MAXIMIZATION OF CO CONVERSION TO LIQUIDS

Chemical Means

In an attempt to increase the per pass methanol productivity, several alternatives were conceived that could displace the chemical equilibrium of methanol synthesis by further reacting methanol to a non-equilibrium end product. This may be accomplished by several methods. The addition of bases to the catalyst or process liquids could promote the formation of higher alcohols. The addition of acids, on the other hand, could promote the formation of dimethyl ether and other volatile ethers. This work was carried out in the small Parr autoclave reactor. The small size of the unit (600 cm^3) allows for rapid equilibration during a variable scan. A schematic diagram is given in Figure 5-5. The reactor is equipped with a mechanical stirrer (1800 rpm), which results in excellent gas-liquid-solid mixing. Temperature control is maintained by an external heating jacket and an internal cooling coil. During the runs, air is used as the cooling medium. This is switched to water for rapid cool-down at the end of a run.

Feed gas is metered to the reactor through a high-pressure rotameter at rates of up to 5 SCFH. Make-up liquid can be added during the course of the reaction. Catalyst for the run was generally reduced in a nearby unit and then loaded under N_2 into the autoclave. At times, various other in situ reduction methods were tried. The effluent was cooled and separated from the product and vaporized process liquid before analysis. Liquid products were also analyzed.

Catalyst Modifications to Produce Dimethyl Ether. The initial tests (Series 1) were performed using the Catalyst A (-16 +20 mesh)(1)/Witco 40 mineral oil/Lurgi gas system in order to check the operability of the experimental unit and to establish a basis for comparative purposes. The results of this and subsequent tests are presented in Table 5-7.

(1) Due to constant agitation in the reactor, the particles were greatly reduced in size during the scan.

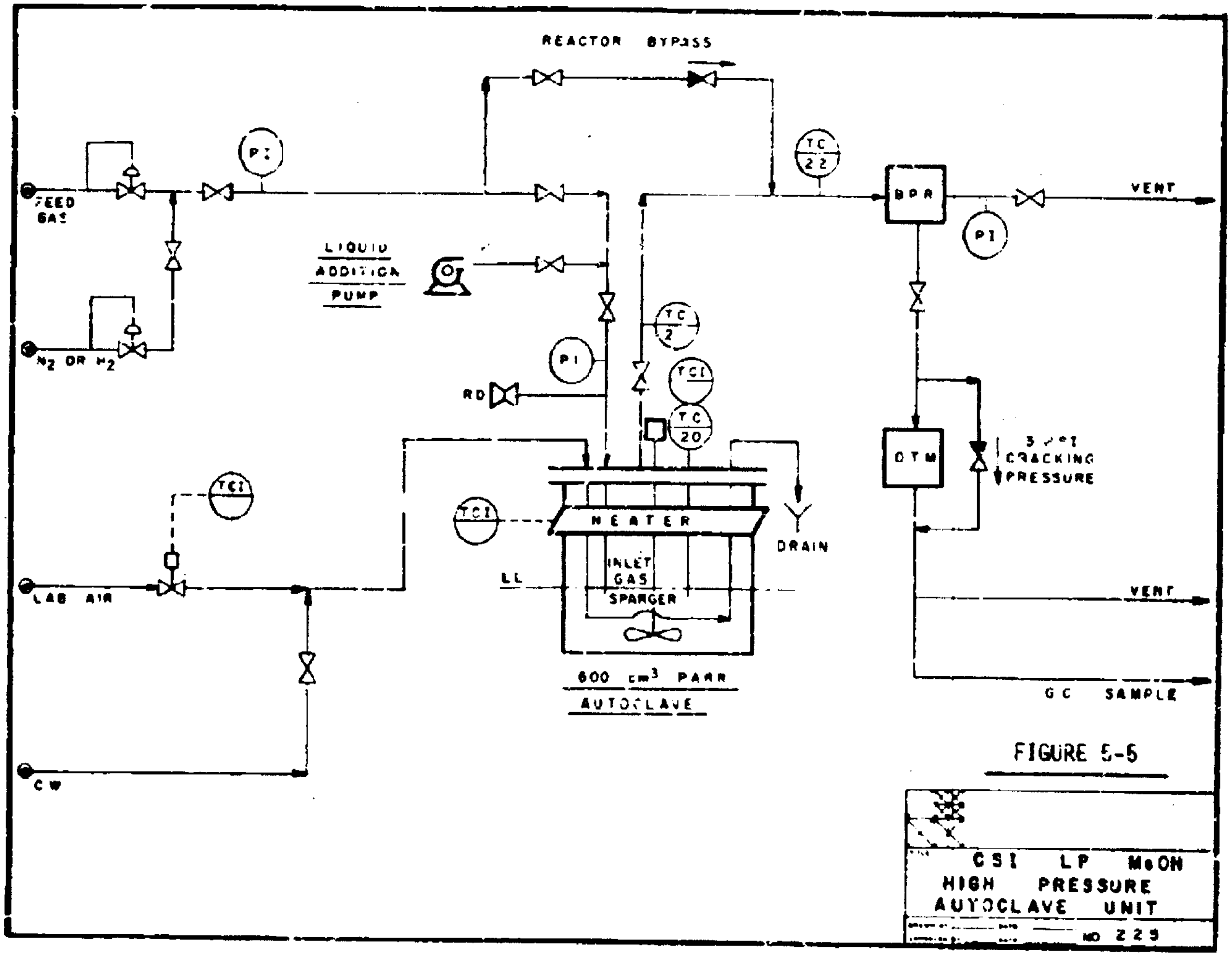

REACTOR BYPASS
FEED GAS
N2 OR H2
PI
LIQUID ADDITION PUMP
RD
PI
TC 2
TCI
TC 20
TC 22
BPR
PI
VENT
DTM
5 PSI CRACKING PRESSURE
DRAIN
HEATER
INLET GAS
SPARGER
LL
TCI
TCI
LAB AIR
CW
600 cm3 PARR AUTOCLAVE
VENT
GC SAMPLE
CSI LP MeOH
HIGH PRESSURE
AUTOCLAVE UNIT
NO 229

FIGURE 5-5

TABLE 5-7
MODIFIED CATALYSTS TO PRODUCE DIMETHYL ETHER
AUTOCLAVE TESTS: CATALYST A/WITCO 40 MINERAL OIL/LURGI FEED GAS(1)

RUN NO.	TEMP. (°C)	PRESS. (psig)	VHSV(6) (Liters Gas/ Kg Cat-Hr)	CO CONV. (%)	H_2 CONV. (%)	MeOH PRODUCTIVITY (Gm Mole MeOH Kg Cat-Hr x 10^3	COMMENT
Series I (2)							
114-4	230	500	2115	18.0	18.2	4.8	Base Case
114-6	250	500	2015	18.0	19.1	4.5	product
114-7	230	1000	3050	24.9	29.6	10.1	contains
114-8	250	1000	3210	26.0	26.8	11.0	trace DME.
Series II (3)							
114-11	250	1000	6435	23.5	23.5	19.5	Product
114-13	230	1000	6915	20.5	19.2	18.1	contains
114-14	280	1000	6600	21.5	19.6	17.8	trace DME.
Series III(4)							
114-16	250	1000	6200	33.7	27.5	28.6	Product
114-17	280	1000	3645	34.3	28.9	18.0	contains
114-18	300	1000	6820	26.4	20.8	24.0	trace DME.
Series VI(5)							
114-30	230	1000	2000	1	1	----	Product contains trace DME.

(1) Lurgi Feed Gas Nominal Composition: 50% H_2, 25% CO, 10% CO_2, and 15% CH_4.

(2) Catalyst: 80 Grams Commercial Catalyst A (-16 +20 mesh). Due to constant agitation in the reactor, the particles were greatly reduced in size during the scan.

(3) Catalyst: 40 Grams Commercial Catalyst A (powder) + 40 Grams γ-Alumina (Harshaw).

(4) Catalyst: 40 Grams Commercial Catalyst A (powder) + 40 Grams 13X Molecular Sieve.

(5) Catalyst: Prepared by impregnating Davison 980 silica-alumina tablets (1/4" Ø x 1/4") with a solution of the nitrate salts of copper, zinc, and chrome, such that the molar ratio of the Cu:Zn:Cr was 60:30:10. The saturated particles were dried overnight (140°C), calcined in a muffle furnace for 48 hours (265°C), and crushed (-30 +50 mesh) prior to reduction.

(6) Based on weight Catalyst A only for Series I, II and III.

The absolute methanol productivities obtained in Parr Autoclave are only 30-60 percent of those obtained in the Bench Scale Unit under similar reaction conditions. However, the purpose of the system was to easily screen a series of catalysts, and it proved to be entirely satisfactory.

The Series II and III runs were perf[illegible]sing a catalyst/co-catalyst pair, consisting of a mixture of Catalyst A and γ-alumina (Harshaw) and Catalyst A and molecular sieves (type 13x), respectively. It was expected that the α-alumina and/or the molecular sieves, due to their acidic nature, would catalyze the formation of ether from methanol;

$$2CH_3\,OH \longrightarrow CH_3OCH_3 + H_2O$$

and thereby react methanol to a non-equilibrium end product. However, as Table 5-7 indicates, while the productivities (based on Catalyst A weight only) are somewhat higher than the Series I tests, this is probably due to reduction procedure variances, since the liquid product and gas effluent streams contained only trace amounts of dimethyl ether.

Based on these initial results an impregnated version of the Series II catalyst/co-catalyst pair, was prepared. Unfortunately, this resulted in a virtually non-reactive system.

Due to the discouraging results, this work was terminated.

Catalyst Modifications to Produce Higher Alcohols. A literature review indicated several possible alternatives for increasing the yields of higher alcohols during methanol synthesis. In addition to particular catalyst compositions, potassium doping and physical catalyst/co-catalyst mixtures were suggested. Eight series of process variable scans were performed in the autoclave system to investigate these proposed alternatives. The results are summarized in Table 5-8.

The Series I and II tests evaluated catalyst compositions reported effective for the synthesis of higher alcohols. The process conditions used were those recommended in the source article. The Series I catalyst

TABLE 5-8

MODIFIED CATALYSTS TO PRODUCE HIGHER ALCOHOLS

AUTOCLAVE TESTS: CATALYST A/WITCO 40 MINERAL OIL/LURGI FEED GAS(1)

RUN NO.	TEMP. (°C)	PRESS. (psig)	WHSV (Liters Gas/ Kg Cat-Hr)	CO CONVERSION (%)	H_2 CONVERSION (%)	ALCOHOL PRODUCTIVITY (Gm Moles/ Kg Cat-Hr) x 10^3	OBSERVATIONS
Series I (2)							
114-20	200	1000	2500	---	---	---	Catalyst died after 60 minutes. No MeOH or EtOH detected in the product stream. CH_4 probably made initially.
114-21	200	1000	2500	---	---	---	
Series II (3)							
114-22	325	1000	3000	40-50	10-20	---	CH_4 and CO_2 made. MeOH and EtOH found in trace amounts.
114-25	325	1000	3000	50-60	15-25	---	
Series III(4)							
114-33	230	1000	3045	19.1	19.3	7.6	Reduction in-situ was successful, although lower activity than externally reduced catalyst

(1) Lurgi Feed Gas Nominal Composition: 50% H_2, 25% CO, 10% CO_2, and 15% CH_4.

(2) Catalyst: Cu-84.6, Zn-13.25, Cr-1.41, K-0.74; Memoir to the Institute of Chemistry, Academy of Sciences of the Ukranian SSR, IV,2, 1937.

(3) Catalyst: $(ZnO)_3$ $(CoO)_1$; Chemical Abstracts, 28, 5806, 4, 1934.

(4) Eighty grams commercial Catalyst A, reduced with a 10 H_2/90 N_2 gas stream at 15 psig while submerged and stirred in the Witco 40 mineral oil.

(Continued)

TABLE 5-8

MODIFIED CATALYSTS TO PRODUCE HIGHER ALCOHOLS

AUTOCLAVE TESTS: CATALYST A/WITCO 40 MINERAL OIL/LURGI FEED GAS(1)

RUN NO.	TEMP. (°C)	PRESS. (psig)	WHSV (Liters Gas/ Kg Cat-Hr)	CO CONVERSION (%)	H_2 CONVERSION (%)	ALCOHOL PRODUCTIVITY (Gm Moles/ Kg Cat-Hr) x 10^3	OBSERVATIONS
Series IV (5)							
114-34	230	1000	1045	10-14	10-15	1.2	Activity much lower than runs from Series III. Selectivity to EtOH is still less than 1%.
Series V (6)							
114-15	230	1000	3045	3-4	3-4	1.2	Activity much lower than runs from Series III. Selectivity to EtOH is still less than 1%.
Series VI (7)							
114-41 e	250	1000	6580	23.3	23.1	19.0	
114-41 g	270	1000	6540	25.5	22.0	19.8	External reduction.
114-41 h	290	1000	6450	27.3	20.0	20.0	Selectivity to higher alcohols less than 1%.

(5)Eighty grams commercial Catalyst A doped with 2 weight percent KOH, reduced with Lurgi feed gas at 1000 psig while submerged and stirred in the Witco 40 mineral oil.

(6)Eighty grams commercial Catalyst A doped with 2 weight percent $K_2 CO_3$, reduced with H_2 at 1000 psig while submerged and stirred in the Witco 40 mineral oil.

(7)Catalyst: 27.5 gm reduced Catalyst A + 24.7 gm reduced Girdler G-82 ammonia synthesis catalyst. WHSV based on Catalyst A weight only. Both catalysts reduced according to manufacturer's specifications.

(Continued)

TABLE 5-8

MODIFIED CATALYSTS TO PRODUCE HIGHER ALCOHOLS

AUTOCLAVE TESTS: CATALYST A/WITCO 40 MINERAL OIL/LURGI FEED GAS(1)

RUN NO.	TEMP. (°C)	PRESS. (psig)	VHSV (Liters Gas/ Kg Cat-Hr)	CO CONVERSION (%)	H_2 CONVERSION (%)	ALCOHOL PRODUCTIVITY (Gm Moles/ Kg Cat-Hr) x 10^3	OBSERVATIONS
Series VII(8)							
114-51	225	1000	2000	2	2	---	External reduction. Very low activity, but selectivity to higher alcohols is 5-10%.
Series VIII(9)							
114-69	240	1000	2000	3	3	---	External reduction. Very low activity, but selectivity to higher alcohols is 5-10%.

(8) Eighty grams commercial Catalyst A doped with 6 weight percent KOAc, reduced with 2 percent H_2/N_2 at atmospheric pressure in an external reduction unit with no oil present (according to manufacturer's specification).

(9) Repeat preparation of catalyst used in Series VII.

exhibited an exotherm when put on stream, but rapidly lost activity over the first 60 minutes. No alcohols were detected. The Series II tests resulted in a catalyst with activity toward methanation only, producing CH_4 and CO_2 with only trace amounts of alcohols detected.

The Series IV, V, VI and VII tests involved some form of potassium doping on a Catalyst A base material. Just prior to these series of runs, a base case run was conducted with a modified Catalyst A (Series III) to attempt the catalyst reduction in the presence of the liquid phase. It was felt at the time that an in situ reduction procedure would reduce the turn-around time between the test series. By and large, all the in situ reduction procedures (Series III, IV, and V) were only partially successful. While the catalysts did display various degrees of activity, none of the tests resulted in activities as high as the externally reduced catalysts. Nevertheless, based on the liquid product compositions ($<1\%$ higher alcohols) for the KOH (Series IV) and K_2CO_3 (Series V) tests, it was concluded that these catalysts are not suitable for producing significant quantities of higher alcohols, relative to methanol. Because of the uncertainty of the in situ reduction procedure, the KOAc doped catalyst was reduced externally prior to testing. As the Series VII test results show, this version of the catalyst had very little activity. It did, though, produce a product containing 5-10 weight percent higher alcohols. The test was repeated with a new batch of KOAc doped catalyst (Series VIII). The results for this second scan were much the same (moderately increased selectivity to higher alcohols with a greatly reduced reactivity) and is apparently a direct result of the KOAc doping.

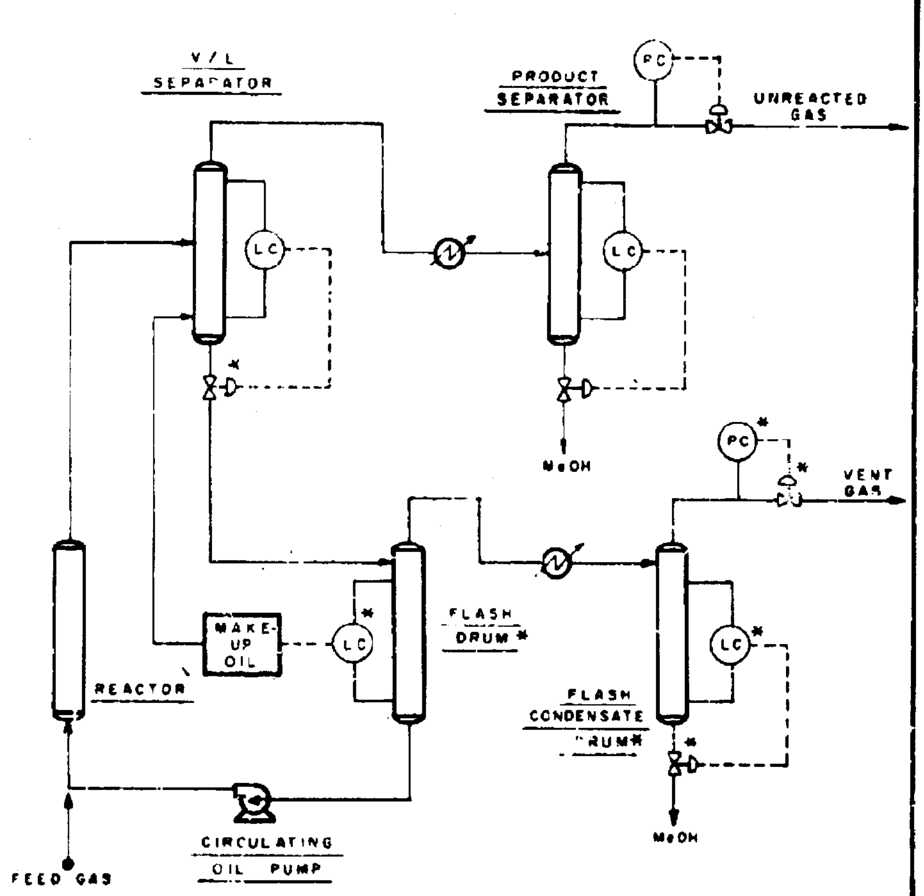
The third form of catalyst modification tested was a simple mixture of Catalyst A with a commercial iron based ammonia synthesis catalyst (Girdler G-82). This co-catalyst was chosen because it is known to be effective in Fischer-Tropsch synthesis reactions for the production of higher hydrocarbons and oxygenated compounds. Each portion of the catalyst/co-catalyst pair was reduced externally prior to running in the autoclave. The results for these tests (Series VI), as outlined in Table 5-8, indicate that the reactivity results chiefly from the Catalyst A present. There was no noticeable increase in higher alcohols due to the presence of the co-catalyst.

Physical Means

Several methods were also considered for the removal of dissolved methanol from the circulating oil phase prior to its recycle to the reactor. In this manner, the circulating liquid phase could behave as a "physical sink" for the product methanol and, therefore, result in increased CO conversions by recovering methanol product not only from the reactor effluent vapor stream but also from the circulating process liquid.

In order to experimentally test this concept, the BSU was modified as shown in Figure 5-6. The circulating oil, as it leaves the vapor/liquid separator, passes through a pressure letdown valve, nto a flash drum, which is maintained at an intermediate pressure. Dissolved methanol and dissolved synthesis gas are flashed off, cooled, condensed and sent to the flash condensate drum. The process liquid from the flash drum is returned to the reactor by the circulating oil pump.

Two preliminary process scans were performed in which the combined liquid/gas reactor effluent was flashed, instead of the liquid stream only. This technique demonstrated the technical feasibility of this concept and at the same time tested the operation of the unit in this flash mode. Both series of experiments used Catalyst A (-16 +20 mesh) and a Lurgi feed gas. The first experiments were performed with Witco 40 mineral oil, while the second series used Exxon Aromatic 150 as the liquid phase. Throughout the flash vaporization runs, the catalyst activity level was continuously monitored, and found to remain fairly stable. The results of these variable scans are presented in Tables 5-9 and 10 and Figure 5-7. The results substantiate the principle of increasing the methanol productivity by allowing the circulating liquid phase to behave as a methanol "sink". It also appears that the increase in productivity is more a function of the ratio of the reactor and flash vessel pressures, rather than the absolute pressure difference. There was no clear-cut distinction as to which process liquid is more sensitive to this process variation, although from solubility considerations



* MODIFICATIONS

FIGURE 5-6

BSU MODIFICATIONS FOR
FLASH VAPORIZATION
SCHEME

NO 226

TABLE 5-[illegible]

FLASH VAPORIZATION OF REACTOR EFFLUENT COMMERCIAL CATALYST A[1] (-16 +20 MESH)/ WITCO 40 MINERAL OIL/LURGI FEED GAS[2]

RUN NO.	TEMP. (°C)	PRESSURE IN REACTOR (psig)	PRESSURE IN FLASH VESSEL (psig)	VHSV Hr⁻¹	CO CONVERSION(3) (%)
95-1	230	500	---	1150	28.9
95-2	230	500	100	1215	44.8
96-1	230	1000	---	1750	52.5
97-1	230	1000	700	1745	54.6
97-2	230	1000	400	1700	55.2
97-3	230	1000	100	1770	68.1
98-1	230	1000	---	4625	41.0
98-2	230	1000	---	1760	49.4
99-1	230	1000	700	4600	42.2
99-2	230	1000	400	4635	46.4
99-3	230	1000	100	4525	50.6
100-1	215	1000	---	4825	36.6
100-2	245	1000	---	4800	37.2
101-1	230	1000	---	4530	39.2
101-2	230	500	---	2305	23.5
102-1	215	500	---	2255	24.3
102-2	245	500	---	2310	19.6

[1] Catalyst Loading = 213.6 grams.

[2] Nominal Composition: 50% H_2, 25% CO, 10% CO_2, and 15% CH_4.

[3] Includes small selectivity to higher alcohols.

TABLE 5-10

FLASH VAPORIZATION OF REACTOR EFFLUENT
COMMERCIAL CATALYST A[1] (-16 + 20 MESH)/
EXXON AROMATIC 150/LURGI FEED GAS [2]

RUN NO.	TEMP. (°C)	PRESSURE IN REACTOR (psig)	PRESSURE IN FLASH VESSEL (psig)	VHSV Hr^{-1}	CO CONVERSION[3] (%)
103-1	230	500	---	2520	18.9
104-1	230	500	---	2570	17.2
104-2	230	500	150	2710	22.9
104-3	230	1000	---	4610	41.3
105-1	230	1000	700	4595	44.1
105-2	230	1000	400	4420	50.3
105-3	230	1000	150	4310	52.8
106-1	230	1000	---	1345	55.9
106-2	230	1000	700	1450	53.7
106-3	230	1000	400	1320	56.1
107-1	230	1000	---	4625	39.1
108-1	230	1000	---	4750	39.6
108-2	245	1000	---	4765	36.4
108-3	215	1000	---	4980	40.1
109-1	230	1000	---	1590	51.9
109-2	230	1000	100	1630	61.9
109-3	230	1000	100	4830	52.5
110-1	230		---	4535	40.9

---

[1] Catalyst Loading = 219.5 grams.

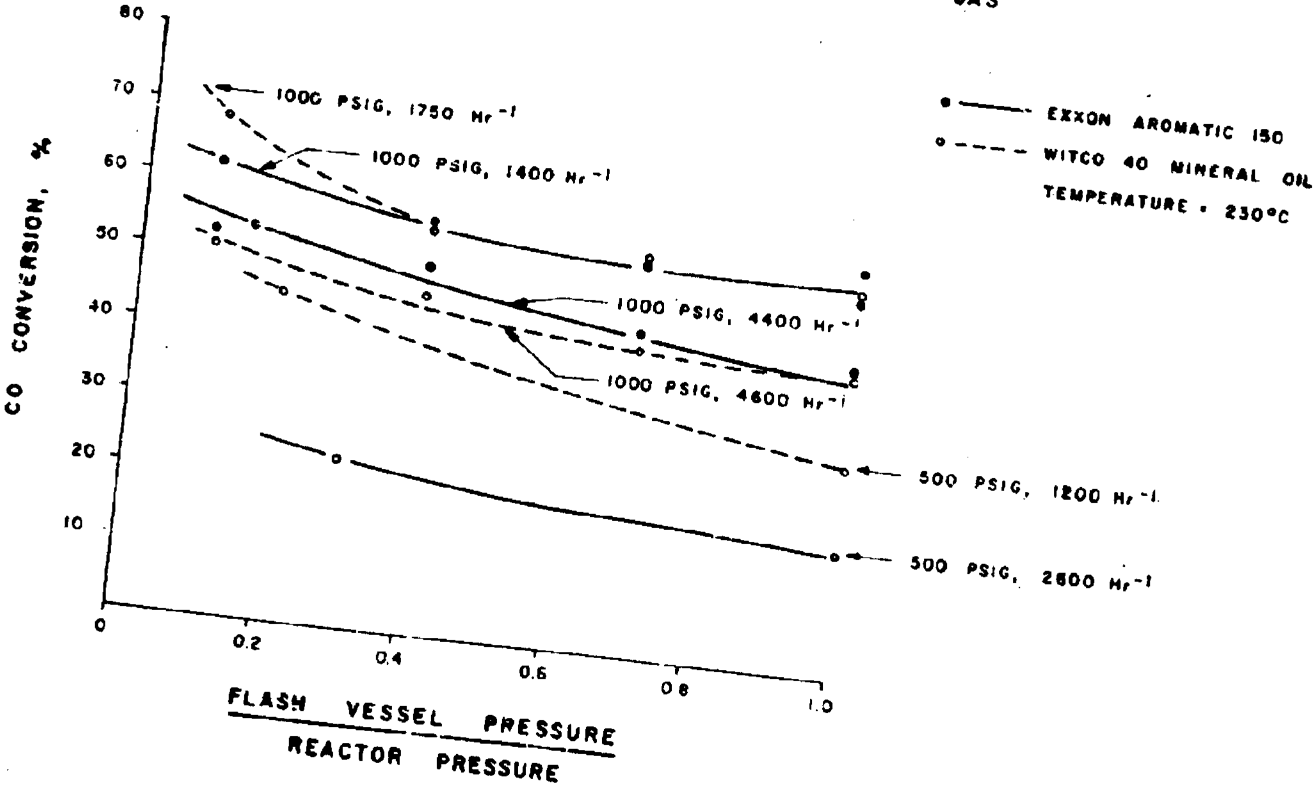
[2] Nominal Composition: 50% H_2, 25% CO, 10% CO_2, and 15% CH_4.

[3] Includes small selectivity to higher alcohols.

FIGURE 5-7

EFFECT OF REMOVAL OF METHANOL FROM THE CIRCULATING PROCESS LIQUID BY FLASH VAPORIZATION

COMMERCIAL CATALYST A / LURGI FEED GAS



(higher MeOH solubility in Exxon Aromatic 150), one would predict that an aromatic process liquid should show greater methanol productivity with this technique.

Initial tests with the fully modified reactor system were made with the UCI C-79-3 (-16 +20 mesh)/Witco 40/Lurgi gas system to help determine the best mode of operation (see Table 5-11). Runs 137-1, 138-1, and 139-1 established the base level activity for this catalyst/liquid pair. (The productivity is approximately one-third that of the Catalyst A system). For all subsequent runs, the circulating oil system was flashed to an intermediate pressure level.

The dissolved methanol, which flashed off was removed prior to re-circulating the oil back to the reactor. In contrast to the first series of runs where the liquid and vapor were blended together, the CO conversion seemed to be less sensitive to the flash level pressure. It was noted that there was no consistent pattern as to the distribution of the unreacted process gas between the vapor-liquid separator and the flash drum. Due to the low molar gas to liquid ratio in the reactor, it was not possible to maintain a consistent pressure balance throughout the reactor system. For this reason, this process scan was discontinued, and the more active Catalyst A was prepared for a similar variable scan (see Table 5-12A). Unfortunately, while the first two base level tests, Runs 148-1 and 148-2, indicated a normal level of activity, the subsequent flash runs showed large daily losses in activity. This test was shortly followed by a repeated loading of Catalyst A (see Table V-12B), which this time indicated a total loss in activity by the second day. While the exact nature of the problem was unknown, some type of contamination was indicated. Subsequent testing of the liquid and catalyst samples for known poisons (S, Cl) was not enlightening.

After the repeated failures in trying to perform the flash vaporization studies, the entire bench scale unit was completely and repeatedly flushed with solvents and fresh mineral oil prior to initiating a new series of experiments. The catalyst charge (Catalyst A) was reduced in the standard manner with particular care given to ending the H_2 treatment as soon as the reduction was completed. The unit was put back

TABLE 5-11

EFFECT OF MeOH REMOVAL BY FLASH VAPORIZATION

CCI-C-79(1) (-16 +20 MESH)/WITCO 40 MINERAL OIL/LURGI FEED GAS(2)

RUN NO.	TEMP. (°C)	REACTOR PRESSURE (psig)	FLASH DRUM PRESSURE (psig)	VHSV SCFH GAS/ CF CATALYST	CO CONVERSION TO MeOH(3) (%)
137-1	230	1000	---	2100	39.1
138-1	230	1000	---	1550	39.9
139-1	230	1000	---	2015	35.8
139-2	230	1000	500	2050	36.5
140-1	230	1000	500	1800	42.1
141-1	230	1000	500	1280	45.0
142-1	230	1000	500	1060	42.3
143-1	230	1000	700	1230	37.7
143-2	230	1000	500	1290	47.7
144-1	230	1000	300	1230	45.0
144-2	230	1000	700	1135	43.5
145-1	230	1000	700	1285	40.5

[1] Catalyst Loading = 219.6 grams.

[2] Lurgi Feed Gas = 50% H_2, 25% CO, 10% CO_2, and 15% CH_4.

[3] Includes small selectivity to higher alcohols.

TABLE 5-12A

EFFECT OF MeOH REMOVAL BY FLASH VAPORIZATION
CATALYST A(1) (-16 +20 MESH)WITCO 40 MINERAL OIL/LURGI FEED GAS(2)

RUN NO.	TEMP. °C	REACTOR PRESSURE (psig)	FLASH DRUM PRESSURE (psig)	VHSV SCFH GAS/ CF CATALYST	CO CONVERSION TO MeOH(3) (%)
148-1	230	1000	---	3225	46.3
148-2	230	1000	---	4825	39.6
149-1	230	1000	700	3325	28.2
150-1	230	1000	400	3070	18.2
151-1	230	1000	---	3375	8.5

TABLE 5-12-B

CATALYST A(1)/WITCO 40 MINERAL OIL

RUN NO.	TEMP. (°C)	REACTOR PRESSURE (psig)	FLASH DRUM PRESSURE (psig)	VHSV SCFH GAS/ CF CATALYST	CO CONVERSION TO MeOH(3) (%)
152-1	230	1000	---	3315	21.2
153-1	230	1000	---	3315	5.9
154-1	230	1000	---	3315	1.8

(1) Catalyst Loading = 219.6 grams.

(2) Lurgi Feed Gas = 50% H_2, 25% CO, 10% CO_2, and 15% CH_4.

(3) Includes small selectivity to higher alcohols.

on-stream in the normal mode of operation, in order to test the base activity level. Table 5-13 shows that the the conversion level obtained for Run 155-1, 45.0%, was typical of the normal system response. These run conditions were repeated on the following day with substantially the same results (Run 156-1). On the third day, the unit was put into the flash vaporization mode. Allowing for the modest improvement which should result due to flashing the oil to this intermediate pressure, the slightly lower conversion for Run 157-1, 42.6%, was probably more a result of the higher gas feed rate for this run, rather than any loss in activity. Unfortunately, when the unit was re-started after a three-day standby, the base case conversion had dropped to 32.4%. Nevertheless, the run was continued for one more day, and the effect of the lower flash pressure levels is dramatically shown by comparing Runs 158-1, 159-1, and 159-2. At this point, a decision was made to load a new charge of catalyst and perform the variable scans on a continuous, round-the-clock basis. The results for these series of runs are presented in Table 5-14 and Figure 5-8. Several of the data points were repeated during the scan in order to check the base activity level. From these results, it appears that the catalyst performance equilibrated after about 10-15 hours on stream.* All of the data are internally consistent, and the effect of flash pressure level compares quite favorably with previous results when the combined liquid and gas reactor effluent stream was flashed to the same intermediate pressure levels. The data at 250°C and the lower VHSV (2,200 Hr^{-1}) indicate a stronger dependence on flash pressure level than the 230°C data. In fact, for Run 160-6, the measured conversion is several points greater than the vapor phase equilibrium calculated for these reaction conditions. If this result is typical, the ability to operate the reactor system at a higher temperature may allow for the more economical generation of higher pressure steam. An economic evaluation of this scheme showed that for a large-sized complex where both methanol and synthesis gas are produced the flash vaporization mode offers savings of approximately $0.10 per MM Btu.

* After Run 160-4.

TABLE 5-13

EFFECT OF MeOH REMOVAL BY FLASH VAPORIZATION

CATALYST A(1)/WITCO 40 MINERAL OIL/LURGI FEED GAS(2)

RUN NO.	TEMP. (°C)	REACTOR PRESSURE (psig)	FLASH DRUM PRESSURE (psig)	VHSV SCFH GAS/ OF CATALYST	CO CONVERSION TO MeOH(3)
155-1	230	1000	---	3030	45.0
156-1	230	1000	---	3160	45.2
157-1	230	1000	500	3570	42.6
158-1	230	1000	---	3125	32.4
159-1	230	1000	100	3610	55.0
159-2	230	1000	700	3145	37.2

(1) Catalyst Loading = 219.6 grams

(2) Lurgi Feed Gas - 50% H_2, 25% CO, 10% CO_2 and 15% CH_4

(3) Includes small selectivity to higher alcohols.

TABLE 5-14

EFFECT OF MeOH REMOVAL BY FLASH VAPORIZATION
CATALYST A(1)/WITCO 40 MINERAL OIL/LURGI FEED GAS(2)

RUN NO.	TEMP. (°C)	REACTOR PRESSURE (psig)	FLASH DRUM PRESSURE (psig)	VHSV SCFH GAS/ CF CATALYST	CO CONVERSION TO MeOH(3) (%)
160-1	230	1000	---	2180	52.1
160-2	230	1000	---	4240	45.9
160-3	250	1000	---	2240	48.9
160-4	230	1000	700	2330	51.7
160-5	230	1000	400	2255	55.6
160-6	250	1000	400	2380	54.4
160-9(4)	250	1000	700	2255	43.4
160-10	230	1000	700	4210	39.1
160-11	230	1000	400	4260	41.2
160-12	230	1000	100	4360	48.2
160-13	250	1000	100	4390	49.8
160-14	250	1000	400	4470	40.1
160-15	250	1000	700	4425	37.2
160-16(5)	250	1000	700	2215	41.2
160-17	250	1000	---	4165	34.6
160-18	250	1000	---	2285	37.9
160-19(6)	230	1000	---	4295	38.1
160-20(7)	230	1000	---	2205	44.6
160-21	230	1000	---	1266	51.2

(1) Catalyst loading = 219.6 grams.

(2) Lurgi feed gas = 50% H_2, 25% CO, 10% CO_2, and 15% CH_4.

(3) Includes small selectivity to higher alcohols.

(4) Runs 160-7 and 160-8 had insufficient feed gas flow to maintain reactor pressure at 1000 psig while flash system operated at 100 psig.

(5) Repeat of Run 160-9.

(6) Repeat of Run 160-2.

(7) Repeat of Run 160-1.

FIGURE 5-8

EFFECT OF FLASH VAPORIZATION

PRESSURE : CATALYST A / WITCO 40

LURGI GAS AT P_R = 1000 PSIG