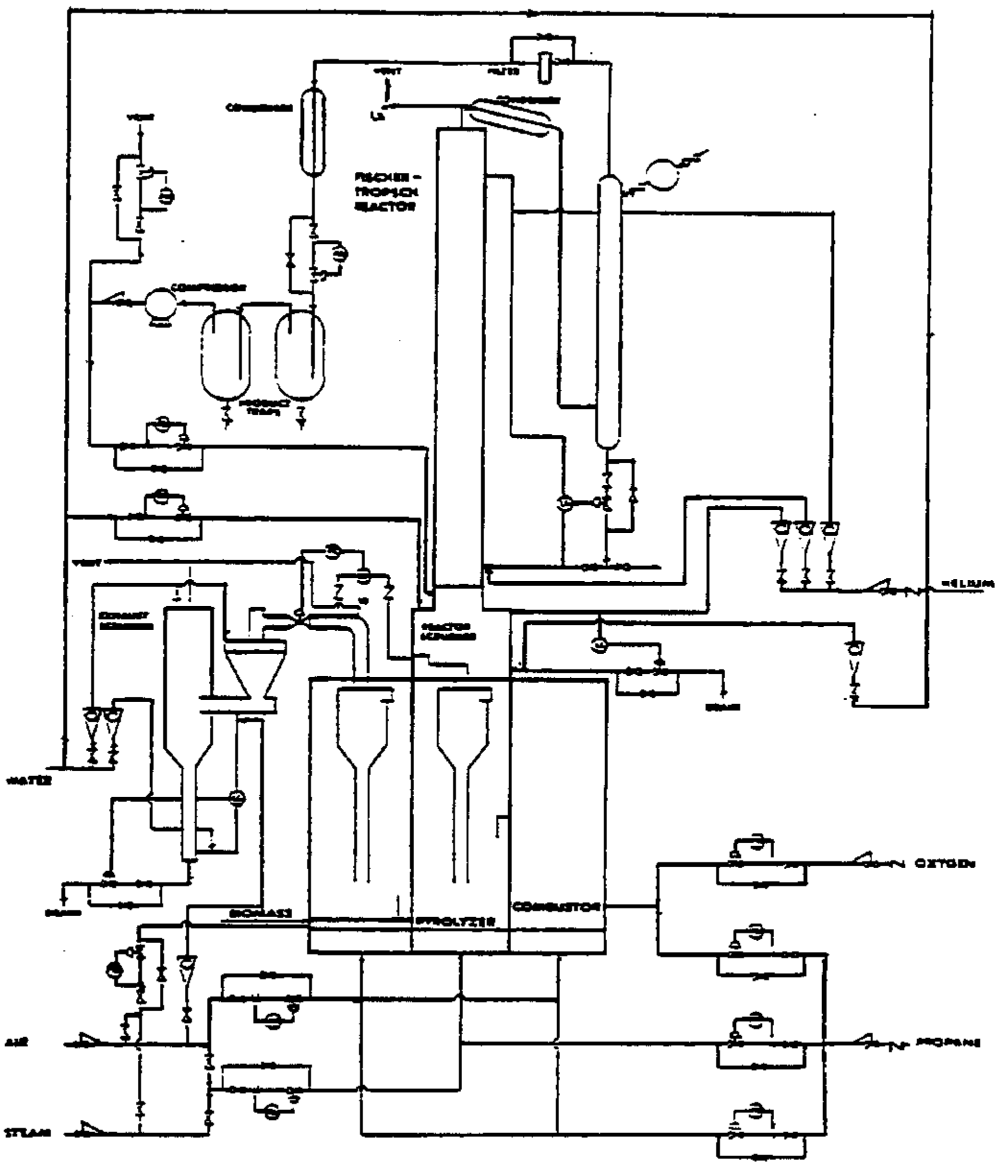

Figure 54. A schematic diagram of a staged reactor system for the indirect liquefaction of biomass.

The water flow rate is a manipulated variable and can be based on the scrubber exit gas temperature. The water level in the scrubber is automatically controlled.

The gas then flows into the Fischer-Tropsch reactor. The flow rate through this reactor can be manipulated but the pressure is controlled automatically. The gas remaining after the product traps can be recycled or vented.

Based on the work of Prasad (1986) the range of desirable operating temperatures for the pyrolyzer is between 900 and 1000 K . The reactor is designed to run at a feed rate of 10 lb/hr. The Fischer-Tropsch reactor will operate at 500 K (Dry, 1976). The entire reactor system will operate at a maximum pressure of 5 atm. Above this pressure significant amounts of methane will be formed. (Bungay, 1981) The Fischer-Tropsch synthesis has been successfully carried out at that pressure using both iron and cobalt catalysts. Iron catalyst promoted by alkali metals seem to show the most promise for low pressure liquefaction. (Dry, 1976)

Additional research is needed to develop an effective low pressure Fischer-Tropsch catalyst. An effective catalyst and promoter combination would have to be located before the actual test reactor can be constructed. Alternatively, a catalyst that promotes the formation of the proper synthesis gas composition in the pyrolyzer at high pressures could be developed. If such a catalyst were located then the reactor could operate at standard Fischer-Tropsch pressures of 10 atm. or greater.

Additional overall configurations can also be examined in order to determine their operating characteristics as opposed to those of the currently proposed design. Two potential alternatives are shown in Figure 18.

The final area that future investigators may find fruitful is the control system to be employed. The control system shown in Figure 54 deals with each unit individually. A more effective technique may be to implement an overall control system that considers several variables before adjusting the operation of the system.

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APPENDIX A

DESIGN RATIONALE

PYROLYZER / COMBUSTOR STAGE

The first dimension determined was the height of the pyrolysis stage. Figure 30 shows the effect of the height of the reactor on the product gas composition from the pyrolysis stage as predicted by the preliminary pyrolysis model. As can be seen, after the initial ten centimeters the bed height has only a marginal effect of the conversion. Thus, the bed must be at least 10 centimeters tall. However, a ten centimeter bed would not be sufficient for physical reasons. The bed would not have a uniform radial concentration profile. In addition, it would not have the thermal stability needed to maintain a constant temperature under the feed conditions that will be encountered. Unfortunately, it is not possible to determine the height needed to remedy these problems with any precision. From experience with the currently existing indirect liquefaction system, it was determined that the desirable bed height was between one and two feet. A height of 16 in. was selected. This height was sufficient to provide for mixing and thermal stability while also providing room for the installation of such internal components as a cyclone. The combustor bed must be the same height as that of the pyrolyzer due to the interconnection between the two beds.

The diameter of the pyrolysis reactor was determined by the desire to be able to use currently available equipment to feed the test reactor. The feeder is capable of introducing feed into the reactor at rates varying from less than 1 lb/hr to around 20 lb/hr. For proper operation, the gas flow through the reactor at the smallest feed rate desired should produce a gas velocity equal to or greater than the minimum fluidization velocity of the reactor while the highest expected velocity will not produce a velocity high enough to cause the reactor to experience slugging.

Using the data obtained by Prasad (1986) it is possible to calculate the amount of steam that must be introduced into the reactor in order ot obtain the optimal product gas. The ratio of steam to biomass obtained is 2.0. Assuming all the biomass gasifies and that the gases approximate ideal behavior it is possible to determine the volumetric flow rate through the reactor as a function of the feed rate.

$$Q = \frac{\left(\frac{F}{20\ \text{g/mole}} + \frac{2F}{18\ \text{g/mole}}\right) \left(0.0821\ \frac{\text{atm liter}}{\text{mole K}}\right) (1089\ \text{K})}{10\ \text{atm}} \qquad \text{(A1)}$$

where Q = volumetric gas flow rate, liter/s

F = feed rate, g/s

Note that the equation assumes that the combustor/pyrolyzer stage of the reactor will operate at 10 atm. and 1400 ^{o}F.

Hsu (1979) found that for the existing indirect liquefaction system, the correlation proposed by Richardson and Zaki was the most successful equation for predicting the minimum fluidization velocity. The correlation is

$$U_{mf} = (\ {}^{n}_{mf}) \ U_t \qquad (A2)$$

where U_{mf} = minimum fluidization velocity, cm/s

$_{mf}$ = bed voidage at minimum fluidization

n = index proposed by Richardson and Zaki, f(Re)

U_t = Terminal settling velocity of the particles, cm/s

For the system of interest the correlation gives a minimum fluidization velocity of 1.83 cm/s. Combining this value with equation (A1) it is possible to obtain an expression for the cross sectional area of the reactor as a function of the feed rate.

$$A = 787.13 \ F_{min} \qquad (A3)$$

where A = reactor cross-sectional area, cm^2

F_{min} = Minimum feed rate, g/s

The minimum flow is not the only concern in determining the diameter of the reactor. It is also necessary to consider the maximum flow that can be used without the the bed begining to exhibit slugging. Slugging is the condition where voids whose diameter is equal to that of the reactor rise though the bed. Each void is separated from the previous one by an area of solids. When slugging occurs very little gas-solids contacting occurs. There are many different equations for predicting the occurrence of slugging. One proposed by Cheremisinoff and Cheremisinoff (1984) is :

$$U_s = 0.8 \frac{U_{mf}}{L^{0.8}} \quad \text{(A4)}$$

where U_s = the minimum gas velocity for slugging to occur, ft/s

U_{mf} = the minimum gas velocity for fluidization to occur, ft/s

L = the bed height, ft

Substituting the values for the bed height and minimum fluidization velocity derived earlier it is possible to arrive at a value of 7.15 cm/s for U_s. Thus it is possible to arrive at an equation for the cross-sectional area of the reactor based on the maximum feed rate.

$$A = 201.46\ F_{max} \quad (A5)$$

where A = the cross-sectional area of the reactor, cm^2

F_{max} = Maximum feed rate, g/s.

Between equations (A3) and (A5) a reactor diameter of 22.0 cm was determined as the most appropriate. This diameter allows a minimum flow of 4.1 lb/hr and a maximum of 15.0 lb/hr. This diameter was converted to 9 in. for practical considerations.

The diameter of the combustor was much easier to determine since the flow rate of gases through the combustor is independent of the feed rate. The only requirements for the combustor are that there be sufficient room between the inner and outer walls for the installation of the needed internal components and sufficient volume to provide a long enough particle residence time to regenerate a catalyst. The regeneration time is dependant on the catalyst used and thus can not be generalized with any precision. It was determined that the volume of the combustor be such that the residence time would be twice that of the pyrolyzer. Combining the two requirements a final diameter of 24 in. was determined.

FISCHER-TROPSCH STAGE

The height of the Fischer-Tropsch reactor is dependant on the type of bed expected. A slurry bed must be considerably taller than a fluidized bed to obtain the same conversion. From figure 37, a height of 160 cm would be sufficient to produce significant conversions in a fluidized bed reactor while figure 42 shows that a height of nearly 7.0 m. would be needed for a slurry reactor. A variable height reactor that could handle bed heights between 50 cm and 300 cm was selected.

The diameter of the Fischer-Tropsch reactor was determined for a fluidized bed reactor since the requirements of a fluidized bed a more restrictive than those of for a slurry reactor. The gas flow arriving at the Fischer-Tropsch reactor will be determined by the flow rate leaving the pyrolyzer, the temperature of the scrubber and the composition of the gas stream leaving the scrubber. The model of the scrubber predicts that the temperature effects will be significantly more important that the composition changes. The following equation can thus be used :