FIXED BED REACTION TESTS

Gas Feed Only and Gas-Liquid Co-feed

Prior to putting the liquid fluidized bed bench scale reactor on-stream, several preliminary process variable scans were performed in a fixed bed reactor in order to: (1) test the recommended method of catalyst reduction, (2) establish the desired regions of reaction conditions, and (3) qualitatively compare the relative reactivity of three catalyst candidates; Catalyst and Chemicals Inc. C-79-1-01, Catalyst A*, and Catalyst B*.

The first catalyst examined in this series of runs was the CCI catalyst. The 1/4" Ø x 1/8" particles were loaded and reduced according to the manufacturer's recommended procedure. Because of the small size of the catalyst mass and the large capacity for heat removal afforded by the molten salt bath used for controlling the reactor temperature, an undiluted simulated Lurgi coal gas was fed directly to the catalyst bed without incurring an excessive temperature rise. The nominal feed gas composition was CO - 25 percent, H_2 - 50 percent, CO_2 - 10 percent, and CH_4 - 15 percent. The results presented in Table 3-5 are limited because a frozen heater relay caused the reactor temperature to rise above 400°C, deactivating the catalyst. However, one can see that even the undiluted feed gas can be reacted to near equilibrium with this catalyst and reactor.

The results for the Commercial Methanol Catalyst A (crushed to a -5 +12 mesh prior to reduction) are presented in Table 3-6. The first two runs were conducted in order to have a basis for comparison with the CCI catalyst vapor phase tests. While the commercial catalyst appears to be much more reactive, this might be partly due to its smaller particle size.

* Confidentiality agreements with catalyst suppliers prohibit any more detailed identification.

TABLE 3-5

METHANOL SYNTHESIS IN A FIXED BED: LURGI FEED GAS
CCI CATALYST:* C-79-1-01 (1/4" O x 1/8")

Run No.	Temperature °C	Pressure psig	VHSV Hr^{-1}	CO Conversion to MeOH, %	Equilibrium Conv. to MeOH, %
1	275	525	2,600	14.5	15.0
2	271	510	4,320	12.8	16.5
3	258	510	4,200	12.5	21.0
4	255	920	4,880	25.0	50.0

* 29 ± grams = 25 cm^3.

TABLE 3-6

METHANOL SYNTHESIS IN A FIXED BED: LURGI FEED GAS

COMMERCIAL METHANOL SYNTHESIS CATALYST A[1] (-5+12 MESH)/
WITCO 40 MINERAL OIL

Run No.	Temperature °C	Pressure psig	VHSV Hr^{-1}	CO Conversion To MeOH, %
1	250	530	2,060	29.0[3] (Vapor Phase)
2	256	530	3,690	24.2[4] (Vapor Phase)
3[2]	252	530	3,830	10.0
4[2]	270	530	3,690	9.4
5[2]	230	530	3,980	9.8
6[2]	255	530	2,850	14.4
7[2]	248	530	1,760	18.6
8[2]	248	530	1,390	19.9
9[2]	258	760	3,170	16.5
10[2]	256	760	1,890	24.0
11[2]	255	760	1,190	30.0
12[2]	208	530	1,060	20.0
13[2]	207	530	1,640	13.5
14[2]	209	530	2,730	10.4

(1) 62.0 grams = 52 cm^3.

(2) Cofeed liquid stream of Witco 40 mineral oil.

(3) Equilibrium conv. = 29.5 percent.

(4) Equilibrium conv. = 26.5 percent.

The balance of the runs were conducted with a slow, concurrent once-through stream of Witco 40 mineral oil in addition to the feed gas. While the liquid flow was insufficient to fluidize the catalyst bed, it did provide qualitative indications as to the liquid phase effects. This limitation not withstanding, these results (see Figures 3-8, -9 and -10) show approaches to equilibrium on the order of 30-75 percent.

Commercial Methanol Catalyst B was also crushed to -5 +12 mesh prior to reduction and the results for these runs are presented in Table 3-7 and Figure 3-11. The initial catalyst charge was lost after seven runs due to a frozen temperature relay. Nevertheless, the process scan was continued with a second charge of catalyst. At the lower pressure level of 37 atmospheres, the results appear independent of operating temperature over the range of 225-275^{o}C; however, at a higher operating pressure (54 atm), the higher operating temperature (250^{o}C vs. 225^{o}C) appears to result in measurably greater reaction rates. This relative insensitivity to operating temperature might be due to the balancing effects of greater reaction rate and decreased driving force to equilibrium with increasing temperature.

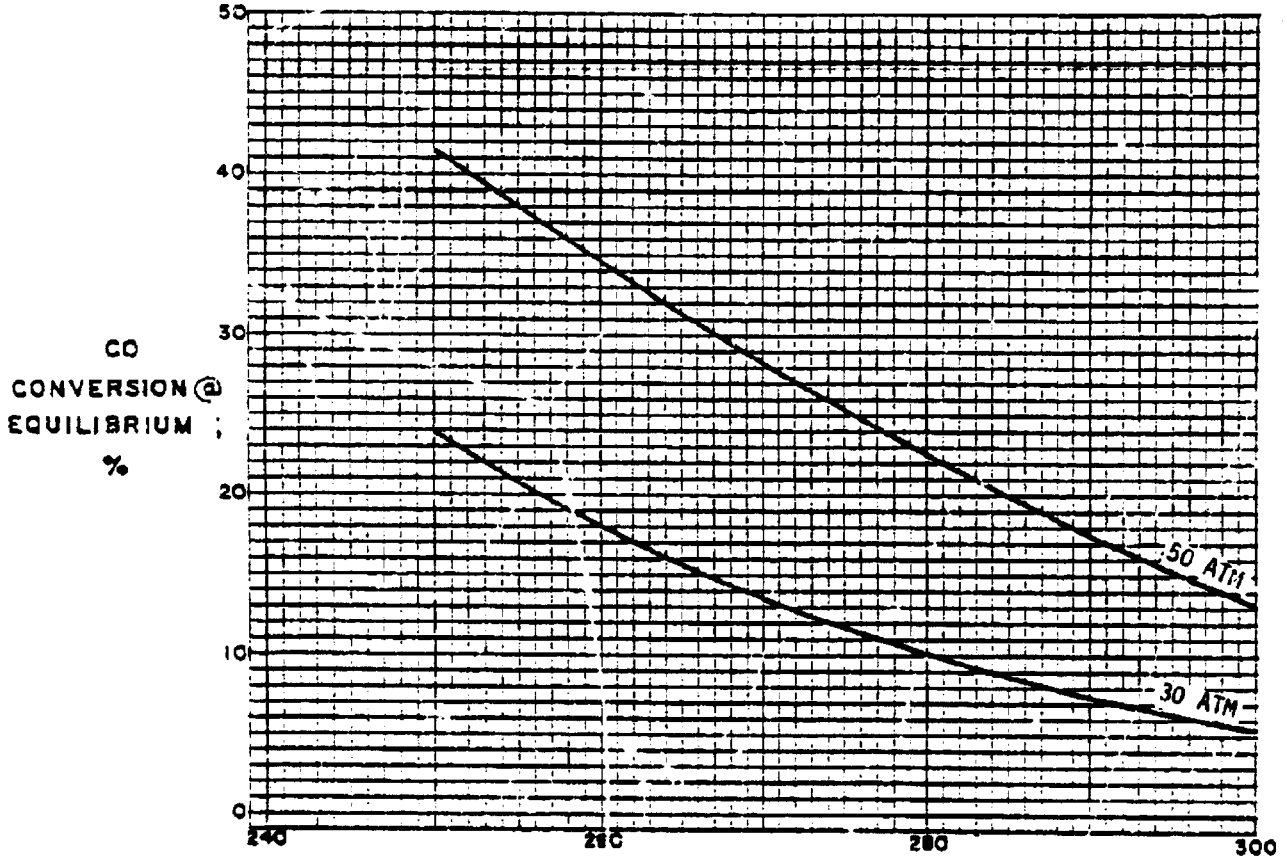
In general, Commercial Catalyst B has similar reaction properties compared to Commercial Catalyst A. Comparison curves are given in Figure 3-12 for both catalyst preparations. The different shapes for the conversion curves are more than likely the result of an insufficient number of data points for the earlier reaction scans with Catalyst A.

At the end of the process scan with the Lurgi feed gas, several runs were performed using a Koppers-Totzek feed gas and while the absolute CO conversions were lower, the actual methanol productivity was essentially equivalent because of the high relative CO feed concentration of a Koppers-Totzek gas.

FIGURE 3-8

THEORETICAL EQUILIBRIUM CO CONVERSION TO METHANOL

LURGI FEED GAS AFTER ACID GAS AND WATER REMOVAL



FEED COMPOSITION

COMPONENT	MOLE %
N_2 , Ar	0.41
CH_4	14.17
C_2 , $C_2^=$	1.29
CO	25.01
CO_2	9.26
H_2	49.83
H_2O	0.23

FIGURE 3-9

THEORETICAL EQUILIBRIUM
CO CONVERSION TO METHANOL

LURGI FEED GAS
AFTER ACID GAS AND WATER REMOVAL

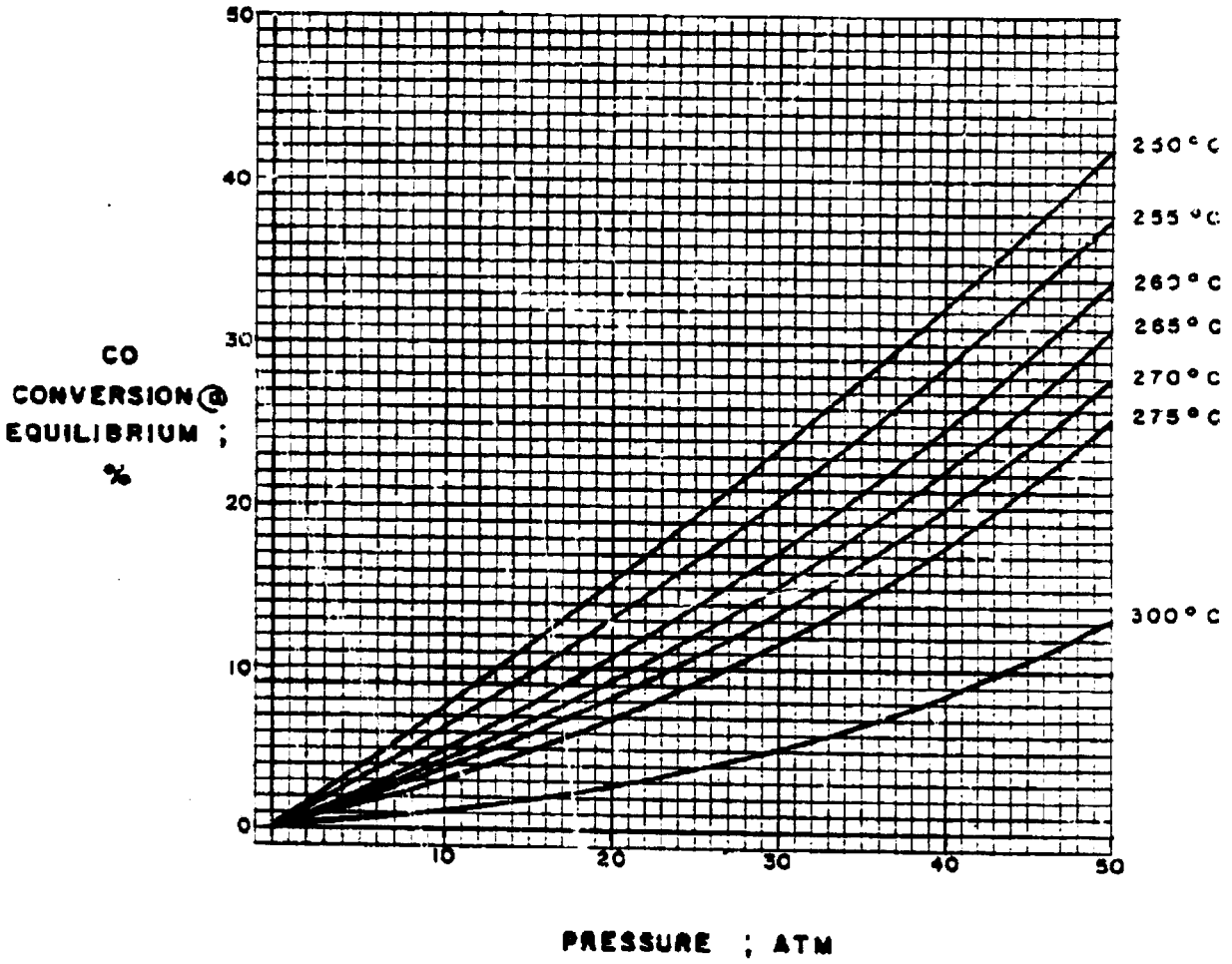
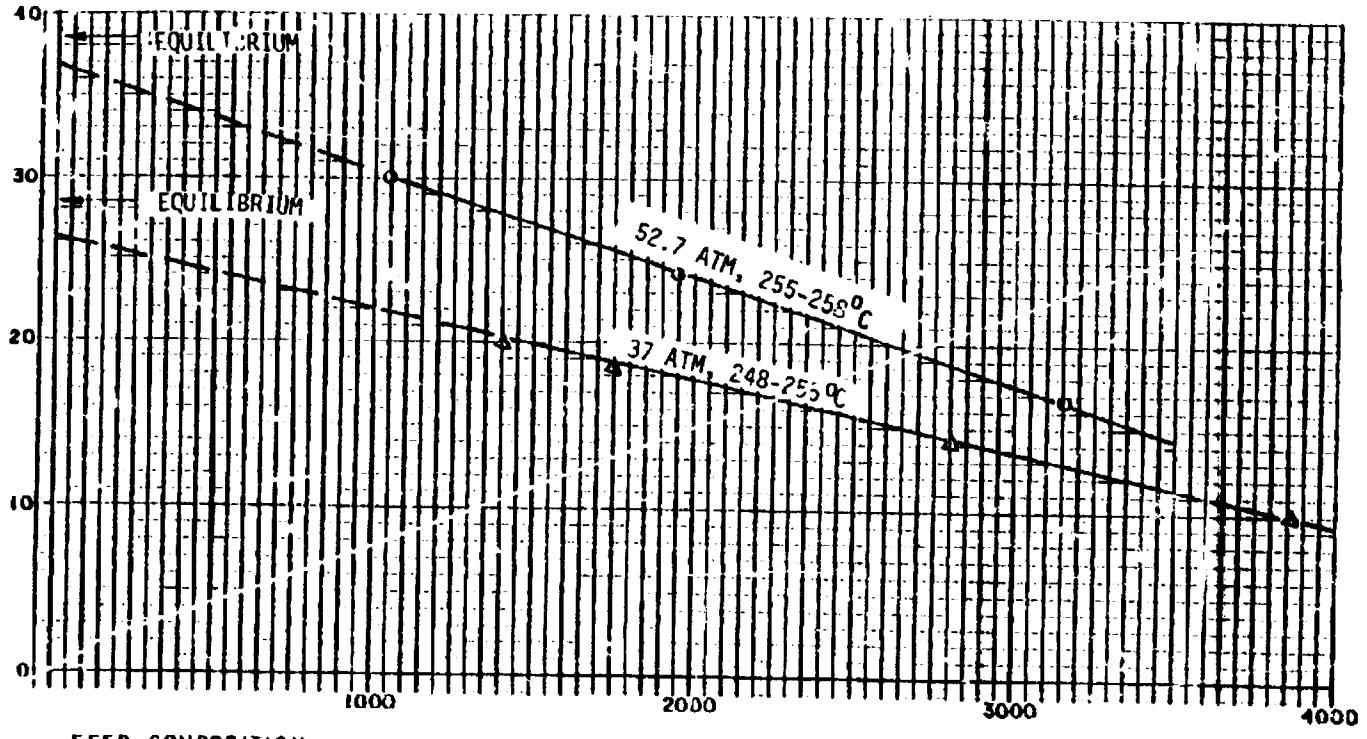


FIGURE 3-10

FIXED BED REACTOR CONVERSION — COMMERCIAL CATALYST A

SIMULATED LURGI FEED GAS

WITCO 40 MINERAL OIL CO-CURRENT FLOW



FEED COMPOSITION

COMPONENT	MOLE %
CH_4	15.0
CO	25.0
CO_2	10.0
H_2	50.0

TABLE 3-7

METHANOL SYNTHESIS IN A FIXED BED
SIMULATED LURGI AND KOPPERS-TOTZEK FEED GASES
COMMERCIAL CATALYST B (-5+12 MESH)/WITCO 40 MINERAL OIL

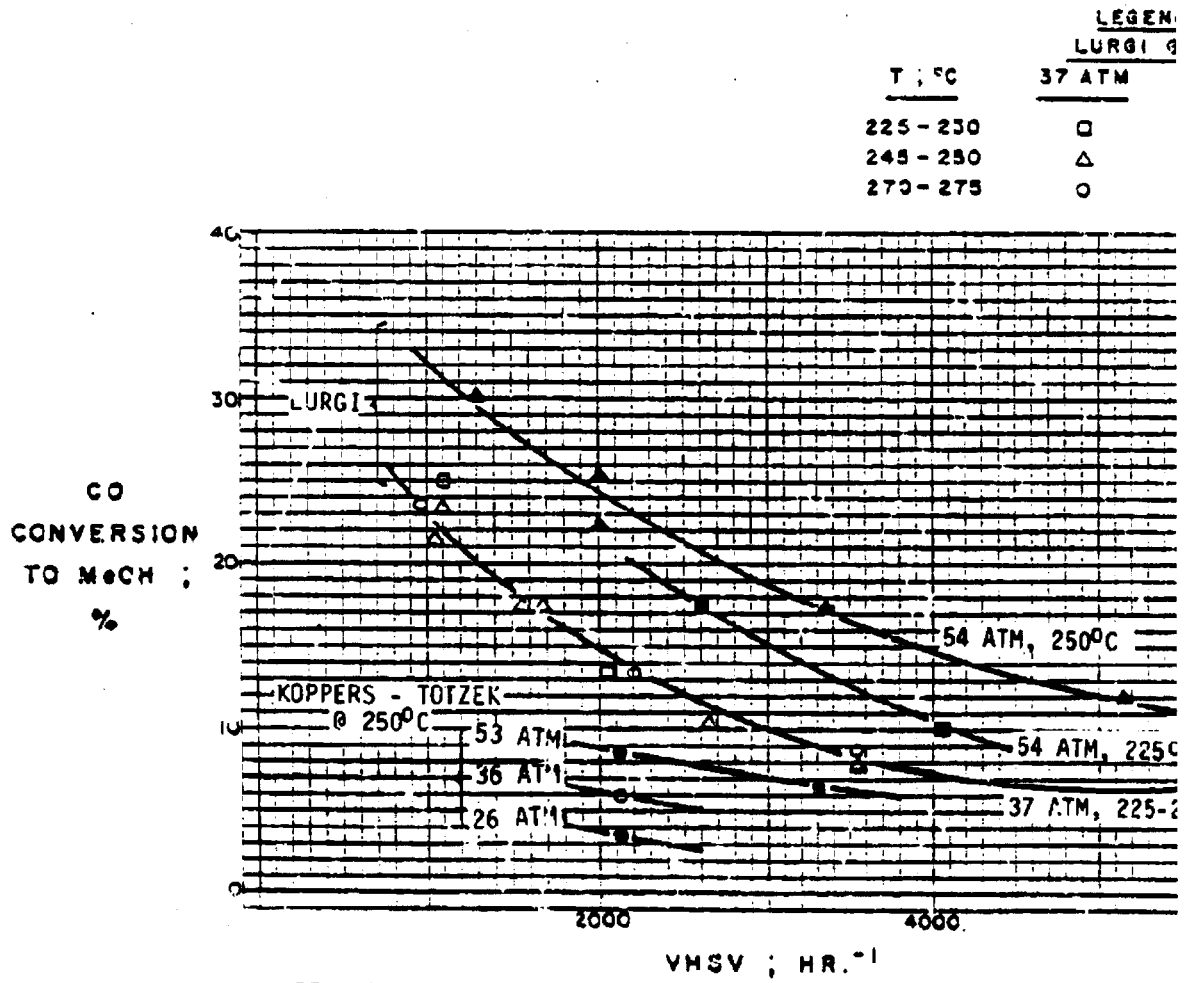
Run No.	Temperature °C	Pressure Atm	VHSV Hr^{-1}	CO Conversion To MeOH, %
	Catalyst Charge No. 1 = 65 gm			
A1	245	37	1,680	17.5
A2	245	37	1,040	21.3
A3	245	37	2,660	10.4
A4	245	37	3,520	7.6
A5	272	37	3,520	8.7
A6	272	37	2,200	13.3
A7	272	37	950	23.7
	Catalyst Charge No. 2 = 65 gm			
B1(1)	246	37	2,460	36.1
B2(1)	246	37	3,560	33.3
B3(1)	246	37	4,900	30.3
B4(1)	247	37	5,560	30.0
B5	250	37	5,500	7.5
B6	248	37	1,560	17.4
B7	252	37	1,100	23.2
B8	225	37	1,100	25.0
B9	252	54	3,360	16.8
B10	252	54	2,000	25.1
B11	252	54	5,160	11.8
B12	225	37	2,020	13.4
B13	225	54	2,600	17.2
B14	225	54	4,040	9.8
B15	248	54	2,000	22.3
B16	248	54	1,300	29.8
B17(2)	247	54	2,100	8.7
B18(2)	247	54	3,300	6.8
B19(2)	248	36	2,100	6.0
B20(2)	247	26	2,100	3.7

(1) Gas flow only.
(2) Koppers-Totzek Feed Gas.

FIGURE 3-11

FIXED BED REACTOR
COMMERCIAL CATALYST B

WITCO 40 MINERAL OIL
SIMULATED LURGI AND
KOPPERS-TOTZEK FEED GASES



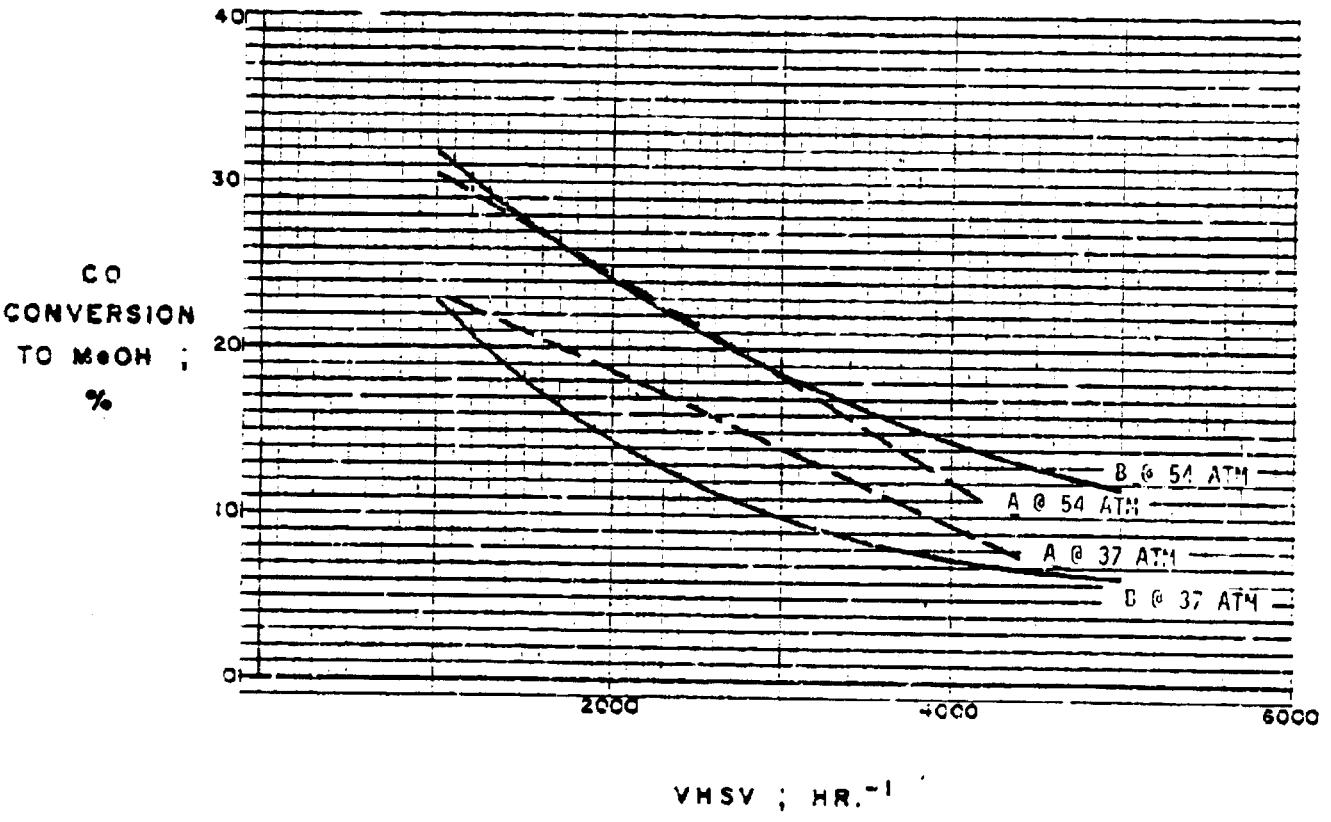
FEED COMPOSITION

COMPONENT	MOLE % LURGI	MOLE % KT
CH_4	15	—
CO	25	60
CO_2	10	2
H_2	50	38

FIGURE 3-12

FIXED BED REACTOR
COMPARISON OF COMMERCIAL CATALYSTS A & B

250°C



Recirculating Liquid

For a short period of time, a bench scale reactor equipped with a circulating liquid pump was available for experiments. Unfortunately, the system was not capable of sufficient heat input at the liquid flows required to fluidize the catalyst. Nevertheless, it was the first time the effect of continuously recirculating the liquid around the synthesis loop was investigated. The results for these experiments are shown in Table 3-8 and Figure 3-13. The scan covered two temperatures and two circulating liquid flow rates. In general, the results agreed with those obtained from the fixed bed reactor (compare Figures 3-11 and 3-13). Increased liquid circulation rate increased productivity at the 270°C level. There was no measurable effect of liquid flow at the 250°C level.

The main results of these initial tests illustrated that:

- Methanol product purity remains consistently high, even at conversion levels of 30 percent per pass.
- On a once-through basis, a high CO concentration feed gas (Koppers-Totzek) performs as well as a more typical methanol synthesis feed gas (Lurgi).
- Operability of commercial methanol catalysts in the presence of a liquid phase, both circulating and once-through was confirmed.

TABLE 3-8

METHANOL SYNTHESIS IN A "RECIRCULATING LIQUID BED" - SIMULATED LURGI GAS
COMMERCIAL CATALYST B[1]/WITCO 40 MINERAL OIL

Run No.	Temperature °C	Pressure Atm	VHSV Hr^{-1}	Liquid Flow Rate	CO Conversion To MeOH, %
C1	250	35	5,050	Low	6.15
C2	250	35	3,690	Low	9.30
C3	250	35	2,290	Low	14.92
C4	270	35	5,125	Low	6.52
C5	270	35	3,850	Low	7.69
C6	270	35	2,225	Low	10.10
C7	250	35	3,700	High	7.75
C8	250	35	2,225	High	13.50
C9	270	35	2,275	High	14.10
C10	270	35	3,825	High	8.31
C11	270	35	5,100	High	6.37

(1) 242 grams.

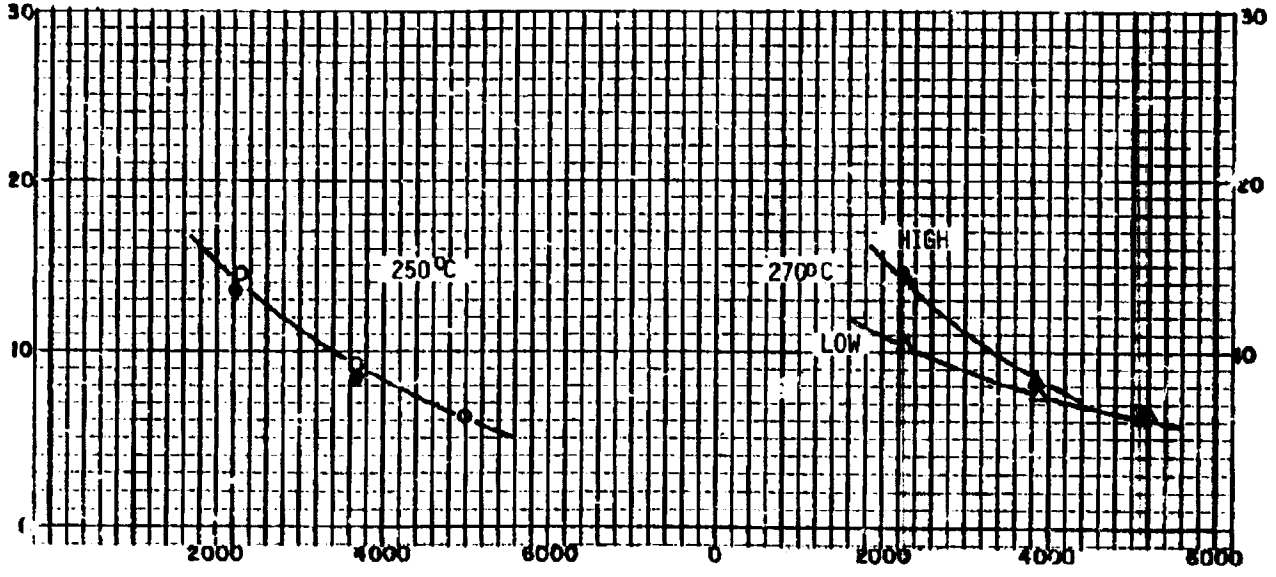
FIGURE 3-13

RECIRCULATING LIQUID REACTOR CONVERSION FOR LURGI GAS

COMMERCIAL CATALYST B / WITCO 40

35 ATM

T ; °C	LEGEND LIQUID RATE LOW	HIGH
250	○	●
270	△	▲



CO CONVERSION TO MeOH ; %

VHSV ; HR.$^{-1}$

FEED COMPOSITION

COMPONENT	MOLE %
CH_4	15
CO	25
CO_2	10
H_2	50

Section 4

BENCH SCALE UNIT DEVELOPMENT PROGRAM

DESCRIPTION OF THE BENCH SCALE UNIT

The bench scale unit was designed for the following operational conditions:

- Pressures to 1500 psig.
- Temperatures to 650°F.
- Gas space velocities ranging from 1000-10,000 SCFH/CF catalyst with actual gas feed rates of 30 SCFH.
- Liquid velocities of 20-100 gpm/ft^2.

Stainless steel (316) construction is used throughout. The unit is, to a great degree, automatically controlled. The unit is housed in a vented, explosion-proof cell with controls and analytical support in a nearby room. A simplified flow diagram of the bench scale unit is shown in Figure 4-1.

The feed gas is pumped from a cylinder bank (using a gas booster compressor (C-100) into the feed gas surge tank (D-103). The gas flow to the reactor is controlled by the feed gas control valve (FV-100). Before entering the reactor, the gas passes through a ZnO reactor (R-101) maintained at 700-800°F in order to remove trace sulfur compounds which would poison the catalyst. Iron carbonyl would also be removed at these temperatures. After mixing with the circulating liquid stream, the feed gas passes upwards in the methanol synthesis reactor (R-100) and then into the vapor/liquid separator (D-100). The oil level in the vapor/liquid separator is maintained by a liquid level controller (LIC-100) which controls the flow from the process oil makeup pump (P-100). The liquid, after leaving the vapor/liquid separator, returns to the process oil circulating pumps (P-101 A, B, C, D) and is then pumped back to the reactor, completing the oil circulation loop. A process oil filter (F-101) is located just after the pumps in order to remove any catalyst fines produced through particle attrition. The reacted product gases leave the vapor/liquid separator passing through

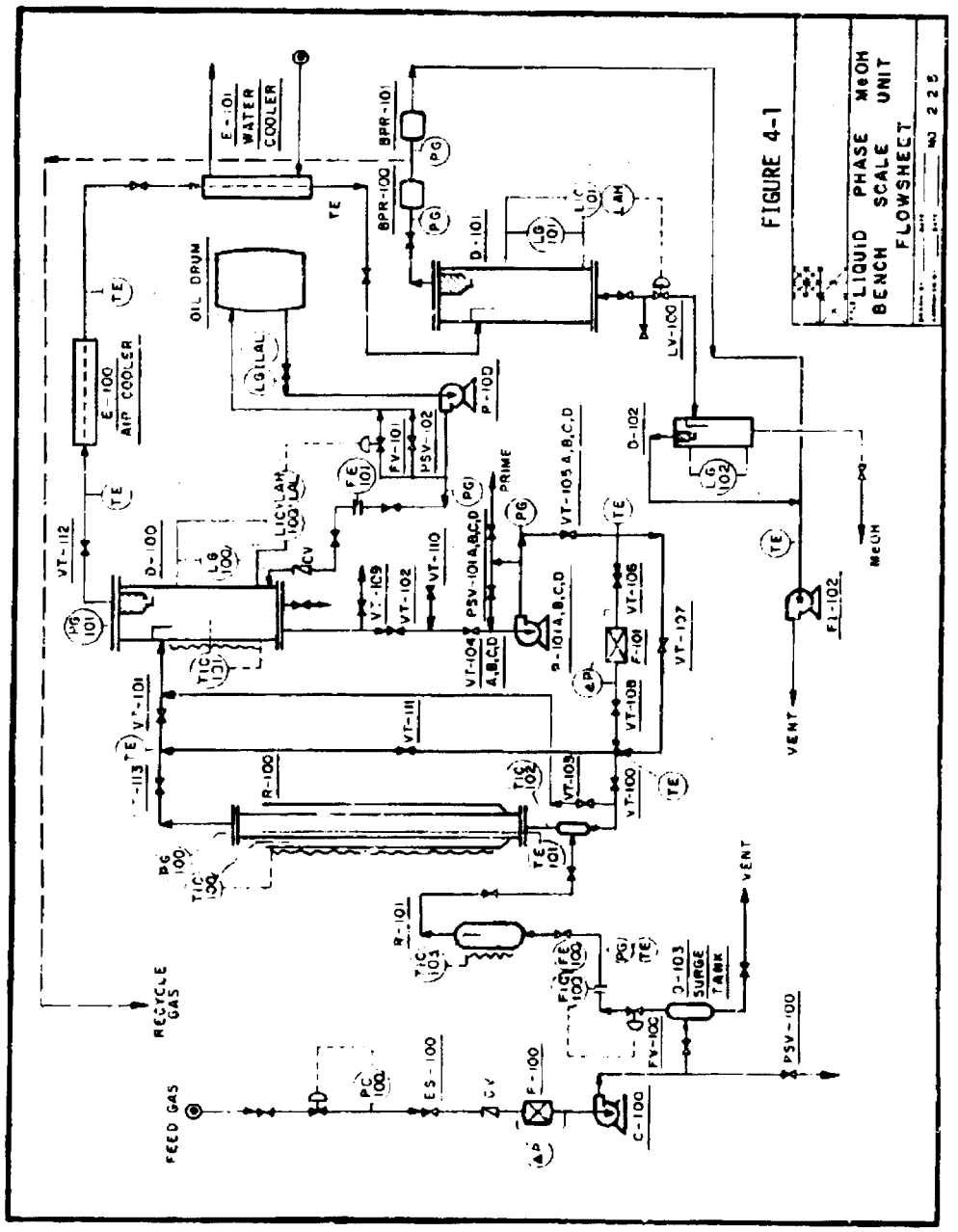


FIGURE 4-1
LIQUID PHASE MeOH
BENCH SCALE UNIT
FLOWSHEET
E-101 WATER COOLER
E-100 AIR COOLER
OIL DRUM
D-100
D-101
D-102
D-103 SURGE TANK
R-100
R-101
P-100
P-1014,B,C,D
C-100
F-100
F-101
F1-102
BPR-101
BPR-100
PSV-100
PSV-101
PSV-102
LV-100
FV-101
FV-100
ES-100
VT-100
VT-101
VT-102
VT-103
VT-104 A,B,C,D
VT-105 A,B,C,D
VT-106
VT-107
VT-108
VT-109
VT-110
VT-111
VT-112
VT-113
PRIME
RECYCLE GAS
FEED GAS
VENT
MeOH

both an air cooler (E-100) and water cooler (E-101) before entering the product vapor/liquid separator (D-101). The product gas stream is vented to the laboratory incinerator for safe combustion. The condensed liquids, containing both product methanol and circulating oil are then dumped into a product accumulator (D-102) through the liquid level control valve (LV-100). The condensed circulating oil is phase separated from the methanol and recycled back into the circulating loop. Residence time in the separator is approximately one hour.

INITIAL OPERATIONS

Having established the feasibility of the basic engineering concept in the early fixed bed test (described in Section 3), an initial series of reaction tests were conducted in the BSU with the following objectives in mind:

- Observe the normal operating behavior of all the mechanical components and make any changes necessary to improve the system behavior.
- Determine the system response time to reach steady-state after a change in operating conditions.
- Calibrate the flow measuring devices in conjunction with the analytical system to enable calculation of the system mass balances.
- Qualitatively examine the improvement in reaction rates for this fluidized bed reactor as compared to the previous fixed bed results.

The process variable scan was carried out with the Catalyst A (12-16 mesh)/Witco 40 mineral oil catalyst-liquid pair and a 75 percent H_2/25 percent CO feed gas. This feed gas composition was chosen for two reasons. Firstly, the straight H_2/CO feed would enable accurate determination of the small methane selectivity expected; and secondly, by using a $3H_2$/1CO feed gas, instead of a $2H_2$/1CO mixture, there would

be an additional check on CO conversion by following the changing product gas composition. It should be noted also that the feed gas was free of CO_2 while numerous literature references indicate CO_2 is desirable to both speed up the reaction rate and aid in maintaining catalyst activity.

The results of the experimental runs for the 3:1 hydrogen/carbon monoxide mixture are given in Table 4-1. The value for the CO conversion to methanol is based on a gas phase carbon balance around the system and is calculated with the following equation:

$$\text{Fractional CO Conversion} = \frac{F_{in}(X_{CO})_{in} - F_{out}\,(X_{CO}+X_{CO2})_{out}}{F_{in}\,(X_{CO})_{in}}$$

The early reaction runs all produced less methanol than was predicted from the gas phase material balance calculations. In trying to determine the reason for the discrepancy, it was hypothesized that a portion of the methanol produced was being condensed at the top of the V/L separator where the Grayloc hub provided a large area for heat losses. It would then reflux back into the separator where it was held up in the large volume and on the large surface area of the packing present in the upper half of the separator. To eliminate the possibility of this occurring, an additional heating tape was added to the top of the V/L separator and the Grayloc hub was further insulated. Also excess condenser capacity was eliminated to reduce the surface area available for liquid holdup. When operating temperature was achieved on the following day, a large excess of methanol continued to flash over into the product separator during most of the run, and even on the last collection of the day, 39 percent more methanol was collected than predicted by the gas phase material balance. For the following runs thereafter, the calculated conversions based on gas phase composition and product methanol collection were in close agreement. CO conversions determined by gas phase analysis are plotted against VHSV in Figure 4-2 and compared to the fixed bed case with recirculating liquid reported previously in

TABLE 4-1

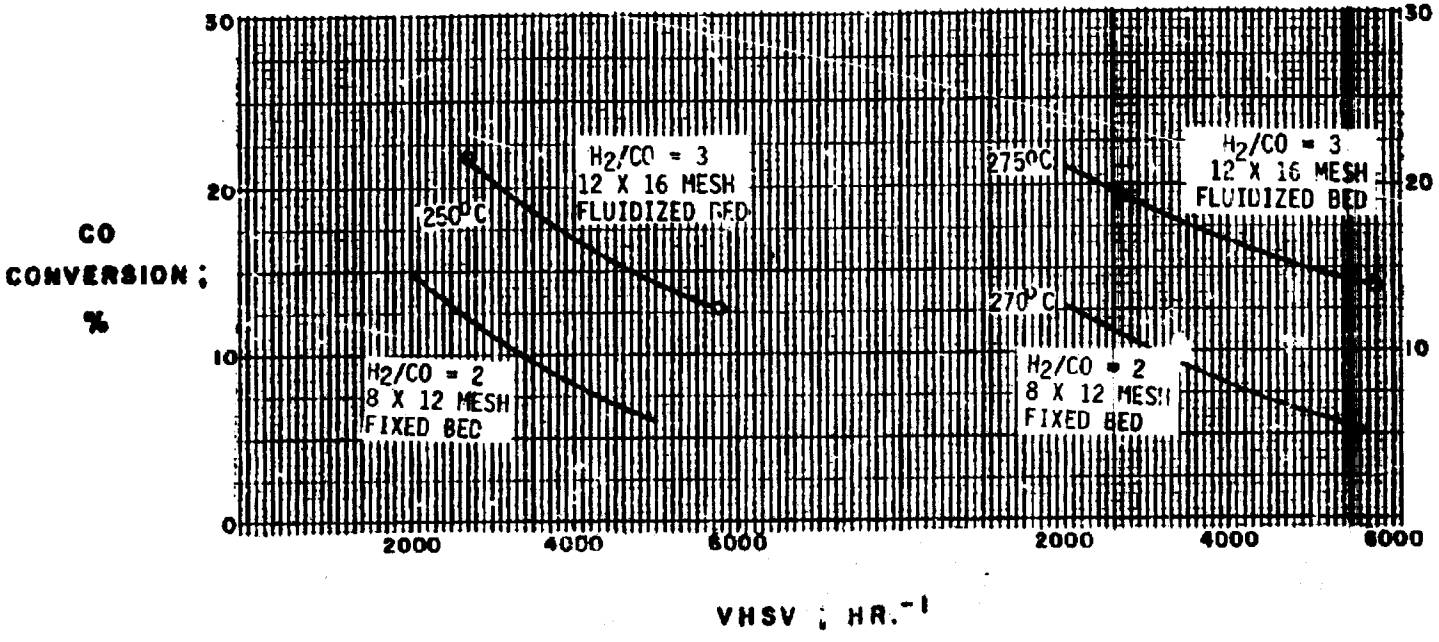
METHANOL SYNTHESIS IN THE BENCH SCALE FLUIDIZED BED
CATALYST A (12-16 MESH)*/WITCO 40 MINERAL OIL
FEED GAS: 75% H_2, 25% CO

Run No.	Temp.°C	Pressure, psig	VHSV, Hr^{-1}	CO Conv. to MeOH, %
1-A	250	505	5,775	13.2
1-B	250	505	2,675	21.7
2-A	225	505	5,800	7.1
2-B	225	505	2,625	13.5
3-A	275	505	5,775	14.2
3-B	275	505	2,650	19.4
4	250	1,015	2,650	46.0
5	250	1,010	5,825	21.4
6	225	1,015	2,760	27.8

* 200 cm^3 = 219.6 grams.

FIGURE 4-2

COMPARISON OF
FIXED AND FLUIDIZED BED REACTION RESULTS
35 ATM



Section 3. They indicate an improvement in the CO conversion which is substantially greater than what would be expected due to the differences in feed composition alone (75 percent H_2, 25 percent CO vs. 50 percent H_2, 25 percent CO and 25 percent inerts in the fixed bed case). The difference is especially significant at the higher VHSV's (3,500 Hr^{-1}) where more than twice the conversion was obtained at the same conditions.

The initial results for effect of pressure on CO conversion to MeOH are presented in Figure 4-3, indicating that a doubling of the pressure approximately doubled the CO conversion at a given VHSV.

In addition to these results, product gas and liquid compositions were continuously monitored in order to quantify the selectivity to by-products. In all cases, the selectivity to CH_4 was less than 0.5 percent and the selectivity to EtOH was less than 0.2 percent; DME was not detected in either the product gas or liquid samples.

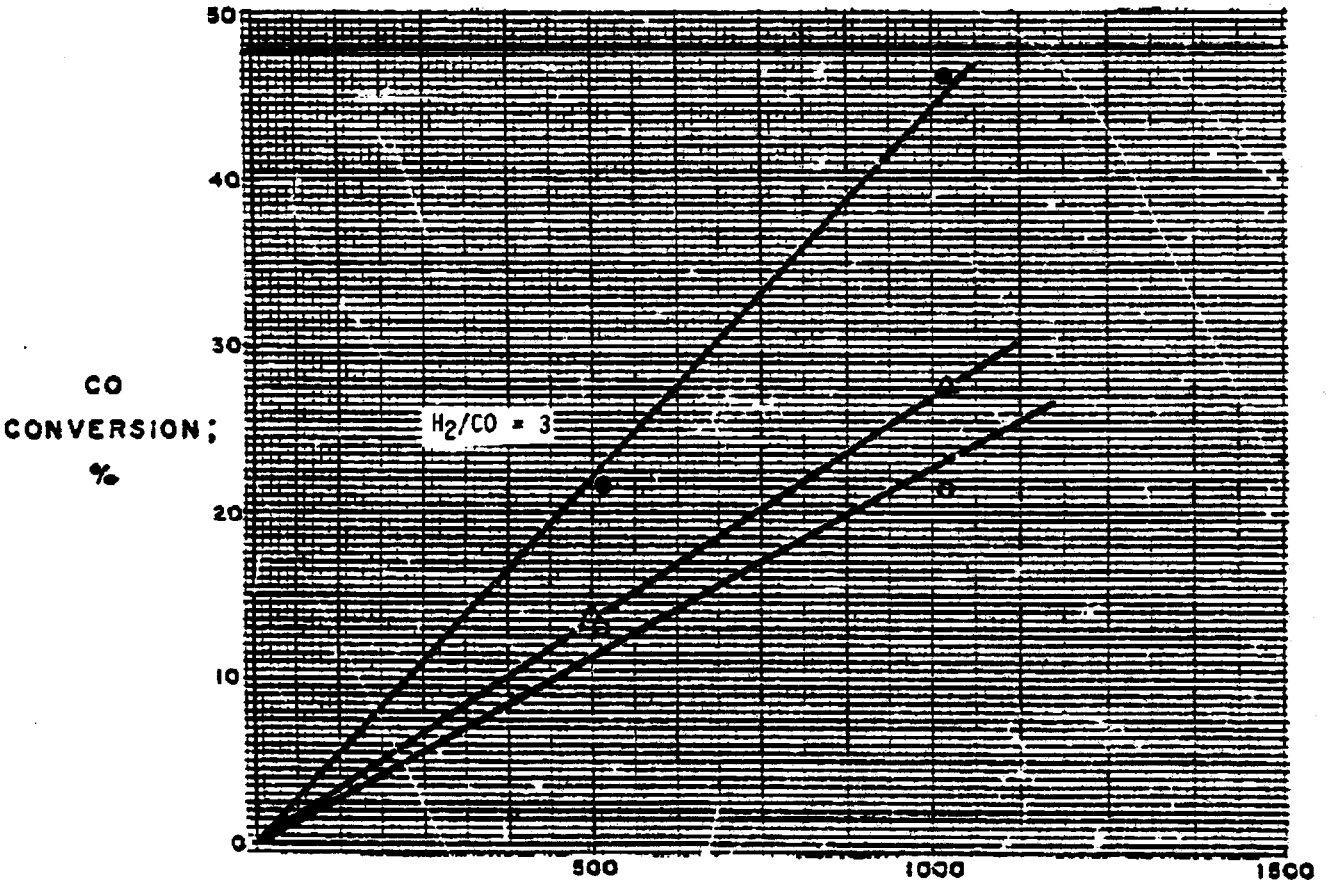
The calculated conversions based on the gas phase analysis and flow rates provide a better indication of reactor performance, as compared to conversion values determined from in-hand methanol collections. However, there is still a fair degree of uncertainty in the results at the lower conversion levels. This is because the computation involves taking the difference between two large numbers, each with a small uncertainty. For example, a 1 percent error in the effluent flow rate measurement would cause a 10 percent relative error in the calculated conversion at the 10 percent CO conversion level. In the future, where possible, feed gases will have an inert (methane or argon) in order to more precisely determine the exact ratio of feed/effluent gas rates and therefore minimize this uncertainty.

PROCESS VARIABLE SCANS - CRUSHED COMMERCIAL CATALYSTS

The main thrust of this portion of the experimental program was to determine the system response as a function of the following basic

FIGURE 4-3

EFFECT OF PRESSURE ON METHANOL CONVERSION



	T °C	VHSV
●	250	2650
○	250	5800
△	225	2700

operating parameters: (1) feed gas composition and flow rate, (2) reaction temperature, (3) reaction pressure, and (4) catalyst particle size. All of the data were obtained using the Commercial Catalyst A/Witco 40 Mineral Oil system and are presented in Tables 4-2, -3 and -4. Table 4-2 covers work with -12 +16 mesh catalyst particles, while Tables 4-3 and 4-4 cover work with -16 +20 mesh catalyst particles. Several different feed gases were investigated. Their nominal compositions are shown below.

Typical Feed Gas Analysis

	Lurgi	KT	$2H_2/1CO/0.1CO_2$	$2H_2/1CO$
H_2	50.0	38.0	64.7	66.7
CH_4	15.0	-	-	-
CO	25.0	60.0	32.3	33.3
CO_2	10.0	2.0	3.0	-

The reaction temperature was varied over the range of 200-275°C. The bulk of the runs were performed at a reaction pressure of 1000 psig with several comparisons made at the 500 and 1500 psig levels. Feed gas rates up to 5000 SCFH/CF-catalyst were examined.

Effect of Feed Gas Composition

While the data taken over any short time period are internally consistent, comparisons of results which are far removed in calendar time are complicated by the attrition and subsequent loss of catalyst during that time period. For both series of runs with the Commercial Catalyst A, catalyst losses to the overhead separator of approximately 30-40 percent were encountered over about a one-month period. (This loss was mainly due to the method of grinding the pellets to the desired size range. Improvement in this operation (e.g. ballmilling) substantially eliminated this problem). For the third loading (-16 +20 mesh) at least,

TABLE 4-2

COMMERCIAL CATALYST A (-12+16 MESH)[1]/WITCO 40 MINERAL OIL

Run No.	Feed Gas	Temp. °C	Pressure psig	VHSV Hr^{-1}	CO Conversion To MeOH[2] %
13-1	Lurgi	225	1,000	3,110	32.3
13-2	Lurgi	225	1,000	2,100	42.0
14-1	Lurgi	275	1,000	3,650	24.2
14-2	Lurgi	275	1,000	1,240	32.5
15-1	Lurgi	250	1,000	3,200	30.6
15-2	Lurgi	250	1,000	3,300	32.6
16-1	Lurgi	225	1,000	1,125	48.5
16-2	Lurgi	225	1,000	4,795	24.6
17-1	Lurgi	225	1,000	3,225	31.0
17-2	Lurgi	225	500	3,540	16.8
18-1	Lurgi	200	1,000	1,125	32.3
18-2	Lurgi	200	1,000	4,790	7.9
19-1	Lurgi	250	1,000	2,130	36.3
19-2	Lurgi	275	1,000	2,010	26.2
20-1	Lurgi	200	1,000	2,170	18.7
21-1	Lurgi	235	1,000	1,610	39.9
21-2	Lurgi	235	1,000	3,500	29.8
22-1	KT	235	1,000	3,510	11.7
23-1	KT	225	1,000	1,320	17.2
23-2	KT	225	1,000	2,430	10.4
24-1	KT	250	1,000	1,650	15.4
24-2[3]	KT	250	1,000	3,600	10.4
25-1	KT	250	1,000	2,430	12.5
25-2	KT	250	500	2,430	5-7[4]
26-1	KT	275	1,000	1,650	15.0
26-2	KT	275	1,000	3,600	10.0
27	No Analysis				
28-1	Lurgi	225	1,000	1,475	27.0
28-2	Lurgi	225	1,000	3,945	11.4
29-1	KT	250	1,000	2,430	12.5
29-2	KT	250	500	2,430	5-7[4]

(1) Catalyst loading (-12 +16 mesh) = 200 cm^3 = 219.6 gm.

(2) Includes small selectivity to higher alcohols.

(3) After this run, the reactor salt bath heaters were inadvertently shut off. Salt bath cooled to 30°C overnight.

(4) Inconsistent data analysis.

TABLE 4-3

COMMERCIAL CATALYST A (-16+20 MESH)(1)/WITCO 40 MINERAL OIL

Run No.	Feed Gas	Temp. °C	Pressure psig	VHSV Hr^{-1}	CO Conversion To MeOH(2), %
30-1	Lurgi	235	1,000(3)	3,870	56.6
31-1	Lurgi	235	1,000(3)	3,825	51.8
32-1	Lurgi	235	500(3)	2,070	34.9
32-2	Lurgi	235	500(3)	3,300	27.8
33-1(4)	Lurgi	235	500	2,870	24.1
34-1(4)	Lurgi	235	1,000	2,580	48.5

(1) Catalyst loading (-16 +20 mesh) ~ 200 cm^3 = 219.6 gm.

(2) Includes selectivity to higher alcohols.

(3) Excessive liquid flow rate caused a portion of the catalyst bed to rise and plug against the catalyst retaining screen. Pressure drop across packed bed portion was on the order of 200-300 psig.

(4) Reactor was cooled and depressurized in order to remove catalyst plug prior to these runs. In addition, liquid flow rate was lowered to prevent reoccurrence of plugging.

TABLE 4-4

COMMERCIAL CATALYST A (-16+20 MESH)[1]/WITCO 40 MINERAL OIL

Run No.	Feed Gas	Temp. °C	Pressure psig	VHSV Hr^{-1}	CO Conversion To MeOH[2], %
35-1	Lurgi	235	500	2,655	22.0
35-2	Lurgi	235	500	1,430	25.9
36-1	Lurgi	235	1,000	2,512	46.8
36-2	Lurgi	235	1,000	4,050	41.0
37-1	Lurgi	220	1,000	2,455	48.3
37-2	Lurgi	220	1,000	4,015	38.2
38-1	Lurgi	200	1,000	2,450	37.3
38-2	Lurgi	200	1,000	4,110	25.0
39-1	Lurgi	250	1,000	2,520	40.5
39-2	Lurgi	250	1,000	4,065	35.9
40-1	KT	250	1,000	2,020	19.5
41-1	KT	270	1,000	2,185	19.6
41-2	KT	270	1,000	3,550	15.2
42-1	KT	235	1,000	3,580	15.2
42-2	KT	235	1,000	3,115	21.3
43-1	KT	235	1,000	2,050	17.3
44-1	$2H_2/1CO/0.1CO_2$	235	1,000	2,800	44.6
45-1	$2H_2/1CO/0.1CO_2$	250	1,000	2,800	41.9
45-2	$2H_2/1CO/0.1CO_2$	250	1,000	4,550	36.0
46-1	$2H_2/1CO/0.1CO_2$	220	1,000	2,760	37.3
46-2	$2H_2/1CO/0.1CO_2$	220	1,000	4,590	34.2
47-1	$2H_2/1CO/0.1CO_2$	235	1,000	4,590	35.7
47-2	Lurgi	235	1,000	3,920	23.2
48-1	$2H_2/1CO$	250	1,000	2,880	35.8[3]
48-2	$2H_2/1CO$	250	1,000	4,740	22.6[3]
49-1	$2H_2/1CO$	220	1,000	2,880	12.5
49-2	$2H_2/1CO$	220	1,000	4,740	7.0
50-1	$2H_2/1CO$	235	1,000	2,880	22.8
50-2	$2H_2/1CO$	235	1,000	4,740	7.4

(1) Catalyst loading (-16 +20 mesh) = 200 cm^3 = 219.6 gm.
(2) Includes small selectivity to higher alcohols.
(3) Presence of CO_2 in system from previous runs resulted in comparatively high conversions on this day.