at matching the trend in bubble velocities with height. As expected however the correlation predicts a larger velocity than was observed. Grace and Harrison (1969) found that in general the gas velocity in the bubbles tends to be larger than the visible bubble velocity. The values predicted for the velocity of the gas in the bubbles were found to be 1.8 times larger than the bubble velocity observed in the simulator. This compares favorably with the predictions of Sit and Grace (1981)

Operationally the simulator demonstrated several important points. It is unlikely that it will be possible to control the entire reactor system by hand. The interplay of all the operational variables is too complex. The pressures of the reactors control the gas flow as well as the the direction of any gas crossflow. However the flow rates have a strong effect on the pressures as well. The extensive interconnection of the operating parameters is a major characteristic of the staged reactor. The transfer tubes tend to isolate the combustor and the pyrolyzer to a degree, but the effect is not large.

The size of the transfer tube that directs the flow into the scrubber must be chosen with care. It must be as large as possible in order to minimize the pressure drop across the scrubber. The pressure drops across each of the distributor plates were sufficient to make operation of the entire plexiglass reactor difficult. This would not be as critical in an actual reactor for steel is not as flexible as plexiglass. However, an excess pressure drop would increase the operational difficulties. The scrubber design seems to be quite effective otherwise. The cyclone action removes the majority of the fine particles present in the gas stream even without the water spray. When actual water jets are installed they should be angled toward the wall instead of straight down. The majority of the initial flow occurs at the walls of the scrubber.

The Fischer-Tropsch section can not be much more than 200 cm to 300 cm tall if a fluidized bed is to be used. Beyond that height the bubbles in the bed have coalesced into a single slug. This slugging severly decreases the contact between the reactive gases and the solid catalyst. In addition, when the initial bed is 127 cm (50 in.) tall the expanded bed is 165 cm (65 in.) high and the maximum height reached by the solid particles is 190.5 cm (75 in.). Thus half the reactor would have to be used as freeboard in order to insure that the catalyst is not blown out of the reactor in very tall beds.

SENSITIVITY STUDIES

The behavior of the models in response to variations in several key parameters was determined. Since computer models are inexpensive and simple to run each variable was examined independently of all the others. When a given parameter was under investigation all the other parameters were held constant at the values which represent the behavior of the proposed staged reactor.

Pyrolyzer.

The values of various parameters associated with the fluidized bed model are shown in Table 6. The model for the pyrolyzer indicated that the reactions occur very rapidly with equilibrium usually being achieved within the first ten centimeters of the bed. The concentration profiles of various key gases are shown in Figure 33. Experiments on the current indirect liquefaction system tend to confirm this finding. (Prasad, 1986) The high steam content of the output stream is due to the fact that steam is used to fluidize the bed. The steam was introduce into the reactor at such a rate that the steam to biomass

Table 6. Values calculated using the correlations for small scale phenomena occuring in a fluidized bed.

Parameter	Value	Principle Independent Variable
U_{BS}	1.36×10^{-4} cm/s	B′ B
U_{CS}	1.66×10^{-4} cm/s	U_{BS}′ C′ C
U_{ES}	2.47 cm/s	U_o, U_{BS}, U_{cs}
U_B	9.93 cm/s	U_o, U_{mf}, D_B
F_{BC}	2.09 s^{-1} s^{-1}	U_{mf}, D_B
F_{CE}	7.19 s^{-1} s^{-1}	U_B, $_{mf}$' D_B, D_G
P_{BC}	4.41×10^{-6} cm/s	U_{BS}
P_{CE}	-5.48×10^{-5} cm/s	U_{CS}, P_{BC}
H_j	3.99 cm	D_P, A, N_D, U_o,U_{mf}
H	41.1 cm	H_{mf}, U_o, U_{mf}, D_{BM}

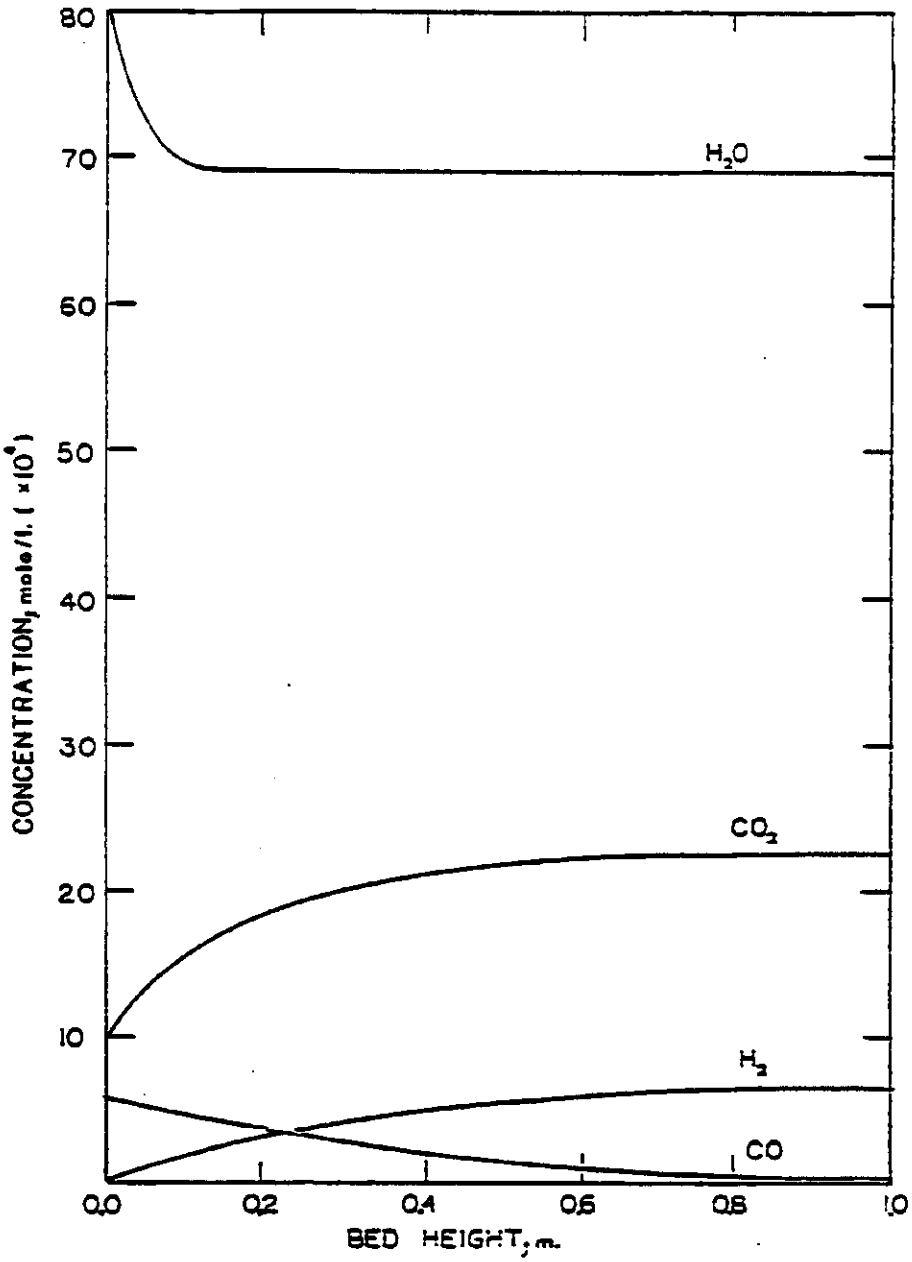

Figure 33. The concentration of several key gases at several heights in the pyrolyzer as predicted by the fluidized bed model.

weight ratio was 2.0 . This ratio was found to produce the optimal pyrolyzer performance by Prasad. (1986)

Three of the parameters tested in conjunction with the pyrolysis model were found to have no measurable effect on the gas compositions in the pyrolyzer. These parameters were : bubble diameter, bubble velocity, and total bed diameter. In the presentation of the effects of the parameters that had significant effects on the behavior of the reactor, the concentration of steam will be used as an indication of the magnitude of the effect. Steam participates in all the chemical interactions modeled. In addition, its high initial concentration will insure that none of the effects are masked by low concentration anomalies.

Figure 34 shows the effect the bubble to cloud mass transfer rate has on the concentration profiles. It should be noted that in all the concentration profiles the cloud phase shows an initial rapid drop in concentration. This is due to the fact that the model assumes that when the biomass gasifies all the resulting gases appear in the emulsion and cloud phases. The bubble phase initially contains only the fluidizing gas. However the circulation between the bubble and cloud phases soon equilibrates the concentrations. This equilibration is responsible for the initial drop in cloud phase concentration. Figure 34

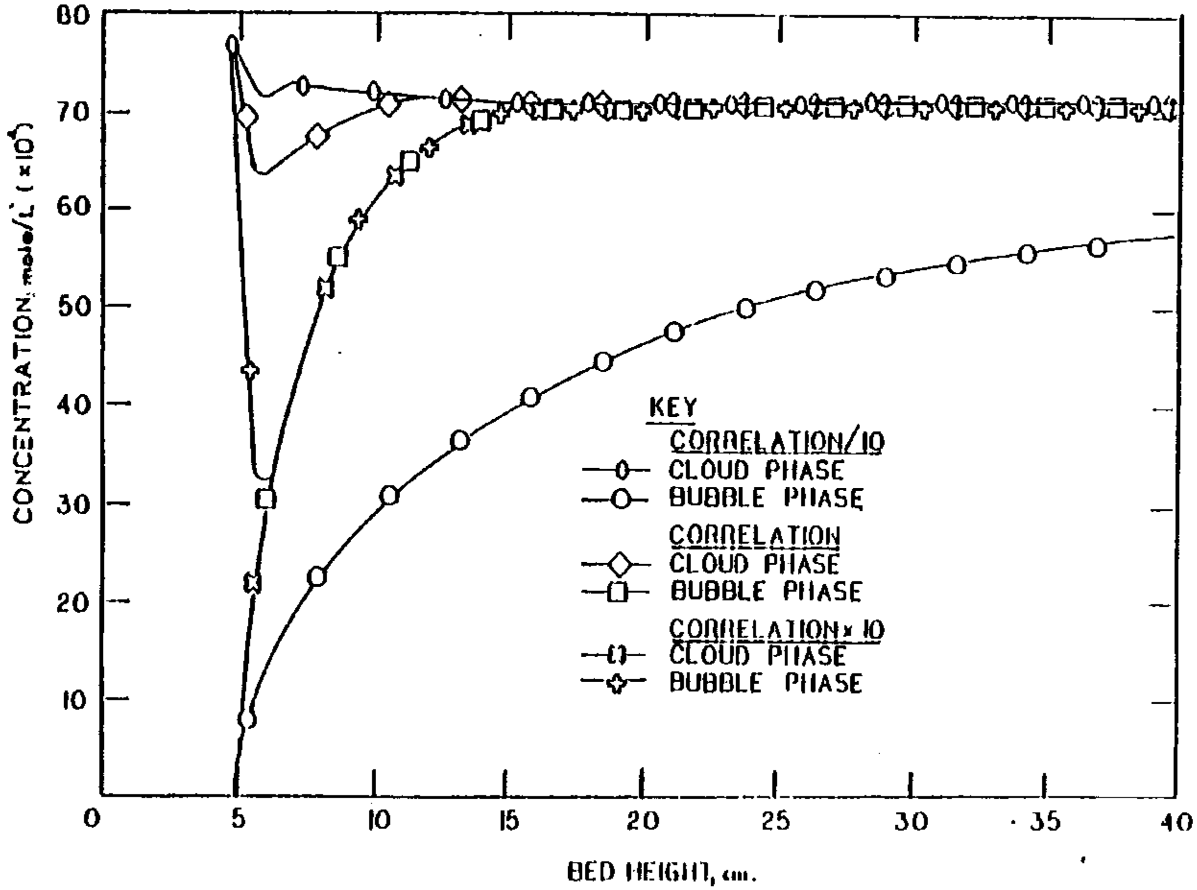

Figure 34. The effect of varying the bubble to cloud mass transfer rate from that obtained using Murry's correlation on water concentration in the gas stream.

indicates that the greater the transfer coefficient the more precipitous this initial drop is as the gaseous circulation is augmented by diffusion. At a transfer rate somewhat larger than the rate predicted by Murry's correlation (Peters et al., 1982) the concentrations become identical. On the other hand at very low rates, those approaching 0.1 of the predicted value, the bubbles and the cloud phase do not reach equilibrium in the 40 cm. bed. This indicates that it may be possible for components with small mass transfer rates to become segregated in the cloud phase. However, it should also be noted that none of the components examined had coefficients varying by more than a factor of two from those of any other component. The gases with the lowest coefficients were the longest hydrocarbon chains. These compounds are also the most likely to react in the emulsion phase. Thus none of the components were susceptible to significant segregation and those that showed the greatest tendency were the mostly likely to be broken down negating whatever difference there may have been.

The transfer rate from the cloud phase to the emulsion phase has a similar effect on the concentration profiles as is shown in Figure 35. At the highest rate tested, ten times Murry's correlation (Peters et al., 1982) value, the concentration of the emulsion and cloud are

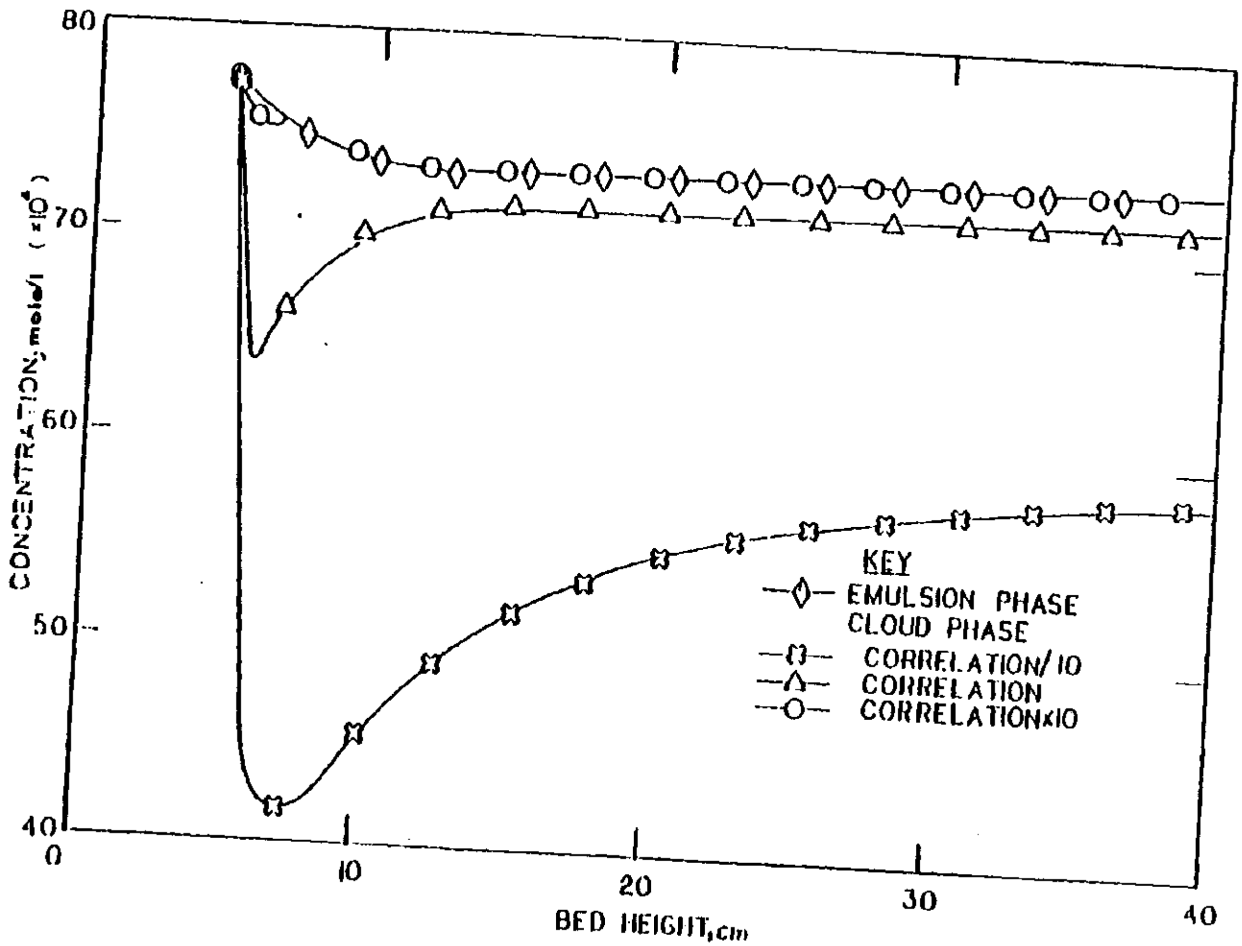

Figure 35 . The effect of varying the cloud to emulsion mass transfer rate on the water concentration in the gas stream.

indistinquishable. The initial drop in cloud concentration has even been eliminated. As the transfer rate drops, the concentrations separate with the cloud phase concentration moving toward that of the bubble phase. Unlike the the cloud to bubble transfer, it appears that the clouds and the emulsion reach equilibrium even at the lowest transfer rate. The transfer rates have little or no effect on the emulsion phase concentration due to its large volume compared with the other two phases.

As expected, the reaction rates used have a significant effect on the concentration profiles. This is unfortunate for the reaction rates are some of the least well known parameters. The rates involved in the gas phase reactions are comparatively well established, however. The key reaction studied was the water gas shift reaction,

$$CO_2 + H_2O \underset{k_1}{\overset{k_2}{\rightleftharpoons}} CO_2 + H_2 \qquad (38)$$

The effects of decreasing k_1 are shown in Figure 36. Varying the rate over approximately two orders of magnitude does not effect the attainment of equilibrium. The variation does, however, significantly increase the final steam content of the outlet gas. The effects of decreasing k_2 are shown in Figure 37. The concentration of steam in the outlet stream decreases with decreasing k_2 as would be

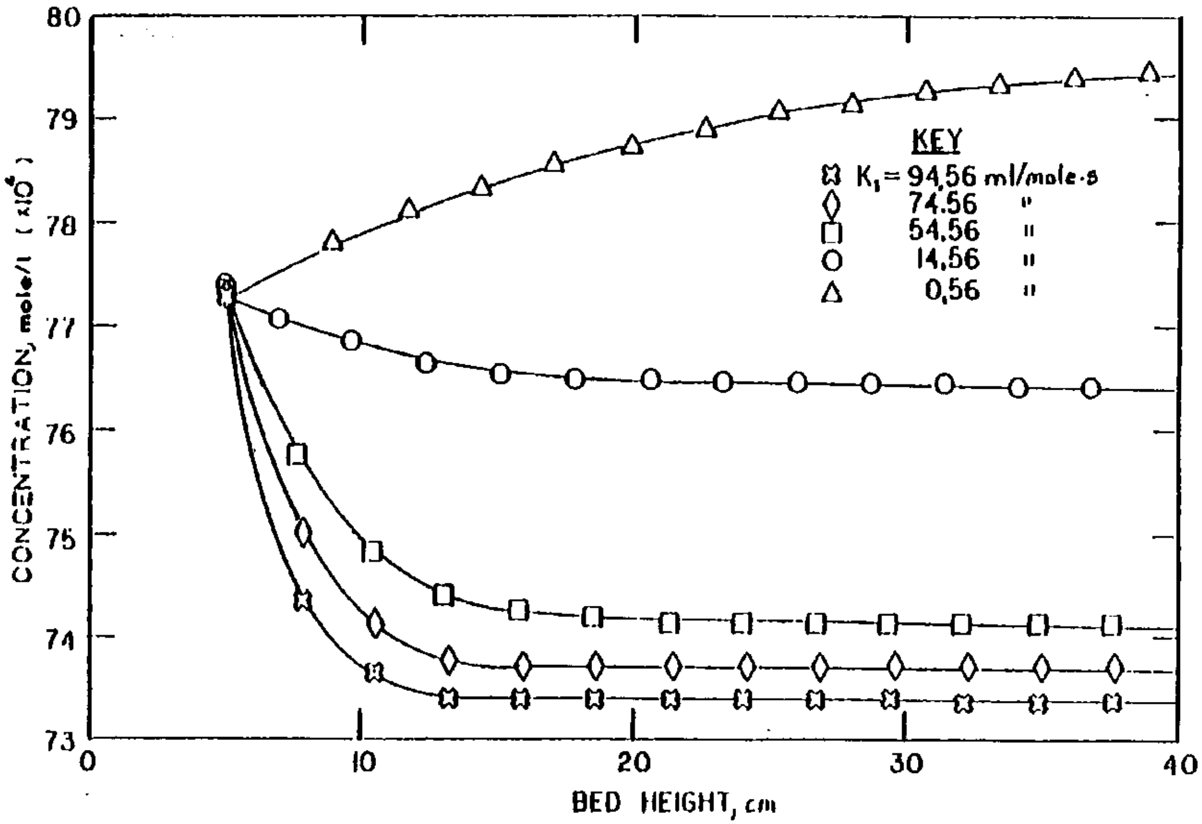

Figure 36. The effect the forward reaction rate constant has on the concentration of water vapor in the gas stream.

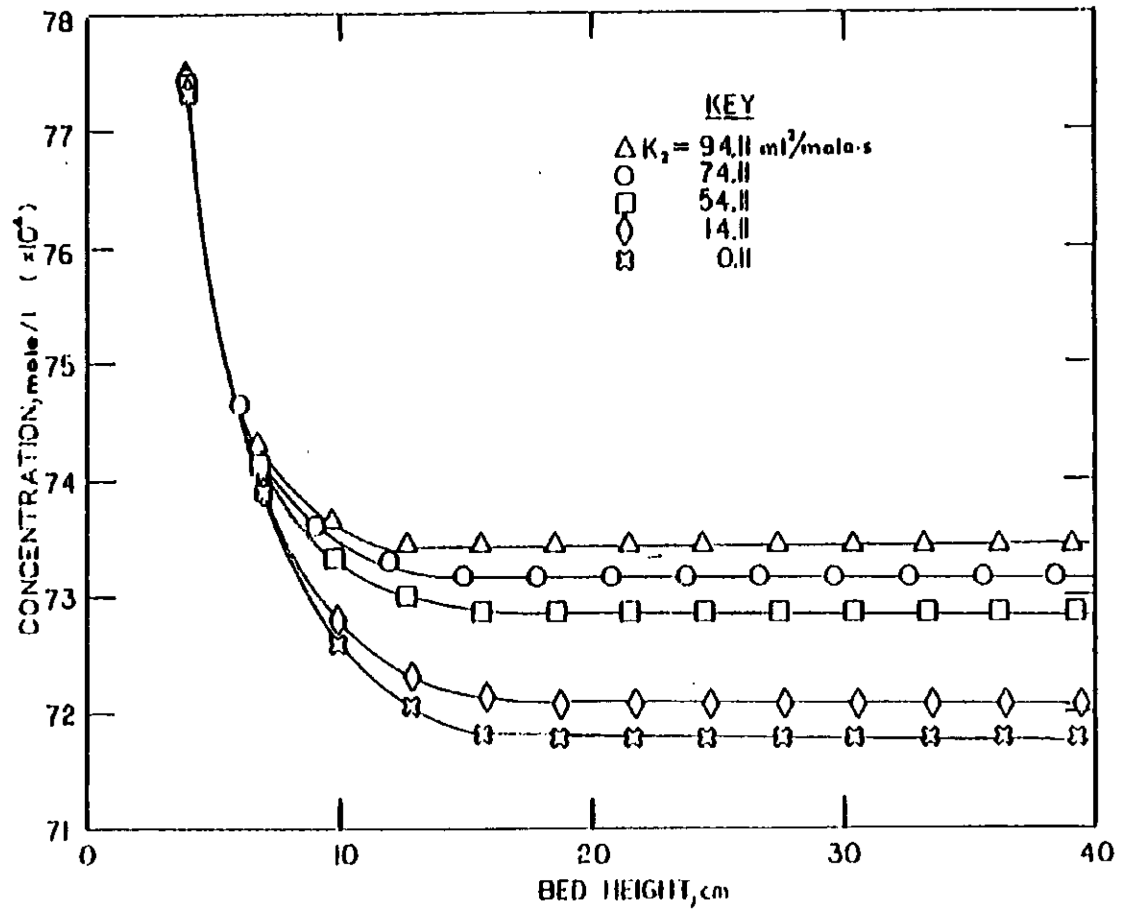

Figure 37. The effect of the reverse reaction rate on the water concentration profile.

expected. The effects are not as pronounced as those that result from decreasing k_1. This is due to the fact that the steam concentration is larger than any other concentration. Since the forward reaction is dependent on the concentration of water vapor the effects of changing that coefficient are magnified more than those of changing k_2 would be.

The final variable that was found to have a significant effect on the concentration profiles was the inlet gas velocity. Figure 38 shows the results of modifying the velocity. The results were not anticipated. The larger velocities resulted in a higher conversion than the smaller velocities. High gas velocities result in larger bubbles and lower residence times. Both these phenomena would tend to decrease conversion. However, increased velocities also result in increased transfer between the bubble and cloud phases. The reaction appears to be sufficiently fast for the transfer effects to offset the effects of increased bypassing.

Fischer-Tropsch Synthesis

Fluidized Bed. The model of the fluidized bed Fischer-Tropsch reactor showed that five of the variables tested had a significant effect on the behavior of the

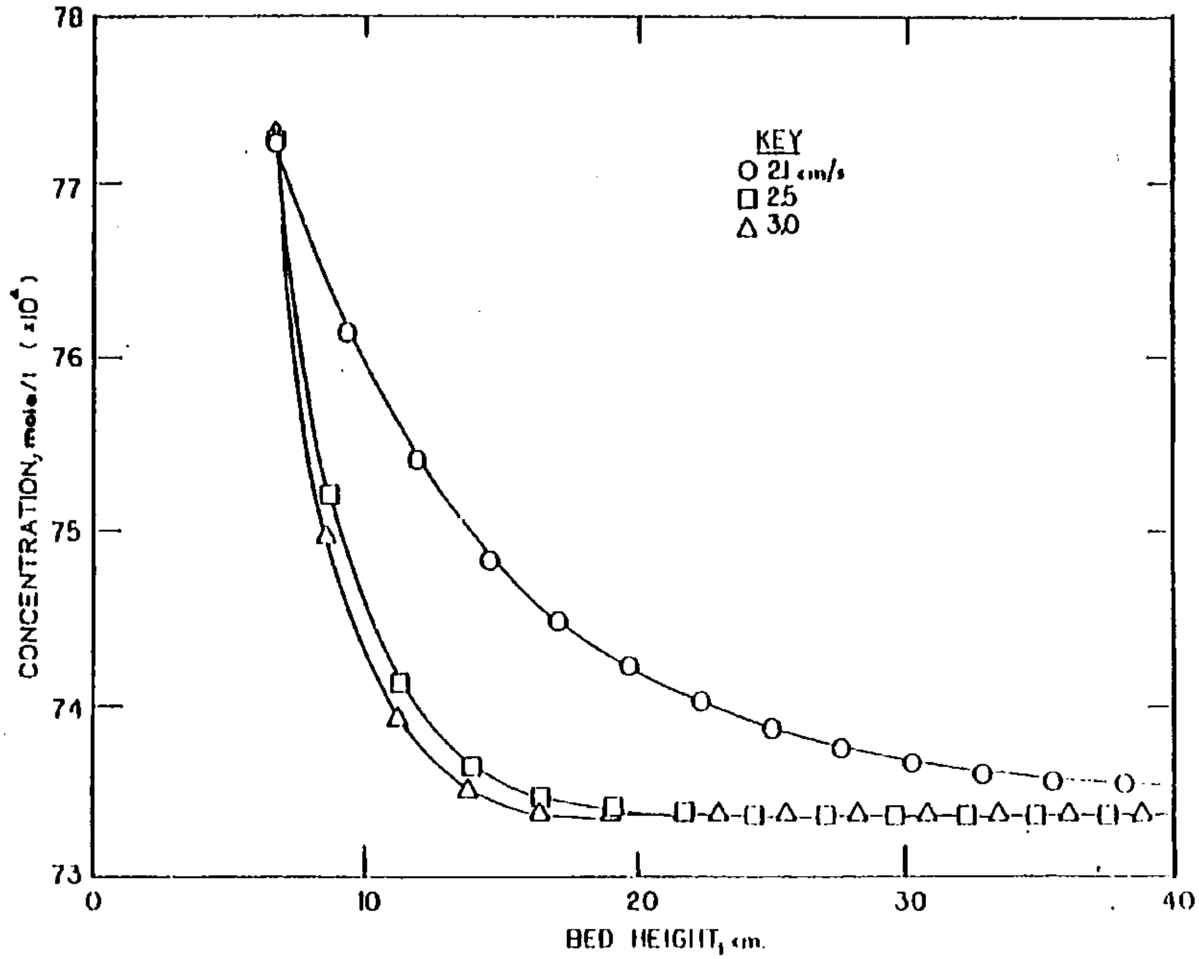

Figure 38. The effect the initial velocity has on the concentration profile of water.

reactor. These variables were the bed height, temperature, rate constant, bed diameter, and gas flow rate. The model exhibited one major deviation when compared to the simulator studies. The bed expansion predicted by the correlation proposed by Peters, Fan, and Sweeney (1982) was only 1 percent of the bed height while the actual bed expanded 30 percent in the taller beds. The correlation was accurate for shorter beds, however. This is to be expected due to the fact that it was developed using data from beds less than 80 cm tall. The actual bed expansion was substituted for the correlated value in the sensitivity studies.

Figure 39 shows the concentration profiles for hydrogen in each of the phases. Hydrogen is used as the key component because the rate expression proposed by Deckwer et al. (1982) predicts conversion based only on the hydrogen concentration. This allows the kinetics to be simplified. Due to the inaccuracies that involved in all fluidized bed models, using the simplified kinetics probably does not introduce a significant new error into the model. As would be expected, the concentration in the emulsion phase is the lowest and the concentration in the bubble phase, where no reactions occur, is the highest. All three concentrations are relatively close together at the beds planned operating conditions.

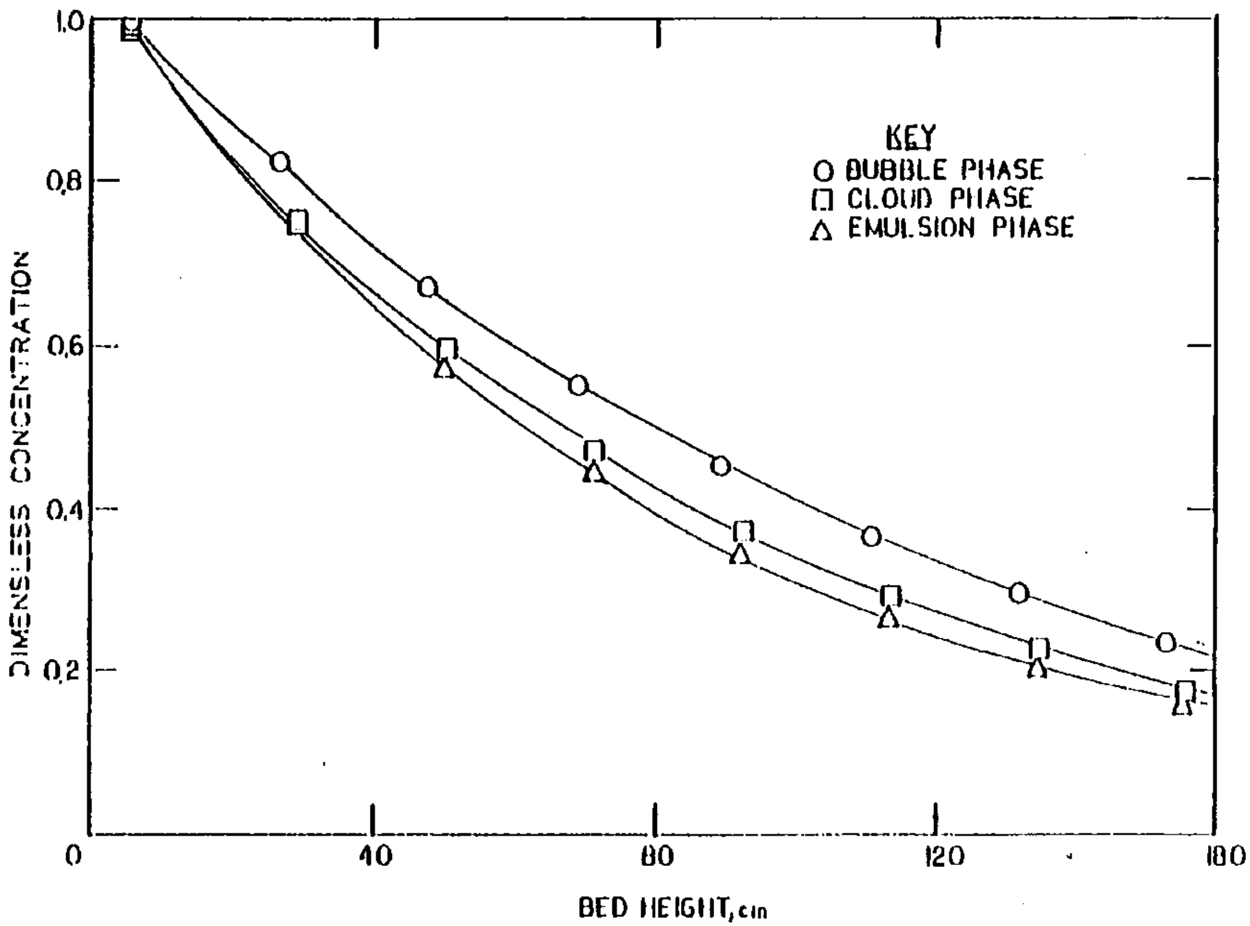

Figure 39. The concentration of hydrogen in each of the three phases in a fluidized bed Fischer-Tropsch reactor with an initial gas velocity of 2.6 cm/s.

The flow rate of the gas had a major effect on conversion. Unlike the pyrolysis reactor the effect was predictable. The higher flow rate, as shown in Figure 40, resulted in a lower conversion. It also tended to exaggerate the concentration differences between each phase. With the bubbles moving more rapidly, the gas at any point in the bed in the bubble phase had a higher than normal concentration of hydrogen. The slower Fischer-Tropsch reaction is probably the main cause of the difference in behavior between the Fischer-Tropsch reactor and the pyrolyzer. In general it appears the the Fischer-Tropsch reaction carried out in a fluidized bed is a reaction rate limited process while the pyrolysis is a mass transfer limited process.

The effect of the reactor temperature was also large. The temperature mainly effected the reaction rate. Figure 41 shows the effects of varying the temperature over a 100 K range. At 550 K it was possible to reach nearly complete conversion in the reactor. It is important to note that the rate expressions in particular and the model in general do not attempt to predict the nature of the product that the reactor produces. The temperature will have a significant effect on the nature of this product. Studies conducted by Wang (1980) on the Fischer-Tropsch synthesis indicate that temperatures above a given level

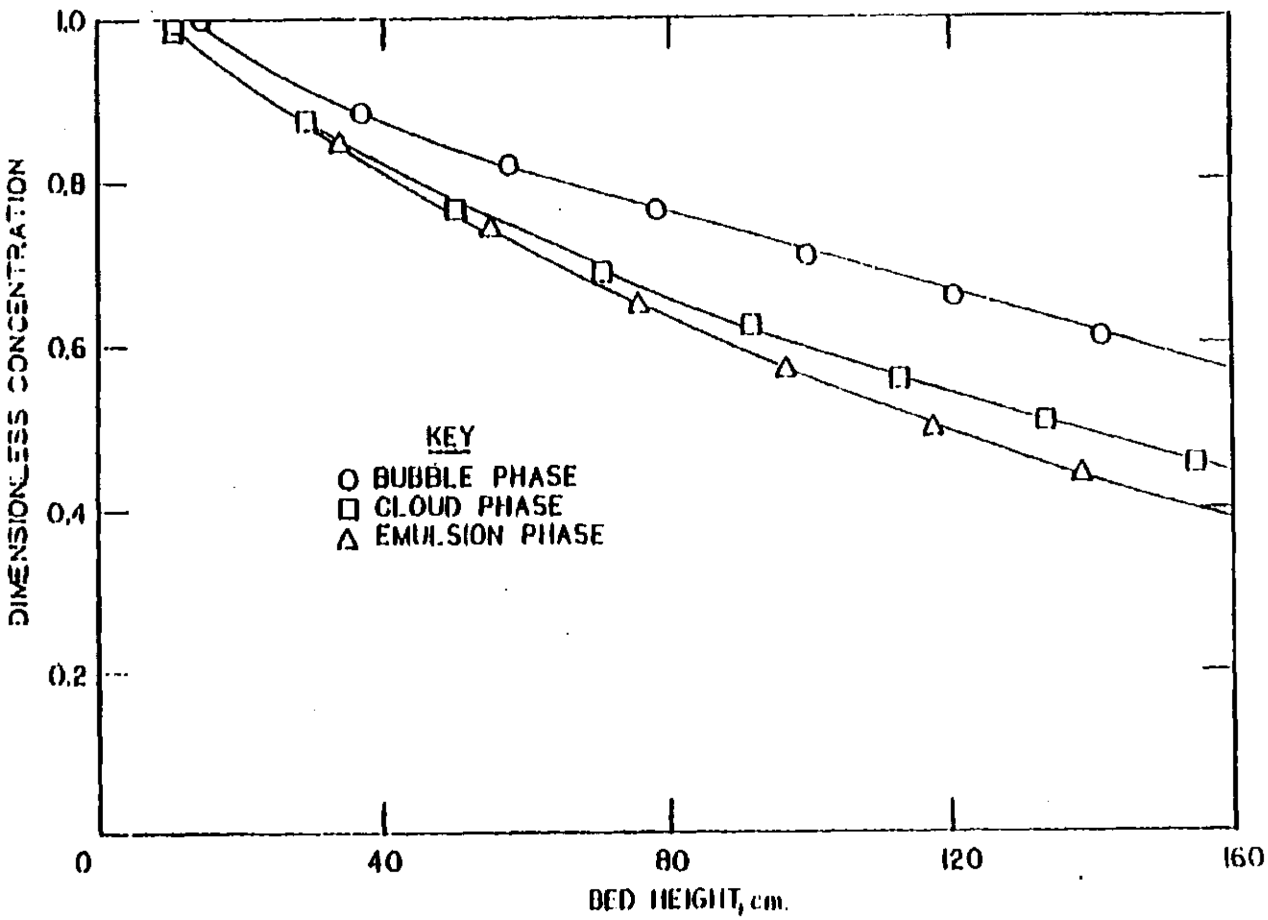

Figure 40. The concentration of hydrogen in each of the three phases in a fluidized bed Fischer-Tropsch reactor with an initial gas flow rate of 5.0 cm/s.

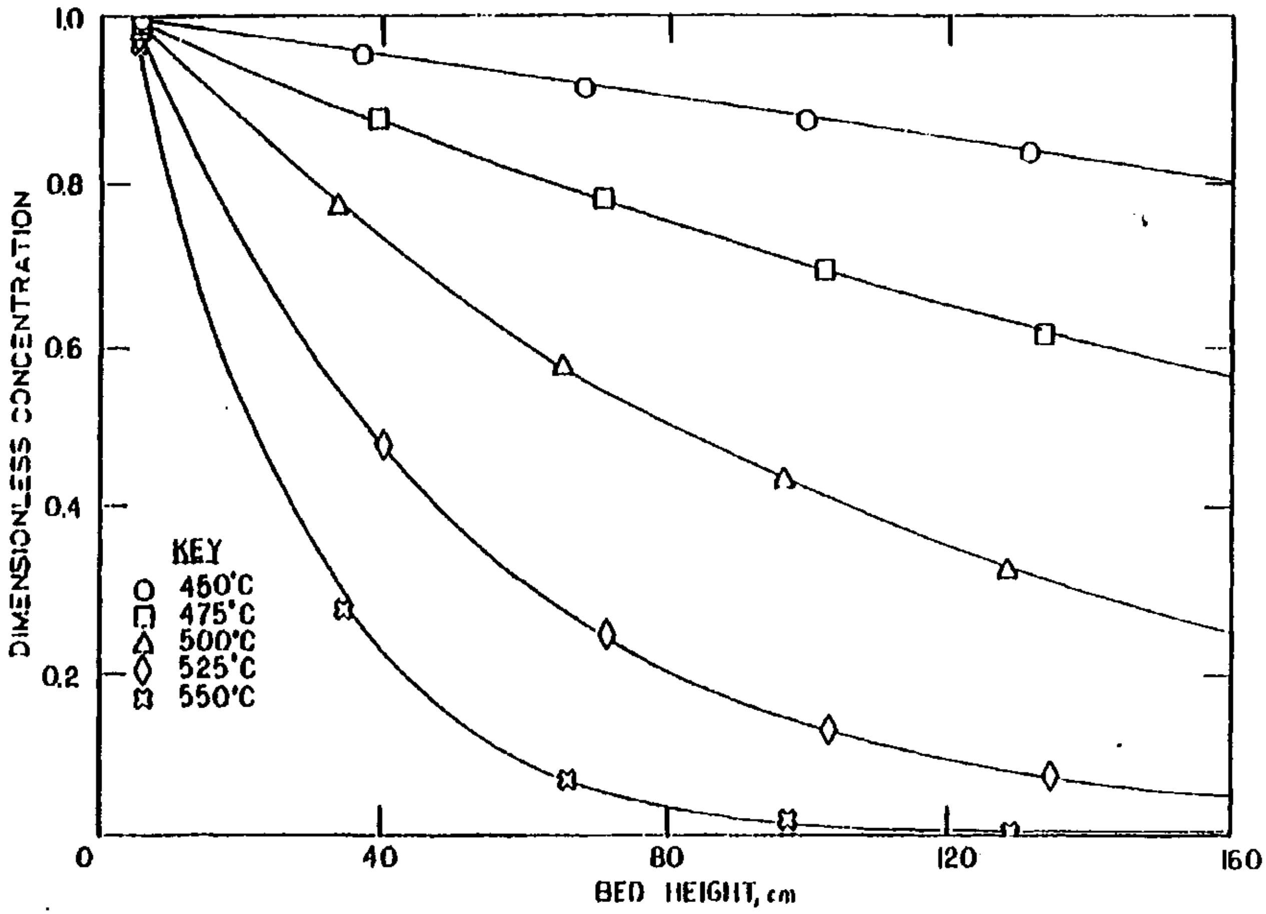

Figure 41 . The effect of reactor temperature on the hydrogen concentration profiles.

will shift the product distribution to shorter hydrocarbons. The optimal temperature was found to be 533 K.

The rate of reaction was varied directly. The results of the variations are shown in Figure 42. As would be expected they are very similar to the temperature effects. From a comparison of Figure 41 and Figure 42 it is possible to estimate the magnitude of the effect that temperature has on the rate constant. A variation of 25 K produces approximately the same effect as 50 percent change in the reaction rate.

Finally the effects of varying the reactor diameter are shown in Figure 43. The smaller diameter reactor exhibited the best conversion. The reactor diameter is important in determining the size of the bubbles that are formed in the reactor. The smaller reactors tend to produce smaller bubbles. This in turn enhances the transfer of gas from the bubbles to cloud and emulsion phases where the reactions can take place. The effect is not very strong due to the fact that the reaction rate is more important in determining the the overall conversion than the mass transfer considerations are for this reactive system. It does point out however that the mass transfer considerations can not be neglected.