>0.0167 mol fraction H20</td>
<td></td>
</tr>
<tr>
<td>Ma03 28.8935 lb/mol</td>
<td>YA03Q2</td>
<td>0.1965 mol fraction Q2</td>
<td></td>
</tr>
<tr>
<td>MA021 36.6792 lb/s</td>
<td>YA03N2</td>
<td>0.7681 mol fraction N2</td>
<td></td>
</tr>
<tr>
<td>MA022 36.6792 lb/s</td>
<td>YA03Ar</td>
<td>0.0091 mol fraction Ar</td>
<td></td>
</tr>
<tr>
<td>MA032 37.4130 lb/s</td>
<td>YA03CD</td>
<td>0.0000 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>MA04 38.9413 lb/s</td>
<td>YA04CD</td>
<td>0.0092 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>MA06 1.5283 lb/s</td>
<td>YA04W</td>
<td>0.165 mol fraction H20</td>
<td></td>
</tr>
<tr>
<td>HA051 2.7465 lb/s</td>
<td>YA04Q2</td>
<td>0.1969 mol fraction Q2</td>
<td></td>
</tr>
<tr>
<td>HA052 2.7465 lb/s</td>
<td>YA04N2</td>
<td>0.7682 mol fraction N2</td>
<td></td>
</tr>
<tr>
<td>MAdwd 2.2061 lb/s</td>
<td>YA04Ar</td>
<td>0.0091 mol fraction Ar</td>
<td></td>
</tr>
<tr>
<td>MCl11 41.5666 lb/s</td>
<td>YA04CD</td>
<td>0.0000 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>MCl12 57.5400 lb/s</td>
<td>YC11CD</td>
<td>0.0754 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>MCl12 57.5400 lb/s</td>
<td>YC11W</td>
<td>0.0803 mol fraction H20</td>
<td></td>
</tr>
<tr>
<td>MCl12 57.5400 lb/s</td>
<td>YC11Q2</td>
<td>0.1132 mol fraction Q2</td>
<td></td>
</tr>
<tr>
<td>MCr26 41.5666 lb/s</td>
<td>YC11N2</td>
<td>0.7218 mol fraction N2</td>
<td></td>
</tr>
<tr>
<td>MRdug 0.2722 lb/s</td>
<td>YC11Ar</td>
<td>0.0086 mol fraction Ar</td>
<td></td>
</tr>
<tr>
<td>MRe6l 41.5666 lb/s</td>
<td>YC11CO</td>
<td>0.0007 mol fraction C02</td>
<td></td>
</tr>
<tr>
<td>MRG</td>
<td>0.7338 lb/s</td>
<td>ye61CD</td>
<td>0.0754 mol frac CO2</td>
</tr>
<tr>
<td>-------</td>
<td>-------------</td>
<td>--------</td>
<td>---------------------</td>
</tr>
<tr>
<td>MRs402</td>
<td>9.4584 lb/s</td>
<td>ye61W</td>
<td>0.0803 mol frac H2O</td>
</tr>
<tr>
<td>MRw4g</td>
<td>0.3262 lb/s</td>
<td>ye6102</td>
<td>0.1132 mol frac O2</td>
</tr>
<tr>
<td>Pa021</td>
<td>132.4185 psia</td>
<td>ye61N2</td>
<td>0.7218 mol frac N2</td>
</tr>
<tr>
<td>Pa022</td>
<td>132.4185 psia</td>
<td>ye61Ar</td>
<td>0.0086 mol frac Ar</td>
</tr>
<tr>
<td>Pa032</td>
<td>128.2466 psia</td>
<td>ye61CO</td>
<td>0.0007 mol frac CO</td>
</tr>
<tr>
<td>Pc111</td>
<td>15.0000 in. h2o</td>
<td>yPGAr</td>
<td>0.0129 mol Ar/mol wd</td>
</tr>
<tr>
<td>Pc112</td>
<td>15.0000 in. h2o</td>
<td>yFGCH4</td>
<td>0.0474 mol CH4/mol wd</td>
</tr>
<tr>
<td>Pc113</td>
<td>10.0000 in. h2o</td>
<td>yFGCO</td>
<td>0.8092 mol CO/mol wd</td>
</tr>
<tr>
<td>Pc13</td>
<td>9.9195 in. h2o</td>
<td>yPGCO2</td>
<td>1.1434 mol CO2/mol wd</td>
</tr>
<tr>
<td>Pc14</td>
<td>3.4265 in. h2o</td>
<td>yPGH2</td>
<td>0.5308 mol H2/mol wd</td>
</tr>
<tr>
<td>Pc212</td>
<td>-9.5735 in. h2o</td>
<td>yPGH20</td>
<td>0.0000 mol H2O/mol wd</td>
</tr>
<tr>
<td>Pc232</td>
<td>15.0000 in. h2o</td>
<td>yFGN2</td>
<td>1.0812 mol N2/mol wd</td>
</tr>
<tr>
<td>Pc26</td>
<td>0.7000 in. h2o</td>
<td></td>
<td></td>
</tr>
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</table>
SUBROUTINE GASIFY(wb51,Tds,Xdirt,Xash,Te01,yrCO2,yrH2O,yrO2,  
 yrN2,yrAr,yrCO,Pg,Eqrat,radloss,xCC0,Te032,yCO2in,yH20in,  
 yO2in,yN2in,yArin,yCOin,Pair,dBurn,Te031,Mra022,mgfg,PGCO,  
 PGCO2,PGH2,PGH2O,PGCH4,PGN2,PGAr,Tpg02,yeCO2,yeH2O,  
 yeO2,yeN2,yeAr,yeCO,Mra032,Pgex,MPG,MRwg,MRdwg,Etfrm,  
 print,IRRtrm,WRpgc)

C
C THIS SUBROUTINE DETERMINES THE TEMPERATURE AND COMPOSITION OF  
C THE PRODUCT GAS EXITING AN UP-DRAFT GASIFIER. THE PROCESS USED  
C IN THIS SUBROUTINE IS AS FOLLOWS: 1) DETERMINE THE ADIABATIC  
C FLAME TEMPERATURE IN THE COMBUSTION ZONE OF THE GASIFIER BY  
C ITERATION ON THE HEAT BALANCE OF THE COMBUSTION OF WOOD. 2) USE  
C THE FLAME TEMPERATURE OBTAINED AS THE PYROLYSIS TEMPERATURE AND  
C DETERMINE THE COMPOSITION OF THE PRODUCT GAS PER 1 MOLE OF DRY  
C WOOD (ASSUMED) USING A NEWTON-RAPHSON SCHEME FOR A SYSTEM OF 5  
C NON-LINEAR SIMULTANEOUS EQUATIONS (3 FOR ELEMENT BALANCE, 1 FOR  
C HEAT BALANCE, 1 FOR CHEMICAL EQUILIBRIUM). 3) DETERMINE THE  
C EXIT TEMPERATURE OF THE PRODUCT GAS AFTER DRYING THE EQUIVALENT  
C WOOD YIELDING 1 MOLE OF DRY WOOD IN THE PYROLYSIS ZONE.

LIST OF VARIABLES

wb51-MOISTURE CONTENT OF THE INLET WOOD ON A WET BASIS.
Tds-INLET WOOD TEMPERATURE.
Xdirt-MASS RATIO OF DIRT IN INLET WOOD.
Xash-MASS RATIO OF ASH IN INLET WOOD.
Te01-AMBIENT AIR TEMPERATURE.
yrCO2-MOLE FRACTION OF CO2 IN AMBIENT AIR ENTERING GASIFIER.
yrH2O-MOLE FRACTION OF H2O IN AMBIENT AIR ENTERING GASIFIER.
yrO2-MOLE FRACTION OF O2 IN AMBIENT AIR ENTERING GASIFIER.
yrN2-MOLE FRACTION OF N2 IN AMBIENT AIR ENTERING GASIFIER.
yrAr-MOLE FRACTION OF Ar IN AMBIENT AIR ENTERING GASIFIER.
yrCO-MOLE FRACTION OF CO IN AMBIENT AIR ENTERING GASIFIER.
Eqrat-MASS RATIO OF ENTERING AIR TO ENTERING WOOD.
PGCO-MOLES OF CO FORMED IN GASIFIER PER ONE MOLE OF WOOD.
PGCO2-MOLES OF CO2 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
PGH2-MOLES OF H2 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
PGH2O-MOLES OF H2O FORMED IN GASIFIER PER ONE MOLE OF WOOD.
PGCH4-MOLES OF CH4 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
Tpg-TEMPERATURE OF PRODUCT GASES EXITING GASIFIER.

implicit real(a-z)
integer i,j,k,jj,kpl,ii,ipl
logical print2
dimension z(5,6),x(5)

C DETERMINE MOLES OF ELEMENTS ENTERING GASIFIER IN UNDERFIRE  
C COMBUSTION AIR.
Moles of dry wood in gasifier that burns \( XX \)

\( XX = EQrat \)

Heat balance equation iterated on to determine the adiabatic flame temperature in the combustion zone.

\[
\text{delta}=1.0 \\
Z1=2.98 \\
Z2=\left( (Ta01-32.0) \times (5.0/9.0) + 273.15 \right) / 100.0 \\
call CSUBP(Z1,Z2,dhO2,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr) \\
LHS=((XX)*(115000.13))+(O2ing*(dhO2))+(N2ing*(dhN2)) \\
\]

Guess an AFT and then calculate the right hand side of the heat balance (products) and compare with left hand side. Iterate on the AFT until the two sides are equal.

\[
Z2=3.00 \\
100 \text{ call CSUBP(Z1,Z2,dhO2,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr) } \\
twd=\left( (Z2 \times 100.0) - 273.15 \right) \times (9.0/5.0) + 32.0 \\
tzl=77.0 \\
cpwd=0.266 + (0.00322 \times (twd + tzl - 64.0)) \\
cpwd=4.1858 * cpwd + 22.84 \\
RHS=((XX)*(-39322.0 + dhCO2)) + ((7*XX)*(-241827.0 + dhH2O)) + ((N2ing)*(dhN2)) + ((1.0 - XX) \times cpwd * (Z2 - Z1) \times 100)) \\
check=RHS-LHS \\
if (check < 0.0) then \\
Z2=Z2+delta \\
goto 100 \\
else if (delta .ge. 0.05) then \\
Z2=Z2-delta \\
delta=delta/10.0 \\
Z2=Z2+delta \\
goto 100 \\
else \\
AFT=Z2*100.0 \\
endif \\
NOW assume that the AFT is the temperature that the water shift reaction occurs at in the reduction zone. Knowing this, the equilibrium constant of that reaction can be determined. This is done by using a curve fit of the plot of \( Keq \) versus temp.

\[
\text{if (AFT .lt. 1000.0) then } \\
\text{Keq=55.2046-(1.0443*AFT)+(5.06016E-05*(AFT**2.0))} \\
\text{else if (AFT .lt. 1200.0) then } \\
\text{Keq=17.7387-(.0268641*AFT)+(-.05697E-05*(AFT**2.0))} \\
\]
else if (AFT .lt. 1400.0) then  
  Keq=0.75579-(.011979*AFT)+((3.75273E-06)*(AFT**2.0))  
else  
  Keq=3.78197-((3.91905E-03) AFT)+((1.08963E-05)(AFT**2.0))  
endif

C
C
C INITIAL GUESS AT THE COMPOSITION OF THE PRODUCT GAS
C
  if (E0rat .le. .75) then  
    c=.68  
    d=.05  
    e=.9  
    f=1  
    g=.05  
  else if (E0rat .le. .95) then  
    c=.68  
    d=.05  
    e=1  
    f=.1  
    g=.05  
  else  
    c=.68  
    d=.05  
    e=.01  
    f=1  
    g=.05  
endif

  coefcdH2  
  coefd=(241827.0+dhH20)  
  coefe=110529.0+dhCO)  
  coeff=(-393522.0+dhCO2)  
  coefg=(-74873.0+dhCH4)  

200  
  z(1,1)=0.0  
  z(1,2)=0.0  
  z(1,3)=1.0  
  z(1,4)=1.0  
  z(1,5)=1.0  
  z(2,1)=2.0  
  z(2,2)=2.0  
  z(2,3)=0.0  
  z(2,4)=0.0  
  z(2,5)=4.0  
  z(3,1)=0.0  
  z(3,2)=1.0  
  z(3,3)=1.0  
  z(3,4)=2.0  
  z(3,5)=0.0  
  z(4,1)=coefc*(-1.0)  
  z(4,2)=coefd*(-1.0)  
  z(4,3)=coefe*(-1.0)
\[
\begin{align*}
    &z(4,4) = \text{coeff}(-1.0) \\
    &z(4,5) = \text{coeffg*(-1.0)} \\
    &z(5,1) = f/(ed) \\
    &z(5,2) = ((c*f)/(e*(d+2.0)))*(-1.0) \\
    &z(5,3) = ((c*f)/(e+2.0+d))*(-1.0) \\
    &z(5,4) = c/(ed) \\
    &z(5,5) = 0.0 \\
    &z(1,6) = (e+f+g-(1.0))*(-1.0) \\
    &z(2,6) = (2.0c)+(2.0d)+(4.0g)-(1.4)*(-1.0) \\
    &z(3,6) = (d+e+(2.0f)-(0.59+(2.002ing)))*(-1.0) \\
    &z(4,6) = (((-1.0)*\text{coeffc*c})+((-1.0)*\text{coeffd}d)+((-1.0)*\text{coffe}e)+ \\
    &> ((-1.0)*\text{coeff*f}+((-1.0)*\text{coeffg}g)-(103977.47 \\
    &> +(N2ing*dN2)))*(-1.0) \\
    &z(5,6) = (((c*f)/(e+d))-\text{Keq})*(-1.0)
\end{align*}
\]

C

WITH MATRIX FILLED, A LINEAR EQUATION SOLVER USING PARTIAL
C
PIVOTING IS NEXT EMPLOYED TO DETERMINE THE SOLUTION SET \((x(1-5))\)
C
WHICH IS THE AMOUNT EACH INITIAL GUESS SHOULD BE INCREMENTED BY.
C

\begin{verbatim}
do 270 k=1,4
   jj=k
   big=abs(z(k,k))
   kpl=k+1
   do 220 i=kpl,5
      ab=abs(z(i,k))
      if (big-ab) 210,220,220
    210  big=ab
      jj=i
   220  continue
   do 230 j=k,6
      temp=z(jj,j)
      z(jj,j)=z(k,j)
      z(k,j)=temp
   230  do 260 i=kpl,5
      quot=z(i,k)/z(k,k)
      do 260 j=kpl,6
         z(i,j)=z(i,j)-quot*z(k,j)
      260  do 270 i=kpl,5
      z(i,k)=0.0
   270 C
end
\end{verbatim}
\[ x(5) = z(5,6) / z(5,5) \]

C remainder of back substitution process

\[
\begin{align*}
& \text{do 290 } i = 1, 4 \\
& \quad \text{sum} = 0.0 \\
& \quad i = 5 - i i \\
& \quad ip1 = i + 1 \\
& \quad \text{do 280 } j = ip1, 5 \\
& \quad \quad \text{sum} = \text{sum} + z(i,j) * x(j) \\
& \quad \quad x(i) = (z(i,6) - \text{sum}) / z(i,i)
\end{align*}
\]

C NOW CHECK FOR CONVERGENCE OF NEWTON-RAPHSON METHOD BY TESTING TO SEE IF ALL INCREMENT ADDITIONS (x(1-5)) ARE LESS THAN .001

\[
\begin{align*}
& \text{if (abs(x(1)) .gt. .001) then} \\
& \quad \text{goto 200} \\
& \quad \text{else if (abs(x(2)) .gt. .001) then} \\
& \quad \quad \text{goto 200} \\
& \quad \quad \text{else if (abs(x(3)) .gt. .001) then} \\
& \quad \quad \quad \text{goto 200} \\
& \quad \quad \quad \text{else if (abs(x(4)) .gt. .001) then} \\
& \quad \quad \quad \quad \text{goto 200} \\
& \quad \quad \quad \quad \text{else if (abs(x(5)) .gt. .001) then} \\
& \quad \quad \quad \quad \quad \text{goto 200} \\
& \quad \quad \quad \quad \quad \text{else} \\
& \quad \quad \quad \quad \quad \quad \text{yPGH2} = c \\
& \quad \quad \quad \quad \quad \quad \text{yPGH2O} = d + H2O\text{ing} \\
& \quad \quad \quad \quad \quad \quad \text{yPGCO} = e \\
& \quad \quad \quad \quad \quad \quad \text{yPGCO2} = f \\
& \quad \quad \quad \quad \quad \quad \text{yPGCH4} = g \\
& \quad \quad \quad \quad \quad \quad \text{yPGN2= N2\text{ing} } \\
& \quad \quad \quad \quad \quad \quad \text{yPGAr= Ar\text{ing} }
\end{align*}
\]

C DETERMINE THE TEMPERATURE DECREASE OF THE PRODUCT GAS BY EVAPORATING THE MOISTURE IN THE INCOMING WOOD. THIS IS DONE BY DETERMINING THE ENERGY REQUIRED TO EVAPORATE THE MOISTURE IN THE WOOD AT THE GIVEN PRESSURE THEN PERFORMING AN ENERGY BALANCE.

\[
\begin{align*}
& \text{Mdirt} = Xdirt * 22.84 \\
& \text{press} = P_g \\
& \text{Tgas} = Tsat WP(press) \\
& \text{Tgas} = (((Tgas - 32.0) * (5.0/9.0)) + 273.15 \\
& \text{Tamb} = (((Ta01 - 32.0) * (5.0/9.0)) + 273.15 \\
& \text{Mtot} = 22.84 / (1.0 - Wb51)
\end{align*}
\]
THE TOTAL ENERGY REQUIRED TO EVAPORATE THE WATER IS CALCULATED.

```
Hfgs1=Hfg*2.326
Qdir2=mdirt*.8374*(Tgas-Tamb)
Qdry=(((4.187*(Tgas-Tamb))+Hfgs1)*Mwat+((Tgas-Tamb)*3.06+22.84)
```

CALCULATE THE TEMPERATURE DECREASE OF THE PRODUCT GAS (dT), AND THE SUBSEQUENT EXIT PRODUCT GAS TEMPERATURE (Tpg02).

```
dT2=((Qdry+Qdir2)/((yPGAr*20.785)+(yPGCO2*37.0392)+(yPGCO*29.1564)+(yPGH2O*28.4182)+(yPGCH4*38.0592)+(yPGN2*29.1648)+
  (yPGH2*33.7014)))
Tpg02=AFT-dT2
```

DETERMINE EQUILIBRIUM TEMPERATURE AT WHICH THE WATER AND GAS EXIT THE GASIFIER

```
delta=1.0
cons=Tpg02/100.0
Z2=3.73
1313 Z1=Tgas/100.0
    call CSUBP(Z1,Z2,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr)
    LS=(Mwat/100.0)*dhH2O
    Z1=Z2
    call CSUBP(Z1,cons,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr)
    RS=yPGN2*dhN2+yPGAr*dhAr+yPGH20*dhH2O+yPGCO2*dhCO2+yPGCO*dhCO+
            yPGCH4*dhCH4+yPGH2*dhH2
    checc=RS-LS
    if(checc .gt. 0.0) then
        Z2=Z2+delta
        goto 1313
    else if (delta .ge. .005) then
        Z2=Z2+delta
        delta=delta/10.0
        Z2=Z2-delta
        goto 1313
    else
        Tpg02-Z2+100.0
    endif
```

NOW ADD THE WATER EVAPORATED TO THE MOLES OF WATER PRODUCED IN PYROLYSIS TO YIELD FINAL WATER AMOUNT.

```
Pgwat=yPGH20
    Pgwat=yPGH20+(Mwat/18.0)
    eng=(yPGCH4*88635.0)+(yPGH2*285638.0)+(yPGCO*282989.0)
    mas=(yPGCO*28.0)+(yPGCO2*44.0)+(yPGH20*18.0)+(yPGH2*2.0)+
            (yPGH4*18.0)+(yPGN2*28.0)+(yPGAr*39.95)
    HHVpgb=eng/mas
```

ASSUME THE PRODUCT GAS IS COOLED IN A CYCLONE/SCRUBBER TO
C CONDENSE OUT THE WATER VAPOR AND TAR AND OILS.
C
mas=mas-(yPGH2O*18.0)
HHVpga=eng/mas
wmass=yPGH2O*18.0
yPGH2O=0.0
T1=298.0

C ENTER THE PRODUCT GAS COMPRESSOR AND DETERMINE EXIT GAS TEMPERATURE
C
HFH2O=(-241827.0)
HFCO2=(-393522.0)
HFCO=(-110529.0)
HFCH4=(-74873.0)
Pfex=Pair-dPburn
Nair=MRa022/(28.97*2.2046244)
NO2=yO2in*Nair
NN2=yN2in*Nair
NH20=yH20in*Nair
NA=Arin*Nair
NCO2=yCO2in*Nair
NCO=yCOin*Nair
P1=(Pg-1.0)*6.894757
P2=(Pair+1.0)*6.894757
H2T2=T1+((P2/P1)**(.290277))
H2OT2=T1+((P2/P1)**(.246420))
COT2=T1+((P2/P1)**(.285714))
CO2T2=T1+((P2/P1)**(.224205))
CH4T2=T1+((P2/P1)**(.230177))
ArT2=T1+((P2/P1)**(.230177))
N2T2=T1+((P2/P1)**(.285714))

C RECIRCULATE THE CLEAN, COMPRESSED, PRODUCT GAS THROUGH A
C RECUPERATOR TO RECLAIM SOME OF THE ENERGY ENTRAINED IN THE DIRTY
C PRODUCT GAS STREAM EXITING THE GASIFIER. (ASSUMES A TERMINAL
C TEMPERATURE DIFFERENCE OF 5 K).
C
if (Tpg02 .gt. 550.0) then
  Ttin=Tpg02-5.0
else
  Ttin=550.0
endif

C ENTER TRIMBURNER PERFORM HEAT BALANCE TO DETERMINE RATE OF PRODUCT
C GAS REQUIRED FOR SPECIFIED HEATING OF AIR STREAM.
C
Z2=Ttin/100.0
Z1=2.98
CALL CSU8P(T1,Z2,do2,dH2O,dH2O2,dH2O2,dHCO2,dHCO2,dH2,dH2,dH2,dCH4,dAr)
dhP6CO=dhCO
dh6CO2=dhCO2
dhP6H2=dhH2
dh6H2O=dhH2O
dhP6CH4=dhCH4
dhP6Ar=dhAr
\[ Z2 = \left(\frac{(T_{031} - 32.0) \times (5.0 / 9.0) + 273.0}{100.0}\right) \]

CALL CSUBP(Z1, Z2, dhO2, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)

NdhCO2 = dhCO2
NdhH2O = dhH2O
NdhO2 = dhO2
NdhN2 = dhN2
NdhAr = dhAr
NdhCO = dhCO

\[ Z2 = \left(\frac{(T_{032} - 32.0) \times (5.0 / 9.0) + 273.0}{100.0}\right) \]

CALL CSUBP(Z1, Z2, dhO2, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)

OdhCO2 = dhCO2
OdhH2O = dhH2O
OdhO2 = dhO2
OdhN2 = dhN2
OdhAr = dhAr
OdhCO = dhCO

KPH2O = NhH2O
KPCO2 = (XCCO) \times (NCO + NC02)
KPO2 = (-XCCO) \times (NCO + NC02)
KPO2 = NO2 + NH2O / 2.0 + NC02 + NH2O / 2.0 - (NCO \times (1.0 - XCCO))

\[ X_{mole} = \frac{(1.0 - \text{radloss}) \times KREAC - KPROD}{VPROD - (1.0 - \text{radloss}) \times VREAC} \]

CALCULATE MASS RATE OF PRODUCT GAS REQUIRED

\[ mP6H2 = yP6H2 \times X_{mole} \times 2.0 \]
\[ mP6H2O = yP6H2O \times X_{mole} \times 18.0 \]
\[ mP6CO = yP6CO \times X_{mole} \times 28.0 \]
\[ mP6CO2 = yP6CO2 \times X_{mole} \times 44.0 \]
\[ mP6CH4 = yP6CH4 \times X_{mole} \times 16.0 \]
\[ mP6N2 = yP6N2 \times X_{mole} \times 28.0 \]
\[ mP6Ar = yP6Ar \times X_{mole} \times 39.948 \]
MRP = mP6H2 + mP6H2O + mP6CO + mP6CO2 + mP6CH4 + mP6N2 + mP6Ar
MRP = 2.046244
CALCULATE PRODUCT GAS COMPRESSOR WORK REQUIRED FOR GIVEN MASS RATE

\[ W_{\text{rp}} = (H_2\text{w} \cdot m_{\text{H}_2}P_{\text{H}_2}) + (H_2\text{Ow} \cdot m_{\text{H}_2O}P_{\text{H}_2O}) + (\text{COw} \cdot m_{\text{CO}}P_{\text{CO}}) + (\text{CO}_2\text{w} \cdot m_{\text{CO}_2}P_{\text{CO}_2}) + \\
> (\text{CH}_4\text{w} \cdot m_{\text{CH}_4}P_{\text{CH}_4}) + (\text{Arw} \cdot m_{\text{Ar}}P_{\text{Ar}}) + (\text{N}_2\text{w} \cdot m_{\text{N}_2}P_{\text{N}_2}) \] / (0.825.9)

DETERMINE WOOD, AIR, MASS RATES IN SUB-SYSTEM

\[ \text{MR}_{\text{Ai}} = \text{MR}_{\text{P}} + \text{MR}_{\text{a}} \]

\[ \text{yyCO}_2 = (\text{UPCO}_2 \cdot \text{Xmole} \cdot \text{KPCO}_2) \]

\[ \text{yyCO} = (\text{UPCO} \cdot \text{Xmole} \cdot \text{KPCO}) \]

\[ \text{yyO}_2 = (\text{UPO}_2 \cdot \text{Xmole} \cdot \text{KPO}_2) \]

\[ \text{yyN}_2 = (\text{UN}_2 \cdot \text{Xmole} \cdot \text{KPN}_2) \]

\[ \text{yyH}_2\text{O} = (\text{UH}_2\text{O} \cdot \text{Xmole} \cdot \text{KPH}_2\text{O}) \]

\[ \text{yyAr} = (\text{UAr} \cdot \text{Xmole} \cdot \text{KPAr}) \]

\[ \text{yettot} = \text{yyCO}_2 + \text{yyCO} + \text{yyO}_2 + \text{yyN}_2 + \text{yyH}_2\text{O} + \text{yyAr} \]

\[ \text{yECO}_2 = \text{yyCO}_2 / \text{yettot} \]

\[ \text{yECO} = \text{yyCO} / \text{yettot} \]

\[ \text{yEO}_2 = \text{yyO}_2 / \text{yettot} \]

\[ \text{yEN}_2 = \text{yyN}_2 / \text{yettot} \]

\[ \text{yEH}_2\text{O} = \text{yyH}_2\text{O} / \text{yettot} \]

\[ \text{yEA} = \text{yyAr} / \text{yettot} \]

\[ \text{wfg} = (\text{yyCO}_2 \cdot 44.) + (\text{yyCO} \cdot 28.) + (\text{yyO}_2 \cdot 32.) + (\text{yyN}_2 \cdot 28.) + (\text{yyH}_2\text{O} \cdot 18.) \]

\[ \text{MR}_{\text{w}} = (\text{Xmole} \cdot 22.84) / (1.0 - \text{wSI}) \]

\[ \text{MR}_{\text{w}} = (\text{Xmole} \cdot 22.84) \]

\[ \text{MR}_{\text{w}} = \text{MR}_{\text{w}} \cdot 2.20462 \]

\[ \text{MR}_{\text{w}} = \text{MR}_{\text{w}} \cdot 2.20462 \]

CONVERT VARIABLES TO ENGLISH UNITS

\[ \text{HHV}_{\text{g}} = \text{HHV}_{\text{pg}} / 2.326 \]

\[ \text{HHV}_{\text{p}} = \text{HHV}_{\text{pg}} / 2.326 \]

\[ \text{AFT} = ((9.0/5.0) \cdot (\text{AFT} - 273.15)) + 32.0 \]

\[ \text{Tpg02} = ((9.0/5.0) \cdot (\text{Tpg02} - 273.15)) + 32.0 \]

\[ \text{WRpge} = \text{WRpge} / 1.055056 \]

\[ \text{TTin} = ((9.0/5.0) \cdot (\text{TTin} - 273.15)) + 32.0 \]

SECOND LAW ANALYSIS OF ENTIRE SYSTEM

\[ \text{MR}_{\text{u}} = (\text{O}_{2} \cdot \text{ing} \cdot 32.0 + \text{N}_2 \cdot \text{ing} \cdot 28.0 + \text{H}_2\text{O} \cdot \text{ing} \cdot 18.0 + \text{CO}_2 \cdot \text{ing} \cdot 44.0 + \text{C}_0 \cdot \text{ing} \cdot 28.0 \]

\[ \text{yAir} = \text{HgasT}((\text{Tds}, \text{yCO}_2, \text{yH}_2\text{O}, \text{yO}_2, \text{yN}_2, \text{yAr}, \text{yC}_0)) - \\
> (\text{Tds} < 45.967) \cdot \text{SgasT}(\text{Te}_3\text{b}, \text{Pair}, \text{Tds}, 14.696, \text{yCO}_2, \text{yH}_2\text{O}, \text{yO}_2, \text{yN}_2, \text{yAr}, \text{yC}_0) - (\text{Tds} > \\
> 45.967) \cdot \text{SgasT}(\text{Te}_3\text{b}, \text{Pf}, \text{Tds}, 14.696, \text{yCO}_2, \text{yH}_2\text{O}, \text{yO}_2, \text{yN}_2, \text{yAr}, \text{yC}_0) - (\text{Tds} > \\
> 45.967) \cdot \text{SgasT}(\text{Te}_3\text{b}, \text{Pf}, \text{Tds}, 14.696, \text{yCO}_2, \text{yH}_2\text{O}, \text{yO}_2, \text{yN}_2, \text{yAr}, \text{yC}_0) - (\text{Tds} > \\
> 45.967) \cdot \text{SgasT}(\text{Te}_3\text{b}, \text{Pf}, \text{Tds}, 14.696, \text{yCO}_2, \text{yH}_2\text{O}, \text{yO}_2, \text{yN}_2, \text{yAr}, \text{yC}_0) \]
WRITE RESULTS

TRIMBURNER SUB-SYSTEM

--GASIFIER SECTION--

WRITE(8,*), 'Gasifier Pressure = ',Pg,' psi'
WRITE(8,*), 'Gasifier Flame Temperature = ',AFT,' F'
WRITE(8,*), 'Product Gas Exit Temperature = ',Tpg02,' F'
WRITE(8,*), 'Inlet Wet Wood Mass Rate = ',MRwwg,' lbs/s'
WRITE(8,*), 'Inlet Dry Wood Mass Rate = ',MRdwg,' lbs/s'
WRITE(8,*), 'HHV of Gas before cyclone = ',HHVogb,' Btu/lb'
WRITE(8,*), 'HHV of Gas after cyclone = ',HHVpga,' Btu/lb'
WRITE(8,*), 'Mass Rate of Product Gas = ',MRPG,' lbs/s'
WRITE(8,*), 'Composition of Product Gas:

CO: ',yPGCO,' moles/mole of dry wood gasified'
CO2: ',yPGCO2,' moles/mole of dry wood gasified'
H2: ',yPGH2,' moles/mole of dry wood gasified'
H2O: ',yPGwat,' moles/mole of dry wood gasified'
CH4: ',yPGCH4,' moles/mole of dry wood gasified'
N2: ',yPGN2,' moles/mole of dry wood gasified'
Ar: ',yPGAr,' moles/mole of dry wood gasified'

--TRIMBURNER SECTION--

WRITE(8,*), 'Mass Rate of Air Entering = ',MRa022,' lbs/s'
WRITE(8,*), 'Mass Rate of Air Exiting = ',MRa032,' lbs/s'
WRITE(8,*), 'Product Gas Inlet Temp. = ',Ttin,' F'
WRITE(8,*), 'Inlet Air Temperature = ',Ta031,' F'
WRITE(8,*), 'Exit Air Temperature = ',Ta032,' F'
WRITE(8,*), 'Work Rate of Gas Compressor = ',WRpgc,' KW'
WRITE(8,*), 'Overall Sub-system Efficiency = ',Eftrm,' %'
WRITE(8,*), 'Overall Irreversibility = ',Irrtrm,' Kw'
WRITE(8,*), 'Second Law Efficiency = ',Seff,' %'
RETURN

END

SUBROUTINE CSUBP(Z1,Z2,dh02,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,>
dhAr)
SUBROUTINE GASIFY

C
C THIS SUBROUTINE DETERMINES THE TEMPERATURE AND COMPOSITION OF
C THE PRODUCT GAS EXITING AN UP-DRAFT GASIFIER. THE PROCESS USED
C IN THIS SUBROUTINE IS AS FOLLOWS: 1) DETERMINE THE ADIABATIC
C FLAME TEMPERATURE IN THE COMBUSTION ZONE OF THE GASIFIER BY
C ITERATION ON THE HEAT BALANCE OF THE COMBUSTION OF WOOD. 2) USE
C THE FLAME TEMPERATURE OBTAINED AS THE PYROLYSIS TEMPERATURE AND
C DETERMINE THE COMPOSITION OF THE PRODUCT GAS PER 1 MOLE OF DRY
C WOOD (ASSUMED) USING A NEWTON-RAPHSON SCHEME FOR A SYSTEM OF 5
C NON-LINEAR SIMULTANEOUS EQUATIONS (3 FOR ELEMENT BALANCE, 1 FOR
C HEAT BALANCE, 1 FOR CHEMICAL EQUILIBRIUM). 3) DETERMINE THE
C EXIT TEMPERATURE OF THE PRODUCT GAS AFTER DRYING THE EQUIVALENT
C WOOD YIELDING 1 MOLE OF DRY WOOD IN THE PYROLYSIS ZONE.
C
C LIST OF VARIABLES
C
C \( w_\text{bi} \)-MOISTURE CONTENT OF THE INLET WOOD ON A WET BASIS.
C \( T_\text{ds} \)-INLET WOOD TEMPERATURE.
C \( x_{\text{dirt}} \)-MASS RATIO OF DIRT IN INLET WOOD.
C \( x_{\text{ash}} \)-MASS RATIO OF ASH IN INLET WOOD.
C \( T_\text{a01} \)-AMBIENT AIR TEMPERATURE
C \( y_{\text{CO2}} \)-MOLE FRACTION OF CO2 IN AMBIENT AIR ENTERING GASIFIER.
C \( y_{\text{H2O}} \)-MOLE FRACTION OF H2O IN AMBIENT AIR ENTERING GASIFIER.
C \( y_{\text{O2}} \)-MOLE FRACTION OF O2 IN AMBIENT AIR ENTERING GASIFIER.
C \( y_{\text{N2}} \)-MOLE FRACTION OF N2 IN AMBIENT AIR ENTERING GASIFIER.
C \( y_{\text{Ar}} \)-MOLE FRACTION OF Ar IN AMBIENT AIR ENTERING GASIFIER.
C \( y_{\text{CO}} \)-MOLE FRACTION OF CO IN AMBIENT AIR ENTERING GASIFIER.
C \( E_{\text{Qrat}} \)-MASS RATIO OF ENTERING AIR TO ENTERING WOOD.
C \( y_{\text{PGCO}} \)-MOLES OF CO FORMED IN GASIFIER PER ONE MOLE OF WOOD.
C \( y_{\text{PGCO2}} \)-MOLES OF CO2 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
C \( y_{\text{PGH2}} \)-MOLES OF H2 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
C \( y_{\text{PGH2O}} \)-MOLES OF H2O FORMED IN GASIFIER PER ONE MOLE OF WOOD.
C \( y_{\text{PGCH4}} \)-MOLES OF CH4 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
C \( T_{\text{pg}} \)-TEMPERATURE OF PRODUCT GASES EXITING GASIFIER.
C
C implicit real(a-z)
C integer i,j,k,jj,kpl,i,ipl
C logical print2
C dimension z(5,6),x(5)
C
C DETERMINE MOLES OF ELEMENTS ENTERING GASIFIER IN UNDERFIRE
C COMBUSTION AIR.
MOLES OF DRY WOOD IN GASIFIER THAT BURNS(XX)

XX=Erat

HEAT BALANCE EQUATION ITERATED ON TO DETERMINE THE ADIABATIC FLAME TEMPERATURE IN THE COMBUSTION ZONE.

delta=1.0
Z1=2.98

DETERMINE INLET AIR TEMPERATURE AFTER COMPRESSION BY INLET COMPRESSOR

T1=((5.0/9.0)*((T01-32.0)))+273.0
P1=101.35
P2=((P+2.0)*6.894757
Ppgs=P+2.0
AIRT2=T1*(((P2/P1)**(.285714))
.Z2=AIRT2/100.0
.call CSUBP(Z2,XX,dh02,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr)
LHS=((XX)*(-115000.13))+((O21ng)*(dh02))+((N2Ing)*(dhN2))
> +(H2Oing*dhH2O)

call CSUBP(Z,XX,dh02,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr)
twd=((Z*100.0)-273.15)*(9.0/5.0)+32.0
tz1=77.0
cpwd=.268+(.000322*(twd+tz1-64.0))
cpwd=4.1868*cpwd+22.84
RHS=((XX)*(-393522.0+dhCO2))+((.7*XX)*(-241827.0+dhH20))
> +(N2Ing)*(dhN2)+((1.0-XX)*cpwd*(Z-Z1)*100.0)

cHECK=RHS-LHS
if (check .lt. 0.0) then
Z=Z+delta
goto 100
else if (delta .ge. .005) then
Z=Z-delta
delta=delta/10.0
Z=Z+delta
goto 100
else
AFT=Z*100.0
endif

NOW ASSUME THAT THE AFT IS THE TEMPERATURE THAT THE WATER SHIFT
REACTION OCCURS AT IN THE REDUCTION ZONE. KNOWING THIS, THE
EQUILIBRIUM CONSTANT OF THAT REACTION CAN BE DETERMINED. THIS
IS DONE BY USING A CURVE FIT OF THE PLOT OF Keq VERSUS TEMP.

\[
\begin{align*}
&\text{if } (AFT \lt 1000.0) \text{ then } \\
&\quad \text{Keq} = 55.2846 - (0.10443 \times AFT) + ((5.06016E-05) \times (AFT \times 2.0)) \\
&\text{else if } (AFT \rt 1200.0) \text{ then } \\
&\quad \text{Keq} = 17.37387 - (0.0268641 \times AFT) + ((1.05697E-05) \times (AFT \times 2.0)) \\
&\text{else if } (AFT \rt 1400.0) \text{ then } \\
&\quad \text{Keq} = 8.75579 - (0.0111979 \times AFT) + ((3.75273E-06) \times (AFT \times 2.0)) \\
&\text{else} \\
&\quad \text{Keq} = 3.78197 - (3.91905E-03 \times AFT) + ((1.08993E-06) \times (AFT \times 2.0)) \\
&\end{align*}
\]

THE NEWTON-RAPHSON METHOD WILL NOW BE USED. THE MATRIX IS LOADED
WITH THE M-MATRIX BEING THE PARTIAL DERIVATIVES OF THE S EQUATIONS
AND THE X-MATRIX BEING THE SOLUTION SET. THE SET IS USED TO
INCREMENT THE MOLE COMPOSITIONS.

INITIAL GUESS AT THE COMPOSITION OF THE PRODUCT GAS

\[
\begin{align*}
&\text{if } (EQt \leq 0.65) \text{ then } \\
&\quad c = 0.68 \\
&\quad d = 0.05 \\
&\quad e = 0.9 \\
&\quad f = 1 \\
&\quad g = 0.05 \\
&\text{else if } (EQt = 0.70) \text{ then } \\
&\quad yPGH2 = 0.0 \\
&\quad yPGH2O = 0.4 + H2O \text{ mng} \\
&\quad yPGCO = 0.0 \\
&\quad yPGCO2 = 0.85 \\
&\quad yPGCH4 = 0.15 \\
&\quad yPGN2 = N2 \text{ mng} \\
&\quad yPGAr = Ar \text{ mng} \\
&\quad \text{goto 507} \\
&\text{else} \\
&\quad yPGH2 = 0.0 \\
&\quad yPGH2O = 0.7 + H2O \text{ mng} \\
&\quad yPGCO = 0.0 \\
&\quad yPGCO2 = 1.0 \\
&\quad yPGCH4 = 0.0 \\
&\quad yPGN2 = N2 \text{ mng} \\
&\quad yPGAr = Ar \text{ mng} \\
&\quad \text{goto 507} \\
&\end{align*}
\]

coeff = dhH2
coeffd = (-104377.0 + dhH2O)
coeffe = (-110529.0 + dhCO)
coefff = (-383522.0 + dhCO2)
coeffg = (-74873.0 + dhCH4)
WITH MATRIX FILLED, A LINEAR EQUATION SOLVER USING PARTIAL
PIVOTING IS NEXT EMPLOYED TO DETERMINE THE SOLUTION SET (x(1-5))
WHICH IS THE AMOUNT EACH INITIAL GUESS SHOULD BE INCREMENTED BY.

```
do 270 k=1,4
   jj=k
   big=abs(z(k,k))
   kpl=k+1
   C search for the largest possible pivot element.
   do 220 i=kpl,5
      abs=abs(z(i,k))
      if (big-abs) 210,220,220
   210   big=abs
   jj=i
   220 continue
   C decision on necessity of row interchange.
   C   if (jj-k) 230,250,230
   C   row interchange.
   230 do 240 j=k,6
      temp=z(jj,j)
```
z(jj,j)=z(k,j)

z(k,j)=temp

C
C calculation of elements of new matrix

C

250
do 260 i=klpl,5
quotn(i,k)/z(k,k)
do 260 j=klpl,6
260 z(i,j)=z(i,j)-quot\*z(k,j)
do 270 i=klpl,5
270 z(i,k)=0.0

C
C first step in back substitution

C

x(5)=z(5,6)/z(5,5)

C
C remainder of back substitution process

C

do 290 ii=1,4
sum=0.0
i=5-ii
ipl=i+1
do 280 j=ipl,5
sum=sum+z(i,j)*x(j)
280 x(i)=(z(i,6)-sum)/z(i,i)
290
C
C NOW CHECK FOR CONVERGENCE OF NEWTON-RAPHSON METHOD BY TESTING
C TO SEE IF ALL INCREMENT ADDITIONS (x(1-5)) ARE LESS THAN .001
C

if (abs(x(1)) .gt. .01) then
goto 200
endif

if (abs(x(2)) .gt. .01) then
goto 200
endif

if (abs(x(3)) .gt. .01) then
goto 200
endif

if (abs(x(4)) .gt. .01) then
goto 200
endif

if (abs(x(5)) .gt. .01) then
goto 200
endif

endif

if (g .lt. 0.0) then
DETERMINE THE TEMPERATURE DECREASE OF THE PRODUCT GAS BY EVAPORATING THE MOISTURE IN THE INCOMING WOOD. THIS IS DONE BY DETERMINING THE ENERGY REQUIRED TO EVAPORATE THE MOISTURE IN THE WOOD AT THE GIVEN PRESSURE THEN PERFORMING AN ENERGY BALANCE.

\[
\text{MDirt} = X_{\text{clirt}} \times 22.84
\]

\[
\text{Press} = P_g
\]

\[
T_{\text{gas}} = (T_{\text{gas}} - 32.0) \times (5.0/9.0) + 273.15
\]

\[
T_{\text{amb}} = (T_{\text{amb}} - 32.0) \times (5.0/9.0) + 273.15
\]

\[
M_{\text{tot}} = 22.84/(1.0 - W_{\text{S1}})
\]

\[
M_{\text{wat}} = W_{\text{S1}} \times M_{\text{tot}}
\]

\[
\text{Enth}_{\text{1}} = W P(\text{press})
\]

\[
\text{Enth}_{\text{water}} = H v W P(T_{\text{gas}}, \text{press})
\]

\[
H_f = r_{\text{enthalpy}} - \text{enthalpy}
\]

THE TOTAL ENERGY REQUIRED TO EVAPORATE THE WATER IS CALCULATED.

\[
H_{\text{fg}} = 2.326
\]

\[
Q_{\text{dir2}} = \text{MDirt} \times 0.8374 \times (T_{\text{gas}} - T_{\text{amb}})
\]

\[
Q_{\text{dry}} = (4.187 \times (T_{\text{gas}} - T_{\text{amb}})) + H_{\text{fg}} + (T_{\text{gas}} - T_{\text{amb}}) \times 3.06 \times 22.84
\]

THE SUBSEQUENT EXIT PRODUCT GAS TEMPERATURE \((T_{\text{pg02}})\) CAN BE CALCULATED AS FOLLOWS:

\[
dT_2 = \frac{(Q_{\text{dry}} + Q_{\text{dir2}})}{(y_{P_\text{GAr}} \times 20.785 + (y_{P_\text{GCO2}} \times 37.0392) + (y_{P_\text{GCO}} \times 29.1648) + (y_{P_\text{GH2O}} \times 33.7014))}
\]

\[
T_{\text{pg02}} = A + F - dT_2
\]

DETERMINE EQUILIBRIUM TEMPERATURE AT WHICH THE WATER AND GAS EXIT THE GASIFIER.

\[
\text{Cons} = T_{\text{pg02}} / 100.0
\]

IF \((\text{Cons} < 3.73)\) THEN

\[
\text{Write}(\text{'gasifier fails to operate at this moisture content'}) \text{Goto 1314}
\]

ELSE

\[
Z2 = 3.73
\]

\[
\text{Delta} = 1.0
\]

\[
Z1 = T_{\text{gas}} / 100.0
\]

CALL CSUBP(Z1, Z2, dh02, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)

\[
L_S = (M_{\text{wat}} / 18.0) \times (dH_{\text{H2O}})
\]

\[
Z1 = Z2
\]

CALL CSUBP(Z1, cons, dh02, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)

\[
R_S = y_{P_\text{GH2O}} \times dH_{\text{H2O}} + y_{P_\text{GH2}} \times dH_{\text{H2}} + y_{P_\text{GAr}} \times dH_{\text{Ar}} + y_{P_\text{GCO2}} \times dH_{\text{CO2}} + y_{P_\text{GCO}} \times dH_{\text{CO}} + y_{P_\text{GH2}} + y_{P_\text{GAr}} + y_{P_\text{GCO2}} + y_{P_\text{GCO}}
\]

\[
\text{checc} = R_S - L_S
\]

IF \((\text{checc} > 0.0)\) THEN

\[
Z2 = Z2 + \text{Delta}
\]
goto 1313
else if (delta .ge. .005) then
    ZZ=ZZ-delta
    delta=delta/10.0
    ZZ=ZZ+delta
    goto 1313
else
    Texit=ZZ*100.0
    Tpg02=Texit
endif
goto 1315

1314

C

NOW ADD THE WATER EVAPORATED TO THE MOLES OF WATER PRODUCED IN C
PYROLYSIS TO YIELD FINAL WATER AMOUNT.
C

1315

yPGH20=yPGH20+(Mwat/18.0)
eng=(yPGCH4*838355.0)+(yPGH2*285803.8)+(yPGCO*282988.0)
mas=(yPGCO*28.0)+(yPG6CO2*44.0)+(yPGH20*18.0)+(yPGH2*2.0)
> + (yPGCH4*16.0)+(yPGN2*28.0)+(yPGAr*39.95)

HHUpgb=eng/mas
C

ASSUME THE PRODUCT GAS IS COOLED IN A CYCLONE/SCRUBBER TO C
CONDENSE OUT THE WATER VAPOR AND TARS AND OILS.
C

mas=mas-(yPGH20*18.0)
HHUpga=eng/mas
wmass=yPGH2018.0
yPGH20=0.0
C

RECIRCULATION OF CLEAN, COMPRESSED PRODUCT GAS THROUGH RECUPERATOR C TO UTILIZE SENSIBLE ENERGY IN DIRTY PRODUCT GAS STREAM EXITING C THE GASIFIER. (ASSUME A TERMINAL TEMPERATURE DIFFERENCE OF 5 K C IN RECUPERATOR.
C

if (Tpg02 .gt. 304.0) then
    Ttin=Tpg02-5.0
else
    Ttin=298.0
endif
C

ENTER THE TRIMBURNER AND PERFORM ENERGY BALANCE C

HFH20=-241827.0
HFCO2=-393522.0
HFCO=-110529.0
HFCH4=-74873.0
Pfgex=Pair-dPburn
Nair=MRa022/(28.97*2.2046244)
N02=y02in*Nair
NN2=yN2in*Nair
\[
\begin{align*}
\text{NH}_2\text{O} &= y\text{H}_2\text{O}\text{nNair} \\
\text{Na}_2 &= y\text{Ar}\text{nNair} \\
\text{NCO}_2 &= y\text{CO}_2\text{nNair} \\
\text{NCO} &= y\text{CO}\text{nNair} \\
Z_2 &= T_{\text{aq02}}/100.0 \\
z &= 2.98 \\
\text{CALL CSUBP}(Z1, Z2, dh\text{O}_2, dh\text{N}_2, dh\text{H}_2\text{O}, dh\text{CO}_2, dh\text{CO}, dh\text{H}_2, dh\text{CH}_4, dh\text{Ar}) \\
dh\text{PG}_0 &= dh\text{C}_0 \\
dh\text{PG}_0 &= dh\text{C}_0 \\
dh\text{PG}_0 &= dh\text{H}_2 \\
dh\text{PG}_0 &= dh\text{H}_2 \\
dh\text{PG}_0 &= dh\text{CH}_4 \\
dh\text{PG}_0 &= dh\text{Ar} \\
dh\text{PG}_0 &= dh\text{N}_2 \\
Z_2 &= ((T_{\text{aq31}}-32.0)*(5.0/9.0)+273.0)/100.0 \\
\text{CALL CSUBP}(Z1, Z2, dh\text{O}_2, dh\text{N}_2, dh\text{H}_2\text{O}, dh\text{CO}_2, dh\text{CO}, dh\text{H}_2, dh\text{CH}_4, dh\text{Ar}) \\
dh\text{CO}_2 &= dh\text{CO}_2 \\
dh\text{H}_2\text{O} &= dh\text{H}_2\text{O} \\
dh\text{O}_2 &= dh\text{O}_2 \\
dh\text{N}_2 &= dh\text{N}_2 \\
dh\text{Ar} &= dh\text{Ar} \\
dh\text{CO} &= dh\text{CO} \\
\text{KPH}_20 &= \text{NH}_2\text{O} \\
\text{KPCO} &= (X\text{CO})(\text{NCO}+\text{NCO}_2) \\
\text{KPCO}_2 &= (1.-X\text{CO})(\text{NCO}+\text{NCO}_2) \\
\text{KP02} &= \text{NCO}_2+\text{NH}_2\text{O}/2.0+\text{NCO}_2/2.0-\text{NH}_2\text{O}/2.0-(\text{NCO}*(1.-X\text{CO})) \\
> -(\text{NCO}*(1.-X\text{CO}))-((\text{NCO}*(\text{XCO}/2.0))-((\text{NCO}*(\text{XCO}/2.0)) \\
\text{KN}_2 &= \text{NN}_2 \\
\text{KAPR} &= \text{Na}_2 \\
\text{UPH}_2 &= y\text{P6H}_2+y\text{PG}_2+y\text{PGH}_2/2.0+y\text{PGH}_2/2.0+2y\text{PGCH}_4 \\
\text{UPC0} &= (\text{XCO})*(y\text{P6C}_0+y\text{PGCO}_2+y\text{PGCH}_4) \\
\text{UPC0}_2 &= (1.-X\text{CO})*(y\text{P6C}_0+y\text{PGCO}_2+y\text{PGCH}_4) \\
\text{UPG2} &= y\text{PG}_2/2.0+y\text{PGCO}_2+y\text{PGCO}_2/2.0-y\text{PG}_2/2.0-y\text{PG}_2/2.0-y\text{PGCH}_4/2.0 \\
> -(y\text{PG}_2*(1.-X\text{CO}))-((y\text{PG}_2*(1.-X\text{CO}))-((y\text{PG}_2*(1.-X\text{CO}))-((y\text{PG}_2*(1.-X\text{CO}))-((y\text{PG}_2*(1.-X\text{CO}))-((y\text{PG}_2*(1.-X\text{CO}))) \\
\text{VPN}_2 &= y\text{PGN}_2 \\
\text{UPV} &= y\text{PG}Ar \\
\text{KREA} &= (\text{NCO}*(\text{HFC0}+\text{NdhC0}))))+(\text{NCO}2(*(\text{HFC0}+\text{NdhC0}2))+(\text{NH}_20+(\text{HFC0}+\text{NdhC0}2))+(\text{HFC0}+\text{NdhC0}2) \\
> \text{NdhH}_2\text{O}+(\text{NaN}2)*(\text{NdhN}_2)+((\text{N2}2)+(\text{NdhN}_22)+((\text{NdhN}_22) \\
\text{VREA} &= (y\text{PGCO}_2(*(\text{HFCO}+\text{NdhC0}2))+(\text{HFCO}+\text{NdhC0}2))+(\text{HFCO}+\text{NdhC0}2) \\
> ((\text{NdhH}_2\text{O}2)+((\text{NdhH}_2\text{O}2))+(\text{NdhH}_2\text{O}2)+((\text{NdhH}_2\text{O}2))+(\text{NdhH}_2\text{O}2)+((\text{NdhH}_2\text{O}2)) \\
\text{KPR00} &= ((\text{KPH}_20)+(\text{HFCO}+\text{NdhC0}2))+(\text{KPCO}2)+((\text{KPCO}2))*H \\
\text{FCO2} &= ((\text{KPCO}2)+(\text{NdhC0}2))+(\text{KPCO}2) +((\text{KPCO}2) \\
\text{VPR00} &= ((\text{VPH}_20)+(\text{HFCO}+\text{NdhC0}2))+(\text{VPCO}2)+((\text{VPCO}2)+((\text{VPCO}2) \\
\text{FCO2} &= ((\text{VPCO}2)+(\text{NdhC0}2))+(\text{VPCO}2) +((\text{VPCO}2) \\
\text{Xmole} &= (((1.0-\text{radloss})*\text{KREA})-\text{KPR00})/((\text{VPR00}-(1.0-\text{radloss})*\text{VREA})) \\
\end{align*}
\]
SECOND LAW ANALYSIS OF ENTIRE SYSTEM

MRue0(02Ing+32.0+N2inv28.0+H2Oing18.0+CO2ing44.0+COing28.0
+Arin39.94)/22.84

+yan(39.94))/22.84

Avair=Hgas(Ta031,Tds,yc02in,yh2oin,yo2in,yn2in,yarin,yC0in)-

(Tds+459.67)*Sgaast(Ta031,Pair,Tds,14.636,yC02in,yH2oin,
yo2in,yN2in,yarin,yC0in,yrC02,yrH20,yr02,yrN2,yrAr,yrC0)

Avud=880.0

Avgas=Hgas(Ta032,Tds,yc02,yh20,ye02,yeN2,yeAr,yeC0)-Tds+

459.67)*Sgaast(Ta032,Pfgex,Tds,14.636,yc02,yH20,ye02,
yeN2,yeAr,yeC0,yrC02,yrH20,yr02,yrN2,yrAr,yrC0)
Temp=77.0
AVwat=HfWT(Temp)-(Tds+459.67)*SfWT(Temp)
AVgair=0.0
AVgair=AVgair+MRa022
AVfgas=AVfgas+MRa032
AVwat=AVwat+((wmass*MRdwg)/22.84)
AVgair=AVgair+MRa032
IRRtrm=(AVgair+AVair+AVud+AVwat+AVpgce)*1.055
Seff=((AVfgas-AVair)/(AVgair+AVud+AVpgce))*100.0
htout=(OdhCO2*ycO2)+(OdhH20*ycH20)+(OdhO2*ycO2)+(OdhN2*ycN2)
htout=(OdhAr*ycAr)+(OdhCO*ycCO)
htin=(NN2*NdhN2)+(N2*NdhO2)+(NH2O*NdhH2O)+(Na*NdhAr)+(NCO2*NhcO2)
EFtrm=((htout-HTin)/(467507.4*Xmole+WRpgce))*100.0
WRITE(8,*)", TRIMBURNER SUB-SYSTEM'
WRITE(8,*)", --GASIFIER SECTION--'
WRITE(8,*)'Gasifier Pressure = ',Pg,' psi'
WRITE(8,*)'Gasifier Flame Temperature = ',AFT,' F'
WRITE(8,*)' Product Gas Exit Temperature = ',Tpg02,' F'
WRITE(8,*)' Inlet Wet Wood Mass Rate = ',MRwwg,' lbs/s'
WRITE(8,*)' Inlet Dry Wood Mass Rate = ',MRdwg,' lbs/s'
WRITE(8,*)' Inlet air mass rate = ',mAIR,' lbs/s'
WRITE(8,*)' HHV of Gas before cyclone = ',HHVpgb,' Btu/lb'
WRITE(8,*)' HHV of Gas after cyclone = ',HHVpga,' Btu/lb'
WRITE(8,*)' Mass Rate of Product Gas = ',MRP6,' lbs/s'
WRITE(8,*)' Composition of Product Gas:
WRITE(8,*)', CO: ',yPGCO,' moles/mole of dry wood gasified'
WRITE(8,*)', CO2: ',yPGCO2,' moles/mole of dry wood gasified'
WRITE(8,*)', H2: ',yPGH2,' moles/mole of dry wood gasified'
WRITE(8,*)', H2O: ',yPGH2O,' moles/mole of dry wood gasified'
WRITE(8,*)', CH4: ',yPGCH4,' moles/mole of dry wood gasified'
WRITE(8,*)', N2: ',yPN6N2,' moles/mole of dry wood gasified'
WRITE(8,*)', Ar: ',yP6Ar,' moles/mole of dry wood gasified'
WRITE(8,*)', --TRIMBURNER SECTION--'
WRITE(8,*)', Mass Rate of Air Entering = ',MRa022,' lbs/s'
WRITE(8,*)', Mass Rate of Air Exiting = ',MRa032,' lbs/s'
WRITE(8,*)', Product Gas Inlet Temp. = ',Ttin,' F'
WRITE(8,*)', Exit Air Temperature = ',Ta031,' F'
WRITE(8,*)', Work Rate of Gas Compressor = ',WRpgc,' KW'
WRITE(8,*)', Overall Sub-system Efficiency = ',EFtrm'
WRITE(8,*)', Overall Irreversibility = ',IRRtrm,' KW'
WRITE(8,*)', Second Law Efficiency = ',Seff,' %'
RETURN
END
SUBROUTINE CSUBP(Z1, Z2, dh02, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, > dhAr)
dh02=((37.432*(Z2-Z1))+(.00804*((Z2**2.5)-(Z1**2.5)))+357.14*( > (Z2**(-.5))-(Z1**(-.5))))-(236.88*((Z2**(-1.0))-(Z1**(-1.0))))
192

\[ \text{dhN}_2 = (39.06 \cdot (Z_2 - Z_1)) + (1025.58 \cdot ((Z_2^{**(-1.0)} - (Z_1^{**(-1.0)})) + (410.2 \cdot ((Z_2^{**(-2.0)} - (Z_1^{**(-2.0)})) + (1072 \\
\text{dhH}_2O = ((143.05 \cdot (Z_2 - Z_1)) - (146.832 \cdot ((Z_2^{**(1.25)} - (Z_1^{**(1.25)})) + (55.167 \cdot (Z_2^{**(1.5)} - (Z_1^{**(1.5)})) - (1.84945 \cdot ((Z_2^{**(2.0)} - (Z_1^{**(2.0)})) + (2.052 \cdot ((Z_2^{**(2.0)} - (Z_1^{**(2.0)})) + (8.066001E-03 \cdot ((Z_2^{**(3.0)} - (Z_1^{**(3.0)})) + (20.78494 \cdot (Z_2 - Z_1)) + (40.265 \cdot ((Z_2^{**(1.75)} - (Z_1^{**(1.75)})) - (401.54 \cdot ((Z_2^{**(1.5)} - (Z_1^{**(1.5)})) + (707.04 \cdot ((Z_2^{**(2.25)} - (Z_1^{**(2.25)})) + (116 \\
\text{dhCO}_2 = ((-3.357 \cdot (Z_2 - Z_1)) + (20.353 \cdot ((Z_2^{**(1.5)} - (Z_1^{**(1.5)})) - (2.052 \cdot ((Z_2^{**(2.0)} - (Z_1^{**(2.0)})) + (8.066001E-03 \cdot ((Z_2^{**(3.0)} - (Z_1^{**(3.0)})) + (20.78494 \cdot (Z_2 - Z_1)) + (40.265 \cdot ((Z_2^{**(1.75)} - (Z_1^{**(1.75)})) - (401.54 \cdot ((Z_2^{**(1.5)} - (Z_1^{**(1.5)})) + (707.04 \cdot ((Z_2^{**(2.25)} - (Z_1^{**(2.25)})) + (116 \\
\text{dhCO}_2 = (56.505 \cdot (Z_2 - Z_1)) - (2010.96 \cdot ((Z_2^{**(1.5)} - (Z_1^{**(1.5)})) + (401.54 \cdot ((Z_2^{**(1.5)} - (Z_1^{**(1.5)})) + (707.04 \cdot ((Z_2^{**(2.25)} - (Z_1^{**(2.25)})) + (116 \\
\text{dhCH}_4 = ((-372.87 \cdot (Z_2 - Z_1)) + (351.792 \cdot ((Z_2^{**(1.25)} - (Z_1^{**(1.25)})) - (14.2143 \cdot ((Z_2^{**(1.75)} - (Z_1^{**(1.75)})) + (647.76 \cdot ((Z_2^{**(1.5)} - (Z_1^{**(1.5)})) + (1072 \\
\text{dhAr} = (20.76494 \cdot (Z_2 - Z_1)) \times 100.0 \\
\text{RETURN} \\
\text{END} \]
SUBROUTINE GASIFY(wb51, Tds, Xdirt, Xash, Ta01, yrCO2, yrH2O, yrO2, 
   yrN2, yrAr, yrCO, Pg, Eqrat, radloss, XCC0, Ta032, yrCO2in, yrH2Oin, 
   yrO2in, yrN2in, yrArin, yrCOin, Pair, dPburn, Ta031, MRA022, mwfg, yPGCO2, 
   yPGH2O, yPGH2, yPGCH4, yPGN2, yPGAr, Tpg02, yeCO2, yeH2O, 
   yeO2, yeN2, yeAr, yeCO, MRA032, Pfgex, MRP6, MRwgg, MRdgg, Eftrm, 
   print, IRRtrm, WRpgc)

C
C THIS SUBROUTINE DETERMINES THE TEMPERATURE AND COMPOSITION OF
C THE PRODUCT GAS EXITING AN UP-DRAFT GASIFIER. THE PROCESS USED
C IN THIS SUBROUTINE IS AS FOLLOWS: 1) DETERMINE THE ADIABATIC
C FLAME TEMPERATURE IN THE COMBUSTION ZONE OF THE GASIFIER BY
C ITERATION ON THE HEAT BALANCE OF THE COMBUSTION OF WOOD. 2) USE
C THE FLAME TEMPERATURE OBTAINED AS THE PYROLYSIS TEMPERATURE AND
C DETERMINE THE COMPOSITION OF THE PRODUCT GAS PER 1 MOLE OF DRY
C WOOD (ASSUMED) USING A NEWTON-RAPHSON SCHEME FOR A SYSTEM OF 5
C NON-LINEAR SIMULTANEOUS EQUATIONS (3 FOR ELEMENT BALANCE, 1 FOR
C HEAT BALANCE, 1 FOR CHEMICAL EQUILIBRIUM). 3) DETERMINE THE
C EXIT TEMPERATURE OF THE PRODUCT GAS AFTER DRYING THE EQUIVALENT
C WOOD YIELDING 1 MOLE OF DRY WOOD IN THE PYROLYSIS ZONE.

LIST OF VARIABLES

---

wb51 - MOISTURE CONTENT OF THE INLET WOOD ON A WET BASIS.
Tds - INLET WOOD TEMPERATURE.
Xdirt - MASS RATIO OF DIRT IN INLET WOOD.
Xash - MASS RATIO OF ASH IN INLET WOOD.
Ta01 - AMBIENT AIR TEMPERATURE
yrCO2 - MOLE FRACTION OF CO2 IN AMBIENT AIR ENTERING GASIFIER.
yrH2O - MOLE FRACTION OF H2O IN AMBIENT AIR ENTERING GASIFIER.
yrO2 - MOLE FRACTION OF O2 IN AMBIENT AIR ENTERING GASIFIER.
yrN2 - MOLE FRACTION OF N2 IN AMBIENT AIR ENTERING GASIFIER.
yrAr - MOLE FRACTION OF Ar IN AMBIENT AIR ENTERING GASIFIER.
yrCO - MOLE FRACTION OF CO IN AMBIENT AIR ENTERING GASIFIER.
Eqrat - MASS RATIO OF ENTERING AIR TO ENTERING WOOD.
yPGCO2 - MOLES OF CO FORMED IN GASIFIER PER ONE MOLE OF WOOD.
yPGH2O - MOLES OF H2O FORMED IN GASIFIER PER ONE MOLE OF WOOD.
yPGH2 - MOLES OF H2 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
yPGCH4 - MOLES OF CH4 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
yPGN2 - MOLES OF N2 FORMED IN GASIFIER PER ONE MOLE OF WOOD.
yPGAr - MOLES OF Ar FORMED IN GASIFIER PER ONE MOLE OF WOOD.
Tpg - TEMPERATURE OF PRODUCT GASES EXITING GASIFIER.

---

implicit real(a-z)
integer i, j, k, jj, kpi, ii, ili
logical print*2
dimension z(5,6), x(5)

DETERMINE MOLES OF ELEMENTS ENTERING GASIFIER IN UNDERFIRE
COMBUSTION AIR.
\[ O_2 \text{ing} = (E \text{rat}) \times (1.055) \]
\[ N_2 \text{ing} = \frac{O_2 \text{ing}}{yrO_2} \times yrN2 \]
\[ CO_2 \text{ing} = \frac{O_2 \text{ing}}{yrO_2} \times yrCO2 \]
\[ H_2O \text{ing} = \frac{O_2 \text{ing}}{yrO_2} \times yrH2O \]
\[ CO \text{ing} = \frac{O_2 \text{ing}}{yrO_2} \times yrCO \]
\[ Ar \text{ing} = \frac{O_2 \text{ing}}{yrO_2} \times yrAr \]

MOLES OF DRY WOOD IN GASIFIER THAT BURNS(XX)

\[ XX = E \text{rat} \]

HEAT BALANCE EQUATION ITERATED ON TO DETERMINE THE ADIABATIC FLAME TEMPERATURE IN THE COMBUSTION ZONE.

\[ \Delta = 1.0 \]
\[ Z1 = 2.98 \]
\[ Z2 = 5.58 \]
call CSUBP(Z1, Z2, dhO2, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)
LHS = \((XX) \times (-115000.13) + (O_2 \text{ing}) \times (dhO2) + (N_2 \text{ing}) \times (dhN2) \)
\]

GUESS AN AFT AND THEN CALCULATE THE RIGHT HAND SIDE OF THE HEAT BALANCE (PRODUCTS) AND COMPARE WITH LEFT HAND SIDE. ITERATE ON THE AFT UNTIL THE TWO SIDES ARE EQUAL.

\[ Z2 = 3.00 \]
100 call CSUBP(Z1, Z2, dhO2, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)
twd = \((Z2 \times 100.0) - 273.15 \times (9.0/5.0) + 32.0 \)
tz1 = 77.0

cpwd = 0.266 + (0.00322 * (twd + tz1 - 64.0))
cpwd = 4.1868 * cpwd * 22.84

t = (XX) \times (-393522.0 + dhCO2) + (7.0 * XX) \times (-241827.0 + dhH2O) + (N_2 \text{ing}) \times (dhN2) + (1.0 - XX) \times cpwd \times (Z1 - Z2) \times 100.0)
check = RHS - LHS
if (check .lt. 0.0) then
\]
Z2 = Z2 + delta
goto 100
else if (delta .ge. 0.05) then
\]
Z2 = Z2 - delta
delta = delta / 10.0
Z2 = Z2 + delta
goto 100
else
AFT = Z2 \times 100.0
endif

NOW ASSUME THAT THE AFT IS THE TEMPERATURE THAT THE WATER SHIFT REACTION OCCURS AT IN THE REDUCTION ZONE. KNOWING THIS, THE EQUILIBRIUM CONSTANT OF THAT REACTION CAN BE DETERMINED. THIS IS DONE BY USING A CURVE FIT OF THE PLOT OF Keq VERSUS TEMP.

if (AFT .lt. 1000.0) then
\]
Keq = 55.2846 - (1.0443 \times AFT) + (5.06016E-05) \times (AFT \times 2.0)
else if (AFT .lt. 1200.0) then
\]
Keq = 17.7387 - (0.266841 \times AFT) + (1.05697E-05) \times (AFT \times 2.0)

endif
else if (AFT .lt. 1400.0) then
  \[ K_{eq} = 8.75579 - (0.0111979 \times AFT) + (3.75273 \times 10^{-6} \times AFT^2) \]
else
  \[ K_{eq} = 3.78197 - (3.91905 \times 10^{-3} \times AFT) + (1.08563 \times 10^{-6} \times AFT^4) \]
endif

C THE NEWTON-RAPHSON METHOD WILL NOW BE USED. THE MATRIX IS LOADED
C WITH THE M-MATRIX BEING THE PARTIAL DERIVATIVES OF THE 5 EQUATIONS
C AND THE X-MATRIX BEING THE SOLUTION SET. THE SET IS USED TO
C INCREMENT THE MOLE COMPOSITIONS.
C
C INITIAL GUESS AT THE COMPOSITION OF THE PRODUCT GAS
C
  if (EQrat .le. .65) then
    c = .68
    d = .05
    e = .9
    f = .1
    g = .05
  else if (EQrat .eq. .70) then
    \( y_{P6H2} = 0.0 \)
    \( y_{P6H20} = 0.4 + H20_{ing} \)
    \( y_{P6CO} = 0.0 \)
    \( y_{P6CH4} = .15 \)
    \( y_{P6CO2} = .85 \)
    \( y_{P6N2} = N2_{ing} \)
    \( y_{P6Ar} = Ar_{ing} \)
    goto 507
  else
    \( y_{P6H2} = 0.0 \)
    \( y_{P6H20} = .7 + H20_{ing} \)
    \( y_{P6CO} = 0.0 \)
    \( y_{P6CO2} = 1.0 \)
    \( y_{P6CH4} = 0.0 \)
    \( y_{P6N2} = N2_{ing} \)
    \( y_{P6Ar} = Ar_{ing} \)
    goto 507
  endif

  coefc = dhH2
  coefd = \(-241827.0 + dhH20\)
  coffe = \(-110529.0 + dhCO\)
  coeff = \(-393522.0 + dhCO2\)
  coefg = \(-7\times 873.0 + dhCH4\)

  \[ z(1,1) = 0.0 \]
  \[ z(1,2) = 0.0 \]
  \[ z(1,3) = 1.0 \]
  \[ z(1,4) = 1.0 \]
  \[ z(1,5) = 1.0 \]
  \[ z(2,1) = 2.0 \]
  \[ z(2,2) = 2.0 \]
  \[ z(2,3) = 0.0 \]
  \[ z(2,4) = 0.0 \]
  \[ z(2,5) = 4.0 \]
  \[ z(3,1) = 0.0 \]
  \[ z(3,2) = 1.0 \]
\[ z(3,3) = 1.0 \]
\[ z(3,4) = 2.0 \]
\[ z(3,5) = 0.0 \]
\[ z(4,1) = \text{coeffc}(-1.0) \]
\[ z(4,2) = \text{coeffd}(-1.0) \]
\[ z(4,3) = \text{coeffe}(-1.0) \]
\[ z(4,4) = \text{coefff}(-1.0) \]
\[ z(4,5) = \text{coeffg}(-1.0) \]
\[ z(5,1) = f/(e*d) \]
\[ z(5,2) = ((c*f)/(e*(d**2.0)))*(-1.0) \]
\[ z(5,3) = ((c*f)/(e**2.0*d)))*(-1.0) \]
\[ z(5,4) = c/(e*d) \]
\[ z(5,5) = 0.0 \]
\[ z(1,5) = (e+f+g)/(1.0)*(-1.0) \]
\[ z(2,6) = ((2.0*e)+2.0*d)+(4.0*g)-(1.4))/(1.0) \]
\[ z(3,6) = (e+2.0*f)-(0.59+2.0*0.2) \]
\[ z(4,6) = (((-1.0)*coeffc+c)+((-1.0)*coeffd)+(e**2.0)*coefff)+(0.0) \]
\[ z(5,6) = ((5.0)/(e*d)) \]

**WITH MATRIX FILLED, A LINEAR EQUATION SOLVER USING PARTIAL PIVOTING IS NEXT EMPLOYED TO DETERMINE THE SOLUTION SET \((x(1-5))\)**

**WHICH IS THE AMOUNT EACH INITIAL GUESS SHOULD BE INCREMENTED BY.**

\[
\begin{align*}
\text{do 270 } & k=1,4 \\
& \text{jj} = k \\
& \text{big} = \text{abs}(z(k,k)) \\
& \text{kpl} = k + 1 \\
\end{align*}
\]

**search for the largest possible pivot element.**

\[
\begin{align*}
\text{do 220 } & i = \text{kpl},5 \\
& \text{ab} = \text{abs}(z(i,k)) \\
& \text{if (big-ab)} = 210,220,220 \\
& \text{big} = \text{ab} \\
& \text{jj} = i \\
& \text{220 continue} \\
\end{align*}
\]

**decision on necessity of row interchange.**

\[
\begin{align*}
\text{if (jj-k)} = 230,250,230 \\
\end{align*}
\]

**row interchange.**

\[
\begin{align*}
\text{230 do 240 } & j = k,6 \\
& \text{temp} = z(j,j) \\
& z(j,j) = z(k,j) \\
& z(k,j) = \text{temp} \\
\end{align*}
\]

**calculation of elements of new matrix**

\[
\begin{align*}
\text{250 do 260 } & i = \text{kpl},5 \\
& \text{quot} = z(i,k)/z(k,k) \\
& \text{do 260 } j = \text{kpl},6 \\
\end{align*}
\]
197

260  \[ z(i,j) = z(i,j) - \text{quot} \cdot z(k,j) \]
    do 270 i=k+1,5
    \[ z(i,k) = 0.0 \]

270

C first step in back substitution
C

\[ x(5) = \frac{z(5,6)}{z(5,5)} \]

C remainder of back substitution process
C

    do 280 i=1,4
        \[ \text{sum} = 0.0 \]
        \[ i = 5 - i \]
        \[ i = i + 1 \]
    do 290 j=k+1,5
        \[ \text{sum} = \text{sum} + z(i,j) \cdot x(j) \]
        \[ x(i) = \frac{z(i,5) - \text{sum}}{z(i,i)} \]

280

290

C NOW CHECK FOR CONVERGENCE OF NEWTON-RAPHSON METHOD BY TESTING TO SEE IF ALL INCREMENT ADDITIONS \((x(1-5))\) ARE LESS THAN \(0.001\)
C

if (abs(x(1)) .gt. .001) then
    goto 200
else if (abs(x(2)) .gt. .001) then
    goto 200
else if (abs(x(3)) .gt. .001) then
    goto 200
else if (abs(x(4)) .gt. .001) then
    goto 200
else if (abs(x(5)) .gt. .001) then
    goto 200
else
    yPGH2=c
    yPGH2O=d+H20ing
    yPGCO=e
    yPGCO2=f
    yPGCH4=g
    yPGN2=N2ing
    yPGAr=Aring
endif

if (g .lt. 0.0) then
    yPGCH4=0.0
endif

C DETERMINE THE TEMPERATURE DECREASE OF THE PRODUCT GAS BY
C EVAPORATING THE MOISTURE IN THE INCOMING WOOD. THIS IS DONE
C BY DETERMINING THE ENERGY REQUIRED TO EVAPORATE THE MOISTURE
C IN THE WOOD AT THE GIVEN PRESSURE THEN PERFORMING AN ENERGY
C BALANCE.
THE TOTAL ENERGY REQUIRED TO EVAPORATE THE WATER IS CALCULATED.

\[ \text{Hfg}_{\text{sl}} = 2.326 \]

\[ \text{Odir} = \text{Mdir} \times 3374 \times (T_{\text{gas}} - T_{\text{amb}}) \]

\[ \text{Qdry} = (4.187 \times (T_{\text{gas}} - T_{\text{amb}})) + \text{Hfg}_{\text{sl}} + \text{Mwat} + ((T_{\text{gas}} - T_{\text{amb}}) \times 3.06) \]

CALCULATE THE TEMPERATURE DECREASE OF THE PRODUCT GAS (\( \Delta T \)), AND THE SUBSEQUENT EXIT PRODUCT GAS TEMPERATURE (\( T_{\text{pg02}} \)).

\[ \Delta T = \left( \frac{\text{Qdry} + \text{Odir}}{(y_{\text{PGAr}} \times 20.785) + (y_{\text{PGCO}} \times 37.0392) + (y_{\text{PGN}} \times 29.1564) + (y_{\text{PGH2}} \times 28.4182) + (y_{\text{PGCH4}} \times 35.0592) + (y_{\text{PGN}} \times 29.1648)} \right) \]

\[ T_{\text{pg02}} = T_{\text{pg01}} - \Delta T \]

PERFORM ENERGY BALANCE TO DETERMINE GASIFIER EXIT TEMPERATURE

\[ \text{cons} = T_{\text{pg02}} / 100.0 \]

if (\( \text{cons} \lt 3.73 \)) then
write(*,*) 'gasifier fails to operate at this moisture content' goto 1314
else
endif
\[ Z2 = 3.73 \]
\[ \text{delta} = 1.0 \]
1313
\[ Z1 = T_{\text{gas}} / 100.0 \]
call CSUBP(Z1, Z2, dh02, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)
\[ \text{LS} = (\text{Mwat} / 18.0) \times (dH2O) \]
\[ Z1 = Z2 \]
call CSUBP(Z1, cons, dh02, dhN2, dhH2O, dhCO2, dhCO, dhH2, dhCH4, dhAr)
\[ \text{RS} = y_{\text{PGH2O}} \times dH2O + y_{\text{PH2}} \times dH2 + y_{\text{PGAr}} \times dAr + y_{\text{PGN}} \times dN2 + y_{\text{PGCH4}} \times dCH4 \]
\[ + y_{\text{PGCO2}} \times dCO2 + y_{\text{PGCO}} \times dCO \]
checc = RS - LS
if (checc .gt. 0.0) then
\[ Z2 = Z2 + \text{delta} \]
goto 1313
else if (\( \text{delta} \geq 0.005 \)) then
\[ Z2 = Z2 - \text{delta} \]
\[ \text{delta} = \text{delta} / 10.0 \]
\[ Z2 = Z2 + \text{delta} \]
goto 1313
else
\[ \text{Texi} = Z2 \times 100.0 \]
\[ T_{\text{pg02}} = \text{Texi} \]
endif
goto 1315
1314 QQQ=\((yPGAr*20.785+yPGC02*37.0392+yPGC0*29.1564+yPGH2*28.4182\)\)
> \(+yPGCH4*36.5092+yPGN2*29.1468+yPGH2O*33.7014\)\)
QQQ=QQQ/((Mwat/18.0)+33.7014)
Texh=(QQQ*Tpg02+Tgas)/(1.0+QQQ)
Tpg02=Texh
C
NOW ADD THE WATER EVAPORATED TO THE MOLES OF WATER PRODUCED IN
PYROLYSIS TO YIELD FINAL WATER AMOUNT.
C
1315 yPGH20=\(yPGH20+\left(Mwat/18.0\right)\)
eng=(yPGCH4*890355.0)+(yPGH2*285838.0)+(yPGC0*282989.0)
mas=(yPGC0*28.0)+(yPGCO2*44.0)+(yPGH2O*18.0)+(yPGH2*2.0)
> \(+yPGCH4*16.0)+(yPGN2*28.0)+(yPGAr*39.95)\)
HHVpg=\(\text{eng/mas}\)
C
ASSUME THE PRODUCT GAS IS COOLED IN A CYCLONE/SCRUBBER TO
CONDENSE OUT THE WATER VAPOR AND TARS AND OILS.
C
mas=mas-(yPGH20*18.0)\)
HHVpg=\(\text{eng/mas}\)
wmass=\(yPGH2O*18.0\)
yPGH2O=0.0
C
ENTER THE TRIMBURNER AND PERFORM ENERGY BALANCE
C
if (wb51 .eq. .05) then
Mair=\(-.966035+(6.81557*EQrat)-(13.8913*EQrat**2.0)\)
> \(+9.30036*EQrat**3.0)\)
else if (wb51 .eq. .15) then
Mair=\(.0163648-(.0394664*EQrat)+(6.684121*EQrat**2.0)\)
> \(-.232909*EQrat**3.0)\)
else if (wb51 .eq. .30) then
Mair=\(.0246916-(.107134*EQrat)+(1.05424*EQrat**2.0)\)
> \(-.290044*EQrat**3.0)\)
else if (wb51 .eq. .40) then
Mair=\(.0316426-(.16553*EQrat)+(1.05424*EQrat**2.0)\)
> \(-.291372*EQrat**3.0)\)
else
Mair=\(-.265539+(1.80054*EQrat)-(3.00059*EQrat**2.0)\)
> \(+2.62528*EQrat**3.0)\)
endif
S01 MRe022=MRe022-Mair
HFH20=-(-241827.0)
HFCO2=(-393522.0)
HFCO=(-110529.0)
HFCH4=(-74873.0)
Pfgex=Pair-dPburn
Mair=MRe022/(28.97*2.2046244)
N01=-y02in*Mair
NN2=yN2in*Mair
NH20=\(yH20in\times Mair\)
NAr=yArin*Mair
NCO2=yCO2in*Mair
NCO=yCOin*Mair
RECIRCULATION OF CLEAN, COMPRESSED PRODUCT GAS THROUGH RECUPERATOR TO UTILIZE SENSIBLE ENERGY IN DIRTY PRODUCT GAS STREAM EXITING THE GASIFIER. (ASSUME A TERMINAL TEMPERATURE DIFFERENCE OF 5 K IN RECUPERATOR.

```fortran
if (Tpg02 .gt. 304.0) then
  TtIn=Tpg02-5.0
else
  TtIn=298.0
endif
Z2=TtIn/100.0
Z1=2.98
CALL CSUBP(Z1,Z2,dhO2,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr)
dhPGCO=dhCO
dhGCO2=dhCO2
dhPGH2=dhH2
dhGH2O=dhH2O
dhPGCH4=dhCH4
dhPGAr=dhAr
dhPGN2=dhN2
Z2=((Ta031-32.0)*(5.0/9.0)+273.0)/100.0
CALL CSUBP(Z1,Z2,dhO2,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr)
NdhCO2=dhCO2
NdhH2O=dhH2O
NdhO2=dhO2
NdhN2=dhN2
NdhAr=dhAr
NdhCO=dhCO
Z2=((Ta032-32.0)*(5.0/9.0)+273.0)/100.0
CALL CSUBP(Z1,Z2,dhO2,dhN2,dhH2O,dhCO2,dhCO,dhH2,dhCH4,dhAr)
  0 dhCO2=dhCO2
  0 dhH2O=dhH2O
  0 dhO2=dhO2
  0 dhN2=dhN2
  0 dhAr=dhAr
  0 dhCO=dhCO
  KPH2O=NH2O
  KPCO=(XCCO)*(NCO+NC02)
  KPCO2=(*1.-XCCO)*(*NCO+NC02)
  KPCO2=NC02+NH2O/2.0+NC02+NC0/2.0-NH2O/2.0-NC0*(1.-XCCO)
  > -(NC02*(1.-XCCO))-(NC0*(XCCO/2.0))-(NC02*(XCCO/2.0))
  KPN2=NN2
  KPAR=NAr
  VPH2O=yPGH2+yPH2O+(2*yPGCH4)
  VPCO=(XCCO)*(yPGCO+yPGCO2+yPGCH4)
  VPCO2=(XCCO)*(yPGCO+yPGCO2+yPGCH4)
  VPG2=yPGH2O/2.0+yPGH2+yPGCO2+yPGCO2/2.0-yPH2O/2.0-yPGH2O/2.0-yPGCH4/2.0
  > -(yPGCO*(1.-XCCO))-(yPGCO2*(1.-XCCO))-(yPGCH4*(1.-XCCO))-(yPGCO
  > (XCCO/2.0))-(yPGCO2*(XCCO/2.0))-(yPGCH4*(XCCO/2.0))
  VPN2=yPGN2
  VPAR=yPGAr
  KREA=(NCO)*(HFCO+NdhCO)+(NC02)*(HFCO2+NdhCO2)+(NH2O)*(HFCO2+NdhCO2)
  > NdhH20)+(NAr)*(NdhAr)+(NN2)*(NdhN2)+(NO2)*(NdhO2)
  VREA=(yPGCO)*(HFCO+yPGCO2)+(yPGCO2)*(HFCO2+yPGCO2)+(yPH2)*
SECOND LAW ANALYSIS OF ENTIRE SYSTEM
```

TRIMBURNER SUB-SYSTEM

--GASIFIER SECTION--

Gasifier Pressure = Pg, psi
Gasifier Flame Temperature = AFT, F
Product Gas Exit Temperature = Tpg02, F
Inlet Wet Wood Mass Rate = MRwwg, lbs/s
Inlet Dry Wood Mass Rate = MRdwg, lbs/s
HHV of Gas before cyclone = HHVggb, Btu/ib
HHV of Gas after cyclone = HHVgge, Btu/ib
Mass Rate of Product Gas = MRPG, lbs/s
Composition of Product Gas:

CO: yPGCO, moles/mole of dry wood gasified
CO2: yPGCO2, moles/mole of dry wood gasified
H2: yPGH2, moles/mole of dry wood gasified
H2O: yPGH2O, moles/mole of dry wood gasified
CH4: yPGCH4, moles/mole of dry wood gasified
N2: yPGN2, moles/mole of dry wood gasified
Ar: yPGAr, moles/mole of dry wood gasified

--TRIMBURNER SECTION--

Mass Rate of Air Entering = MRa022, lbs/s
Mass Rate of Air Exiting = MRa032, lbs/s
Product Gas Inlet Temp. = Tin, F
Inlet Air Temperature = Ta031, F
```

```
WRITE(8,*) 'TRIMBURNER SUB-SYSTEM'
WRITE(8,*) '---GASIFIER SECTION---'
WRITE(8,*) 'Gasifier Pressure = Pg, psi'
WRITE(8,*) 'Gasifier Flame Temperature = AFT, F'
WRITE(8,*) 'Product Gas Exit Temperature = Tpg02, F'
WRITE(8,*) 'Inlet Wet Wood Mass Rate = MRwwg, lbs/s'
WRITE(8,*) 'Inlet Dry Wood Mass Rate = MRdwg, lbs/s'
WRITE(8,*) 'HHV of Gas before cyclone = HHVggb, Btu/ib'
WRITE(8,*) 'HHV of Gas after cyclone = HHVgge, Btu/ib'
WRITE(8,*) 'Mass Rate of Product Gas = MRPG, lbs/s'
WRITE(8,*) 'Composition of Product Gas:
WRITE(8,*) 'CO: yPGCO, moles/mole of dry wood gasified'
WRITE(8,*) 'CO2: yPGCO2, moles/mole of dry wood gasified'
WRITE(8,*) 'H2: yPGH2, moles/mole of dry wood gasified'
WRITE(8,*) 'H2O: yPGH2O, moles/mole of dry wood gasified'
WRITE(8,*) 'CH4: yPGCH4, moles/mole of dry wood gasified'
WRITE(8,*) 'N2: yPGN2, moles/mole of dry wood gasified'
WRITE(8,*) 'Ar: yPGAr, moles/mole of dry wood gasified'
WRITE(8,*) '---TRIMBURNER SECTION---'
WRITE(8,*) 'Mass Rate of Air Entering = MRa022, lbs/s'
WRITE(8,*) 'Mass Rate of Air Exiting = MRa032, lbs/s'
WRITE(8,*) 'Product Gas Inlet Temp. = Tin, F'
WRITE(8,*) 'Inlet Air Temperature = Ta031, F'
```
WRITE(8,*),' Exit Air Temperature = ',Ta032,' F'
WRITE(8,*),' Work Rate of Gas Compressor = ',WRpgc,' KW'
WRITE(8,*),' Overall Sub-system Efficiency = ',Eftrm,' %'
WRITE(8,*),' Overall Irreversibility = ',Irrtrm,' Kw'
WRITE(8,*),' Second Law Efficiency = ',Seff,' %'
RETURN
END

SUBROUTINE CSUBP(Z1,Z2,dh02,dhN2,dhH20,dhCO2,dhCO,dhh2,dhCH4,dhAr)
  dh02=((37.432*(Z2-Z1))+.00804*((Z2**2.5)-(Z1**2.5))+(357.14*
  (Z2**(-.5))-(Z1**(-.5)))-(236.88*((Z2**(-1.0))-(Z1**(-1.0))))
  )*100.0
  dhN2=((39.06*(Z2-Z1))+(1025.58((Z2*(-.5))-(Z1**(-.5))))-(1072
  )+7*((Z2**(-1.0))-(Z1**(-1.0)))+(410.2*((Z2**(-2.0))-(Z1**(-2.
  ))))*100.0
  dhH20o=((143.05*(Z2-Z1))-(146.832((Z2**(1.25))-(Z1**(1.25))))+(*
  55.16701*(Z2**(1.5))-(Z1**(1.5)))-(1.84945*((Z2**2.0)-(Z1*
  *(2.0)))+100.0
  dhCO2((-3.7387(Z2-Z1))+(20.353((Z2**(1.5))-(Z1**(1.5))))-(2.
  )+52*((Z2**2.0)-(Z1**2.0)))+(0.068001E-03*((Z2**(3.0))-(Z1*
  *(3.0))))*100.0
  dhCO=((-69.14501*(Z2-Z1))-(40265*((Z2**(1.5))-(Z1**(1.5)))+(*
  401.54*((Z2**(1.5))-(Z1**(1.5)))+(707.044*((Z2**(2.5))-(Z1*
  *(2.5)))+100.0
  dhH2o((58.505*Z2-Z1))-(2810.96((Z2**(2.5))-(Z1**(2.5)))+116
  )+5.0*(alog(Z2)-(alog(Z1)))+(1121.4((Z2**(-.5))-(Z1**(-.5))))
  )*100.0
  dhH2((-672.87*(Z2-Z1))+(351.792((Z2**(1.25))-(Z1**(1.25))))-(
  14.2143((Z2**(1.75))-(Z1**(1.75)))+(647.766((Z2**(0.5))-(Z1*
  *(0.5)))))*100.0
  dhAr=(20.78494*(Z2-Z1))*100.0
RETURN
END