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Catalytic Gasification of Bagasse for the Production of Methanol

E. G. Baker
M. D. Brown
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October 1985

Prepared for the U.S. Department of Energy
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Pacific Northwest Laboratory
Operated for the U.S. Department of Energy
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CATALYTIC GASIFICATION OF BAGASSE FOR THE PRODUCTION OF METHANOL

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Pacific Northwest Laboratory
Richland, Washington 99352

SUMMARY

The Pacific Northwest Laboratory, operated for the U.S. Department of Energy by Battelle Memorial Institute has tested the gasification of bagasse in the presence of catalysts in a study co-sponsored by the Sugar Research Institute and the U.S. Department of Energy. The purpose of the study was to evaluate the technical and economic feasibility of catalytic gasification of bagasse to produce methanol.

In previous studies for the U.S. Department of Energy, Pacific Northwest Laboratory (PNL) developed a catalytic steam gasification process which converted wood to methanol synthesis gas in one step using nickel based catalysts in a fluid-bed gasifier. Tests in a nominal 1 ton/day process development unit (PDU) gasifier with these same catalysts showed bagasse to be a good feedstock for fluid-bed gasifiers, but the catalysts deactivated quite rapidly in the presence of bagasse.

The exact mechanism of deactivation of the nickel catalysts is not known, but laboratory studies indicate it is probably a function of both sulfur poisoning and carbon deposition. The sulfur level in bagasse (200-400 ppm) is about five times higher than the sulfur level in wood. Laboratory catalyst screening tests showed K_2CO_3 doped on the bagasse to be a promising catalyst for converting bagasse to methanol synthesis gas.

PDU tests with 10 wt% K_2CO_3 doped on bagasse showed the technical feasibility of this type of catalyst on a larger scale. A high quality synthesis gas was produced and carbon conversion to gas was high. The gasifier was successfully operated without forming agglomerates of catalyst, ash, and char in the gasifier. There was no loss of activity throughout the runs because catalyst is continually added with the bagasse. Laboratory tests showed about 80% of the potassium carbonate could be recovered and recycled with a simple water wash.

An economic evaluation of the process for converting bagasse to methanol showed the required selling price of methanol to be significantly higher than the current market price of methanol. A previous economic evaluation of the process using wood as a feedstock showed the required selling price of methanol

to be competitive with the market price of methanol (Mudge 1981). Several factors make this current evaluation using bagasse as a feedstock less favorable: 1) capital costs are higher due to inflation and some extra costs required to use bagasse, 2) smaller plant sizes were considered so economies of scale are lost, and 3) the market price of methanol in the U.S. has fallen 44% in the last six months. Production of methanol from bagasse may still be profitable in the long term if a shortage of transportation fuels occurs and prices increase significantly.

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INTRODUCTION

Bagasse produced as an unavoidable by-product of cane sugar represents a significant energy resource. It is primarily used as a fuel; however, more efficient use of bagasse or conversion of it to higher value products is being considered.

One possibility is conversion of bagasse to methanol. Methanol can be produced from biomass materials such as bagasse via relatively simple gasification technology. Methanol is a valuable chemical in the international market used primarily as a feedstock for production of other chemicals, particularly formaldehyde. However, the primary new interest in methanol is its potential as a transportation fuel. It is already being used as a gasoline additive and neat in fleet vehicles and is being considered as a fuel for gasoline engines, diesel engines and gas turbines.

Since 1977 Pacific Northwest Laboratory (PNL) has been developing a process for the catalytic gasification of wood under sponsorship of the U.S. Department of Energy. The process uses steam, indirect heat, and a catalyst to produce methanol synthesis gas in a fluidized bed gasifier. This report details the results of a project, sponsored jointly by the Australian Sugar Research Institute and the U.S. Department of Energy to test catalytic gasification of bagasse for the production of methanol. The project included gasification tests in laboratory scale gasifiers and a nominal 1 ton/day process development unit and an economic evaluation of the process.

LABORATORY STUDIES

Laboratory gasification tests were initiated after the first Process Development Unit (PDU) tests with supported nickel catalysts showed rapid catalyst deactivation. Two different types of catalyst systems were evaluated in the laboratory: 1) supported metal catalysts, primarily nickel based, and 2) alkali metal catalysts which were impregnated on the bagasse feed materials.

BAGASSE ANALYSIS

Bagasse used for the gasification tests was supplied by the Sugar Research Institute and was shipped from Australia. The bagasse was supplied in three different forms: loose, baled, and pelletized. The three different materials are described in more detail in the section on PDU tests. For laboratory gasification tests the pellets were ground to smaller than 1.4 mm (14 mesh).

Table 1 shows the analysis of the Australian bagasse pellets and some Hawaiian bagasse pellets used for comparison in some laboratory gasification tests. The ash and sulfur content of bagasse are significantly higher than wood. Otherwise the proximate and ultimate analyses are quite similar to hardwoods (Mudge 1983). Bagasse is primarily cellulose (38%), hemicellulose (38%) and lignin (20%). This also closely resembles the composition of hardwoods (Shafizadeh 1982).

DESCRIPTION OF LABORATORY GASIFIER

Two continuous-feed, fixed-catalyst-bed reactors constructed of quartz glass were used for these studies. The reactors, which are operated at atmospheric pressure are shown in Figure 1. Bagasse is fed into the top of the reactors at a rate of 8-12g/hr and gasified in the presence of steam. The volatile material passes over a bed of catalyst in the bottom of the reactor. The space velocity over the catalyst is typically 1000-2000 cm^3/g catalyst/hr. Char and ash accumulate on a support just above the catalyst bed and are vacuumed out periodically.

Analytical equipment used with the laboratory gasification system includes a Carle AGC-S analytical gas chromatograph for gas analysis, a Perkin Elmer 240

TABLE 1. Analysis of Bagasse Used for Gasification Tests

Source of Bagasse	Australia	Hawaii
Composition of Raw Bagasse (wt%)	--	--
Sugar		2
Fibre		43
Ash		7
Moisture		48
Proximate Analysis (wt% dry basis)	--	--
Volatile Matter		65
Fixed Carbon		22
Ash		13
Ultimate Analysis (wt% dry basis)	--	--
C	44.7	43.0
H	5.9	5.0
O	44.4	37.6
N	0.2	0.3
S ppm	300	--
Ash	4.8	13.3
Heating Value (dry basis)	--	--
kJ/kg	18,300	18,100
Btu/lb	7,870	7,780
Moisture Content (wt basis)	9.9	8.3
Form	pellets	pellets

elemental analyzer for wood and char analysis and a Dohrmann DC80 carbon analyzer for water analysis. Other analytical equipment was used for special purposes and will be discussed where appropriate.

The laboratory reactors are designed primarily for catalyst screening studies. The space velocity over the catalyst is similar to the space velocity in the PDU and past experience with wood has shown this system provides a good indication of catalyst performance in the PDU. Because of the differences in gas/solid contacting and residence time between the laboratory gasifiers and the PDU, the rate of gasification and overall carbon conversion can not be compared.

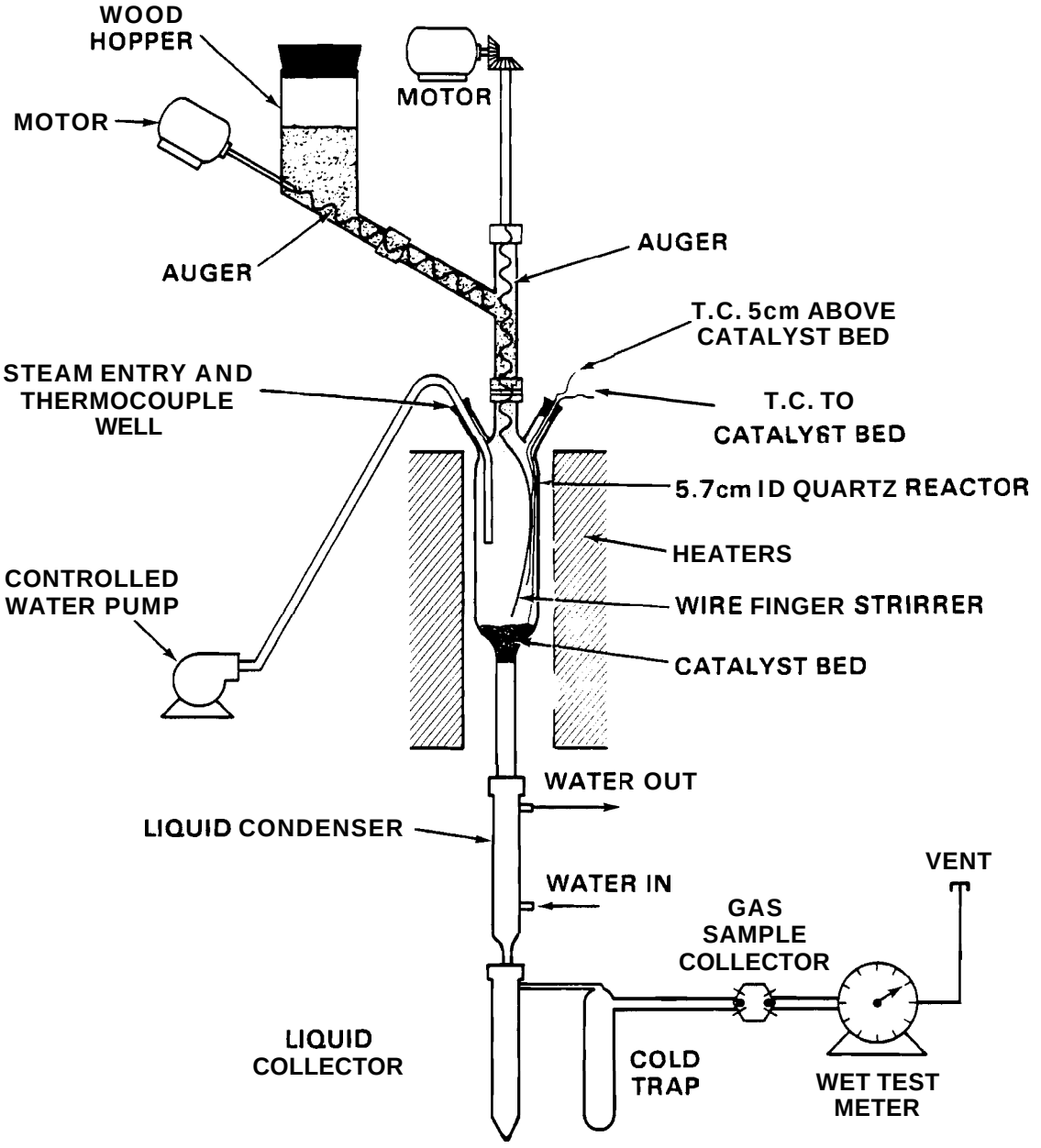

FIGURE 1. Laboratory Gasifier

TESTS WITH SUPPORTED METAL CATALYSTS

Previous catalyst screening studies with wood as a feedstock showed supported nickel catalysts to be the most effective for the production of methanol synthesis gas. Some of the nickel catalysts tested showed rapid deactivation; however, several catalysts were identified which showed the potential for a long lifetime. These included a series of trimetallic catalysts, supplied by

W. R. Grace, Inc. in a fluidizeable form (40-70 mesh spheres) (Mudge 1983, Baker 1982). A large batch of a Ni-Cu-Mo/SiO_2-Al_2O_3 catalyst was purchased from W. R. Grace, Inc. for PDU tests with wood and was also used for the bagasse studies. When this catalyst showed rapid deactivation in PDU tests with bagasse, laboratory studies were initiated to determine the cause of deactivation and identify other catalysts that could be used.

Test Results

The results of gasification tests with bagasse and nickel catalysts are shown in Table 2. The results of Run 2 are typical of an active nickel catalyst. This was a quick run to check the activity of a catalyst which was used for the PDU shakedown run and no loss of activity was noted in the short time of this run. Loss of catalyst activity is indicated by an increase in hydrocarbons (methane, ethane, ethylene) in the product gas and the appearance of significant quantities of tars and oils in the condensate from the reactor. Higher synthesis gas yield shown in the table is also a good overall indication of catalyst activity. Deactivated catalysts reform less methane and tars to hydrogen and carbon monoxide and typically have synthesis gas yields below 0.7 nm^3/kg.

Run 3 and 4 confirmed the early deactivation of the Ni-Cu-Mo/SiO_2-Al_2O_3 catalyst observed in the PDU. Doping of the catalyst with sodium which is sometimes used to reduce coking with steam reforming catalysts (Rostrup-Nielson 1975) did not help (Run 3). The gas composition and conversion data given in Table 2 is averaged over the entire run. At the start of the catalyzed runs the results were similar to Run 2 and then gradually deteriorated until they approached the results obtained without a catalyst (Run 6).

A steam reforming catalyst which showed potential for long lifetime in studies with wood also deactivated rapidly (Run 7). Addition of a ZnO sulfur guard catalyst extended the life of the NiCuMo/SiO_2-Al_2O_3 catalyst but not nearly long enough to be an attractive alternative (Run 12). A Ni-Mo/Al_2O_3 desulfurization catalyst was not active enough even at the start of the run (Run 14). No loss of activity was noted in a prior test with bagasse obtained from Hawaii, so we reran the Hawaiian bagasse and got results similar to the

TABLE 2. Laboratory Gasification Tests with Australian Bagasse and Nickel Catalysts[a]

Run No.	2	3	4	7	12	14	24	6
Catalyst	Ni/Al_2O_3	Ni-Cu-Mo w/10% Na	Ni-Cu-Mo SiO_2-Al_2O_3	Ni/Ceramic	ZnO + NiCuMo on SiO_2-Al_2O_3	NiMo/Al_2O_3	Ni/Al_2O_3	None
Reactor Temp., °C	770	770	770	780	790	710	750	780
Catalyst Temp., °C	735	730	720	720	720	750	760	720
Feed Rate g/min	0.22	0.19	0.23	0.20	0.24	0.27	0.15	0.20
Steam Rate g/g dry bagasse	1.4	0.9	0.64	1.0	1.06	0.6	1.4	0.9
Gas Composition, Vol. %								
H_2	58.2	44.3	42.4	46.4	50.7	39.3	51.8	30.9
CO_2	24.5	28.4	23.2	24.6	25.3	23.9	27.5	22.1
CH_4	1.8	10.2	9.7	8.5	6.8	11.6	6.0	14.4
CO	15.6	14.5	22.3	18.1	16.0	21.7	13.5	27.6
C_2	0.0	2.7	2.5	2.4	1.1	3.6	1.4	4.9
H_2S ppm in	--	190	240	200	--	--	85	100
H_2S ppm out	--	2	2	180	20	450	70	60
H_2/ CO	3.7	3.1	1.9	2.6	3.2	1.8	3.8	1.1
Synthesis Gas Yield (H_2 + CO) m^3/Kg dry bagasse	1.25	.64	.64	.64	.80	.42	.92	.55
Carbon Conversion, wt%								
to gas	84	76	71	67	72	70	85	63
to char	18	21	26	53	24	29	19	24
to liquids	tr	1	1	2	3	Tr	Tr	2
Carbon Balance (%)	102	98	98	121	99	99	104	89
Total Bagasse Fed (g)	27	45	100	61	79	100	77	96

(a) Results are averaged over the entire run.

TABLE 3. Comparison of Results with Bagasse from Australia and Hawaii[a]

Run Number	8	9	4
Feed Source	Hawaii	Hawaii	Australia
Catalyst	NiCuMo on SiO_2-Al_2O_3	NiCuMo on SiO_2-Al_2O_3	NiCuMo on SiO_2-Al_2O_3
Reactor Temp., °C	790	780	770
Catalyst Temp., °C	720	710	720
Feed Rate, g/min	.30	.20	0.23
Steam Rate, g/g dry bagasse	0.8	1.6	0.7
Gas Composition, Vol%			
H_2	44.3	45.7	42.4
CO_2	25.4	25.6	23.2
CH_4	9.4	9.3	9.7
CO	19.5	17.1	22.3
C_2	1.4	2.2	2.5
H_2S, ppm in	160-210	200	190-240
H_2S, ppm out	120-220	5-140	0-5
Synthesis Gas Yield, (H_2 + CO) m^3/kg dry bagasse	0.63	0.63	.64
H_2/CO	2.3	2.7	1.9
Carbon Conversion wt%			
to gas	67	69	71
to char	37	25	26
to liquids	<1	<1	1
Carbon Balance, %	104	94	98
Total Bagasse fed, g	100	100	100

(a) Results are averaged over the entire run.

Australian bagasse as seen in Table 3. A review of the previous test data showed some deterioration in gas quality near the end of the run indicating it may have been terminated before the loss of activity was readily apparent.

Several other supported catalysts were tested but they were not active enough to be considered further (Table 4). These catalysts did not show any loss of activity through the course of the run.

Cause of Catalyst Deactivation

The usual cause of catalyst deactivation in our gasification tests is carbon deposition. The trimetallic catalysts made by W. R. Grace, Inc. had long lifetimes in the laboratory tests despite some carbon deposition. One of these catalysts reached a lifetime of 1470 g wood/g catalyst without loss of activity. No long term tests have been made in the PDU.

The rapid loss of activity with the $NiCuMo/SiO_2-Al_2O_3$ catalyst in Runs 3 and 4 indicated the possibility of some other deactivation mechanism. During previous runs with Hawaiian bagasse and Run 2 with Australian bagasse a sulfur odor was detected by the gasifier operator. In Runs 3 and 4 the gases were analyzed for H_2S and the results are in Table 2. For the H_2S measurement identified as "H_2S ppm in" a gas sample was removed from the top of the reactor before the gas had passed over the catalyst. The results of both Runs 3 and 4 indicated that a significant quantity of H_2S was present and that it was being taken up by the catalyst. After the catalyst was saturated the H_2S content of the product gas rose to the level of the gas in the top of the reactor. The results of all of the H_2S measurements on gases are summarized in Table 5.

Samples of different feed materials and catalysts were also analyzed for sulfur and the results are shown in Table 5. There are some discrepancies in the data which we assume are due to the difficulty in getting a truly representative sample. The results do show that the Australian bagasse contains 200-400 ppm sulfur and the deactivated catalysts have a significant quantity of sulfur on them.

That H_2S adsorption on the catalyst contributes to deactivation is suggested by several factors:

TABLE 4. Laboratory Gasification Tests with Other Supported Catalysts[a]

Run Number	10	11	17
Catalyst	$CoMo/Al_2O_3$	25% K_2CO_3 on Al_2O_3	SiO_2-A_2O_3
Reactor Temp., °C	780	790	740
Catalyst Temp., °C	720	720	810
Feed Rate, g/min	.21	.20	.21
Steam Rate, g/g dry bagasse	1.2	1.2	0.8
Gas Composition, Vol%			
H_2	43.7	43.9	45.9
CO_2	27.7	31.7	20.1
CH_4	11.2	11.8	11.3
CO	15.7	9.5	21.3
C_2	1.6	3.1	1.4
H_2S, ppm in	--	--	--
H_2S, ppm out	140-200	200-400	--
H_2/CO	2.8	4.6	2.1
Synthesis Gas Yield (H_2 + CO) nm /kg dry bagasse	0.57	0.51	0.72
Carbon Conversion, wt%			
to gas	68	67	74
to char	33	28	18
to liquids	<1	<1	<1
Carbon Balance, %	101	95	92
Total Bagasse Fed, g	78	100	100

(a) Results are averaged over the entire run.

- H_2S is a known catalyst poison for nickel reforming catalysts in quantities as low as 2 ppm (Rostrup-Neilsen 1975).
- There is a close correlation between the amount of sulfur on the catalysts when they lose activity and the amount of sulfur required to sulfide the active metal surface area of the catalyst (Table 6).

TABLE 5. Sulfur Analyses for Bagasse Gasification Tests

Gas Samples Feed Material	ppm H_2S by volume in gas (before catalyst) Range	Average
Australian Bagasse	80-240	167
Hawaiian Bagasse	160-210	190
Wood	20-30	25

Solid Samples	ppm H_2S by weight (a)	(b)	(c)
Australian Bagasse Pellets	1200	250	310, 375
Bales	--	--	230, 245
Hawaiian Bagasse (pellets)	--	627	--
Wood	--	47-145	--
Bark	--	204-330	--
Catalyst from Laboratory Run #4	1600	3000	--
Catalyst from PDU at the end of C-2	2000	250	--
Catalyst from PDU after regeneration at end of C-2	--	830	--

(a) Hanford Engineering Development Laboratory, Richland, WA. Leco Analyzer
(b) Schwartzkopf Microanalytical Laboratory, Inc. Woodside, NY. Leco Analyzer
(c) Sugar Research Institute. ICP

- There is a good comparison between the time that should be required to sulfide the catalyst and the time when deactivation occurs. Dividing the amount of sulfur on the catalyst (#3 in Table 6) by the rate of sulfur input the system (#1 or #2) indicates 5-8 hours should be required to sulfide the catalyst. Deactivation in the laboratory studies was readily apparent 3-5 hours after startup.

TABLE 6. Sulfur Calculations for Laboratory Run 4 with $NiCuMo/SiO_2-Al_2O_3$ Catalyst

1.	Sulfur in the Feed 300 ppm x 13.8 g/hr	4.1×10^{-3} g/hr
2.	Sulfur as Hydrogen Sulfide in Gas 167 ppm x 17.7 ℓ/hr x 0.8 g/ℓ	2.4×10^{-3} g/hr
3.	Sulfur on the Catalyst at end of run 2000 ppm x 10g catalyst	2.0×10^{-2} g
4.	Sulfur required to Saturate Monolayer of Catalyst	
	9×10^{-5} atoms H/g catalyst[a] x 10g catalyst x 0.74 atoms S/H[b] x 32g/atoms	2.1×10^{-2} g

(a) Measured using a hydrogen chemisorption apparatus described by Yudge, et al. (1983).
(b) J. R. Rostrup-Nielsen, Steam Reforming Catalysts. Danish Technical Press, Copenhagen 1975, pp. 63-/6.

- Adding ZnO sulfur guard catalyst to the system extended the life of the $NiCuMo/SiO_2-Al_2O_3$ catalyst (Run 12).
- A catalyst with more active surface area (Ni/Al_2O_3) takes longer to deactivate than $NiCuMo/SiO_2/Al_2O_3$. Just the opposite was true with wood where carbon deposition was the primary cause of deactivation.

Wood does have some sulfur, although it is lower than bagasse. A similar catalyst ran for over 1000 hours in a laboratory test with wood without deactivating. Rostrup-Nielsen (1975) in studies of H_2S chemisorption of hydrogen sulfide on steam reforming catalysts found the chemisorption of hydrogen sulfide on nickel to be reversible, the coverage on the catalyst being a function of the ratio pH_2S/pH_2. The amount of adsorbed sulfur in equilibrium with the H_2S in the gas increases with increasing H_2S concentration until constant fractional coverage of the surface is reached. Using a 13.3% nickel catalyst at 627°C the constant fractional coverage was reached at 50-60 ppm H_2S (Stern 1982). In another study (Houghtby, 1980) a catalyst found to be active for reforming ethane in the presence of sulfur deactivated quite rapidly on No. 2 fuel oil. The same catalyst was presulfided and was initially quite

active for reforming fuel oil. In both cases significant carbon was present on the catalyst. Only when the sulfur level in the fuel oil was less than 100 ppm would the catalyst maintain its original activity for an extended period of time and no carbon deposition was noted (Houghtby 1982). Therefore, it appears that the low sulfur levels in wood allow enough of the nickel on the catalyst to remain exposed and active.

Heavy deposition of carbon was also found on the catalysts, particularly those from the PDU. Figure 2 shows the accumulation of carbon on nickel catalysts used with bagasse. Carbon deposition was higher in the PDU than the laboratory and both are significantly higher than we have seen with wood. The $NiCoMo/SiO_2-Al_2O_3$ which ran over 1000 hours in the laboratory had only 5 wt% carbon on it at the end. Carbon deposition was a general indicator of deactivation and no catalysts were found which were active with more than 10 wt% carbon on them (Mudge 1983).

Based on the high carbon deposition on the catalysts used with bagasse and the results of Houghtby (1980) one could conclude that the sulfided catalysts are more susceptible to carbon deposition which actually causes the loss of activity. However, it is generally agreed that sulfiding nickel reforming catalysts decreases carbon formation on the catalyst (Rostrup-Nielsen 1975; McCarty 1981). Sulfiding reduces the activity of the catalyst and in a commercial reformer may allow the feed material to pass to a hotter part of the reactor where thermal cracking and coking may occur.

From our studies it is not possible to positively identify the mechanism for deactivation, but it appears that it is definitely a function of the sulfur content of bagasse. Coking may also play a part but the exact relationship between the sulfur adsorption and carbon deposition is not known.

TESTS WITH ALKALI CARBONATE CATALYSTS

The application of alkali metal catalysts to coal gasification has been known for many years (Taylor 1921) and has been studied extensively in recent years (Cox 1974; McKee 1983; Walker 1983; McCoy 1983; Nahas 1983). These catalysts have also been studied for use with biomass (Mudge 1979; Hawley 1983; Sealock 1982).

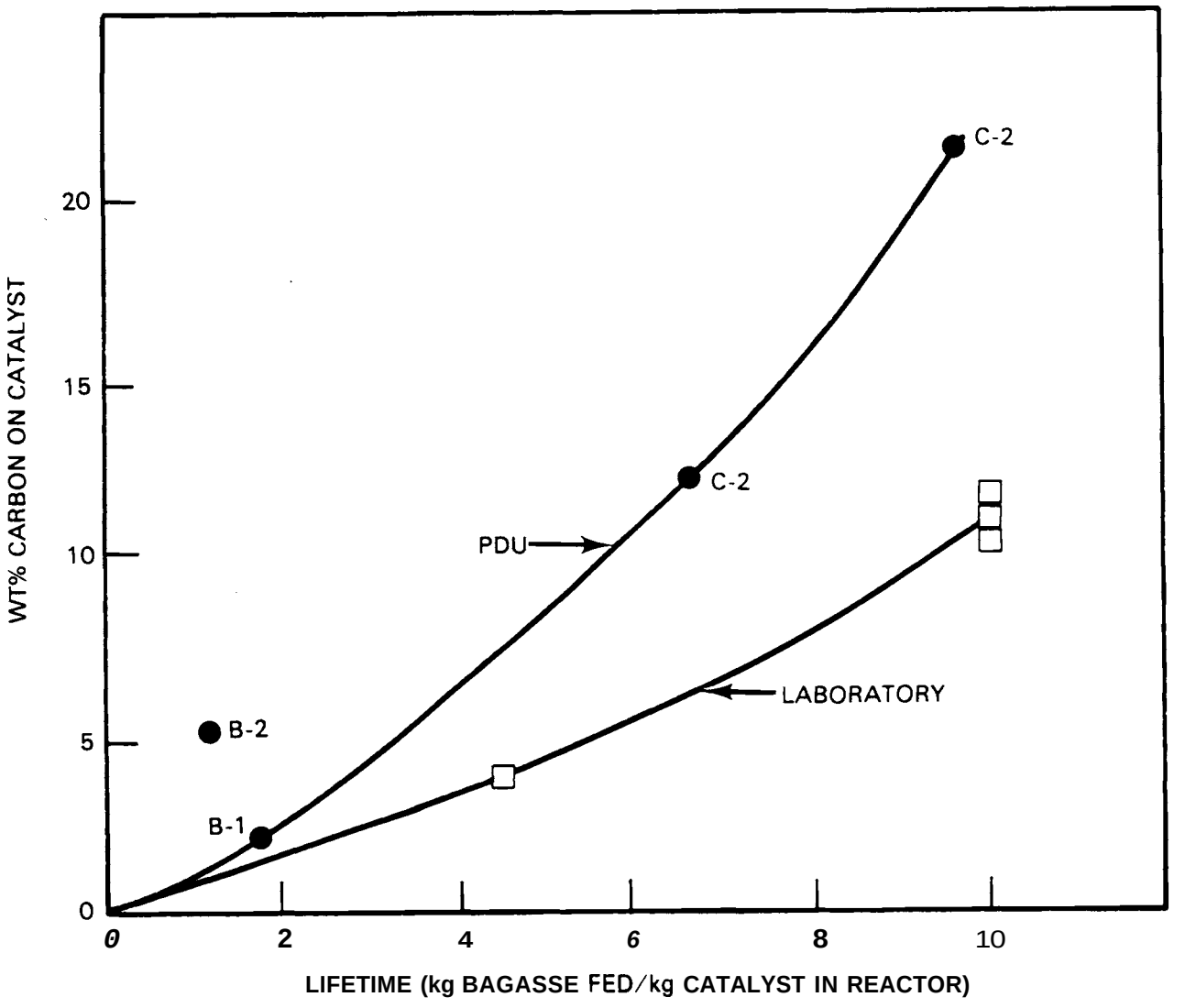

FIGURE 2. Carbon Accumulation on Nickel Catalysts

Typically alkali metal salts are solution impregnated or dry mixed with the feed material to improve the kinetics of the reaction of char with steam, carbon dioxide, and other gases. In the course of our studies with wood we found that alkali catalysts also reduce the yield of tars, oils, and gaseous hydrocarbons and catalyze the water gas shift reaction making it possible to produce a high quality synthesis gas without the use of a secondary catalyst.

When it became apparent that we would be unable to find a long-lived supported nickel catalyst for bagasse gasification we switched our attention to the alkali metal salt catalysts. The alkali metal salts are not quite as

active as fresh nickel catalysts in terms of synthesis gas yield, but because catalyst is being added continuously there is no loss of activity. The catalyst can be recovered from the char and recycled.

Test Results

Table 7 shows the results of laboratory tests with alkali metal salt catalysts. These tests showed the potential of this type of catalyst for the production of methanol synthesis gas from bagasse. There are many possible catalyst variations that can be studied including cation type (K, Na, Li, Cs, Ca) anion type (CO_3, OH, Cl), catalyst loading, and contacting method (solution impregnation or dry mixing). A complete investigation of these variables was not within the scope of this project. The selection of those catalysts tested was based on our previous work with wood and data from the literature.

Alkali metal carbonates are generally the most active catalysts for gasification of char with H_2O and CO_2 (McKee 1983, Johnson 1981). Potassium carbonate is generally recognized as offering the best combination of activity and low cost. In other laboratory studies at PNL with biomass we found the differences between K_2CO_3 and Na_2CO_3 to be small (Sealock 1982). In PDU tests where the catalyst was physically mixed with wet wood, K_2CO_3 was much superior to Na_2CO_3 (Mudge 1983). We attributed this primarily to the fact that the moisture content of the wood was sufficient to dissolve most of the K_2CO_3, thus effectively impregnating the K_2CO_3 on the wood. Sodium carbonate is less soluble and remained primarily as a solid when mixed with the wood.

Laboratory tests with bagasse showed little difference between K_2CO_3 and Na_2CO_3 when both were impregnated (Runs 5 and 15). A mixture of the two also performed about the same.

In another project for the U.S. Department of Energy we are currently looking at the effect of catalyst loading on the gasification of wood. Preliminary results from this work show only a small difference between 17 wt% and 8 1/2 wt% K_2CO_3. Based on this we selected 10 wt% K_2CO_3 as a starting point for the bagasse studies. The results are about the same as we have seen for wood even at higher catalyst loadings. Run 19 with 5 wt% K_2CO_3 showed a significant increase in char and liquid yields.

TABLE 7. Laboratory Gasification Tests with Alkali Carbonate Catalysts[a]

Run Number	5	13	15	16
Catalyst	10 wt% K_2CO_3	10 wt% K_2CO_3 dry mixed w/bagasse	10 wt% Na_2CO_3	5 wt% Na_2CO_3, 5 wt% K_2CO_3
Reactor Temp., °C	750	740	735	735
Feed Rate, g/min	.19	.21	.27	.4
Steam Rate, g/g bagasse	0.9	0.8	0.7	.4
Gas Composition, Vol%				
H_2	49.8	46.0	50.0	50.5
CO_2	23.7	24.5	21.3	18.9
CH_4	5.9	6.9	5.4	4.9
CO	18.6	19.8	20.8	23.3
C_2	2.1	2.8	2.5	2.3
H_2S, ppm in	80-120	--	--	--
H_2S, ppm out	15-60	--	--	--
H_2/CO	2.7	2.3	2.4	2.2
Synthesis Gas Yield (H_2 + CO) nm /kg dry bagasse	1.0	.85	0.94	.92
Carbon Conversion, wt%				
to gas	93	87	82	78
to char	8	15	13	20
to liquids	<1	1	2	<1
Carbon Balance, %	101	103	97	98
Total Bagasse Fed, g	90	90	86	32

(a) Results are averaged over the entire run.

TABLE 7. (contd)

Run Number	19	20	21	22	23
Catalyst	5% K_2CO_3	10% K_2CO_3	10% K_2CO_3 Dry Mixed at PDU	10% K_2CO_3 Impregnated at PDU	8% KOH
Reactor Temp., °C	745	755	755	750	750
Feed Rate, g/min	.31	.45	.18	.18	.26
Steam Rate, g/g dry bagasse	0.6	0.4	1.1	1.1	0.8
Gas Composition, Vol%					
H_2	50.4	50.3	53.9	54.1	54.4
CO_2	21.5	15.0	23.7	24.0	18.5
CH_4	5.3	3.7	5.0	4.7	3.3
CO	20.2	29.2	15.9	15.8	22.9
C_2	2.6	1.8	1.5	1.5	1.9
H_2S ppm in	--	--	--	--	
H_2S ppm out	--	--	--	--	
H_2/CO	2.5	1.7	3.4	3.4	2.4
Synthesis Gas Yield (H_2 + CO) nm^3/Kg dry bagasse	0.81	1.07	1.07	1.11	1.09
Carbon Conversion, wt%					
to gas	75	85	89	92	86
to char	18	8	5	2	7
to liquids	4	<1	<1	<1	<1
Carbon Balance, %	97	94	95	94	93
Total Bagasse Fed,g	93	82	45	58	166

We have tested two methods of contacting the alkali salt with biomass: 1) impregnating a solution of the salt on the biomass, and 2) dry mixing the salt with the biomass. Moist biomass materials readily absorb dry mixed potassium carbonate powders. Sodium carbonate is much more difficult due to decreased solubilities, and requires solution in hot water. All of the tests in Table 7 were made with impregnated bagasse except where noted. The results of Run 13 with 10% K_2CO_3 dry mixed with dry bagasse were not quite as good as those obtained with an impregnated feedstock.

Runs 21 and 22 were made with feed materials prepared at the PDU. The bagasse in Run 21 was prepared by wetting dry bagasse pellets to 10-15 wt% moisture and then mixing the K_2CO_3. The bagasse in Run 22 was sprayed with a 40 wt% solution of K_2CO_3. The results were quite good with both materials indicating either method could be used to prepare the bagasse for PDU runs.

A significant finding on these laboratory bagasse studies is that a much lower than expected steam to bagasse ratio is needed to achieve the 2 to 1 hydrogen to carbon monoxide ratio required for methanol synthesis. With wood and a nickel secondary catalyst a steam to wood weight ratio of 0.7 was required for a 2 to 1 hydrogen to carbon monoxide ratio. For bagasse with the alkali metal catalysts it appears that 0.4 to 0.5 kg of steam per kg of bagasse would be appropriate as shown in Figure 3.

Catalyst Recovery

Because of the cost of alkali metal catalysts and the loadings required it is necessary for economic reasons to recover the catalyst from the gasification residue (char, ash, catalyst). Previous studies with wood showed 90-95% of the potassium charged to the gasifier could be recovered by washing with large quantities of water (Mudge 1981). Exxon is currently planning on 90% recovery of the potassium used in their catalytic gasification process (Furlong 1978, Fant 1980).

The quantity of residues generated from a bagasse gasification run with 10% K_2CO_3 will range from about 0.2 kg/kg of dry bagasse at 90% carbon conversion to 0.25 kg/kg dry bagasse at 80% carbon conversion. The residue will be 40-50 wt% K_2CO_3 with the remainder being ash and char.

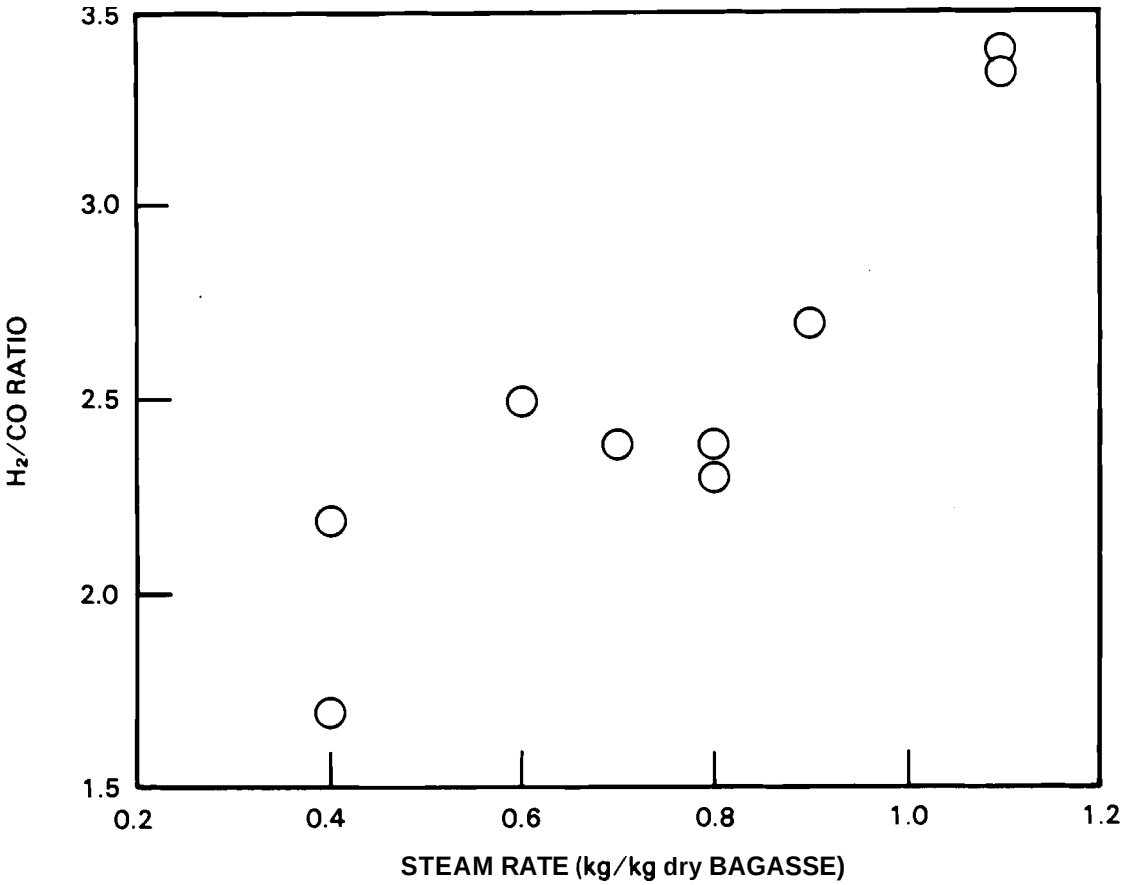

FIGURE 3. Effect of Steam Rate on the H_2/CO Ratio in the Product Gas with K_2CO_3 Catalyst

The lowest water/residue ratio used in previous testing was 10 ml/g (Mudge 1981). Four samples of residue from Run 20 were washed with smaller quantities of water and the solutions analyzed by atomic absorption. The results are shown in Figure 4. About 80% recovery was achieved with as little as 2 ml of water per gram of residue. Additional water did not significantly improve recovery.

The remainder of the potassium probably reacts with mineral matter in the bagasse to form water insoluble compounds. Wood has very little ash so higher recoveries are possible. Exxon digests the char from their catalytic coal gasification process with $Ca(OH)_2$ to free additional water-soluble catalyst to increase their recovery to 90% (Furlong 1978; Fant 1980). This method could probably be used with bagasse char to increase the recovery of potassium.

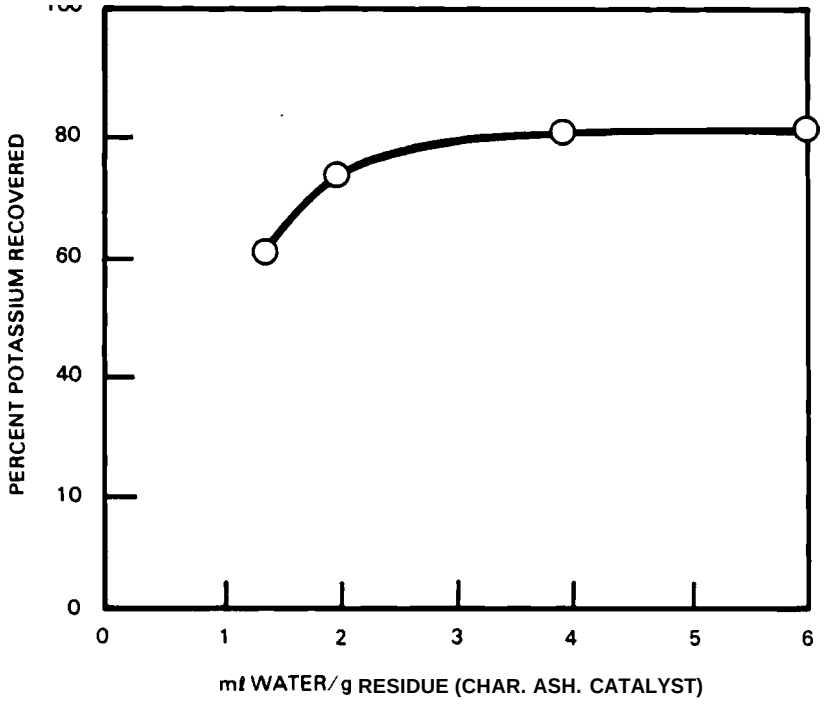

FIGURE 4. Catalyst Recovery from Gasification Residue

PROCESS DEVELOPMENT UNIT STUDIES

The main objectives of the process development unit (PDU) studies are to determine the technical feasibility of producing a methanol synthesis gas from bagasse and to provide information on equipment and catalyst performance for further scale up.

PDU DESCRIPTION

The process flow diagram for the PDU is shown in Figure 5. Steam is superheated in the gas heater and contacts bagasse in the fluid bed gasifier shown in Figure 6. Product gas and char exiting the gasifier are separated by a cyclone and filter. A heat exchanger condenses steam and organic compounds which are then separated from the gas in the demister. The gas can be recycled to the gas heater via the recycle compressor. Product gas is released through a letdown valve. A parallel off-gas system employs a venturi scrubber; however, it was not used and will not be discussed. All other components are discussed in detail.

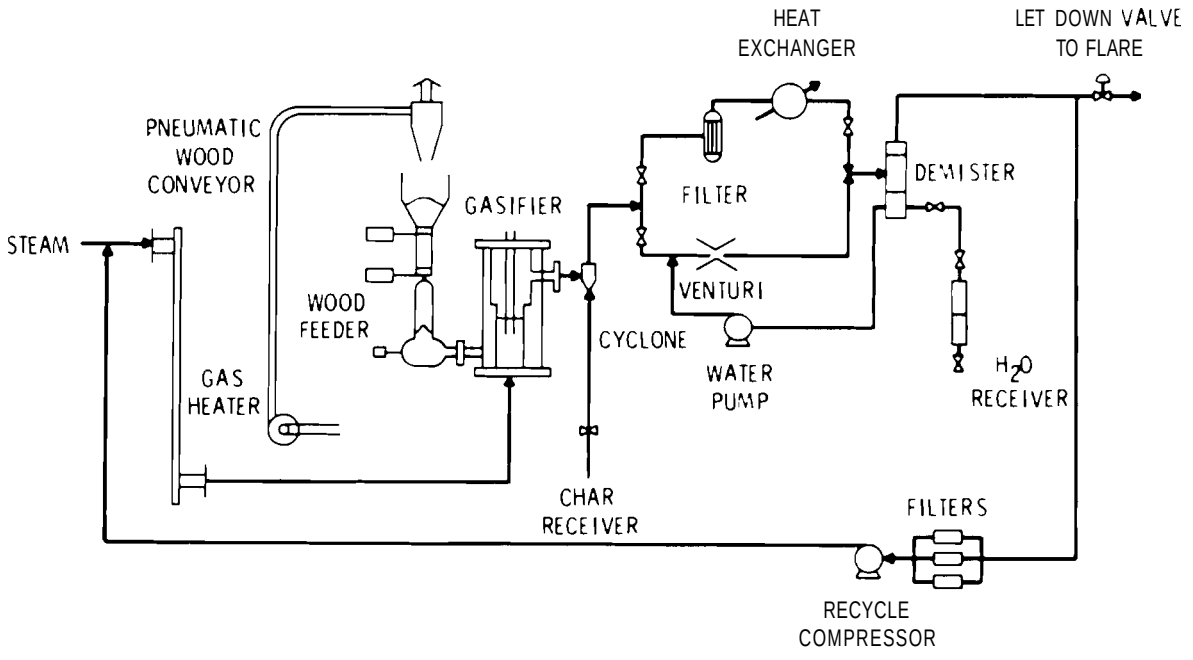

FIGURE 5. PDU Flow Diagram

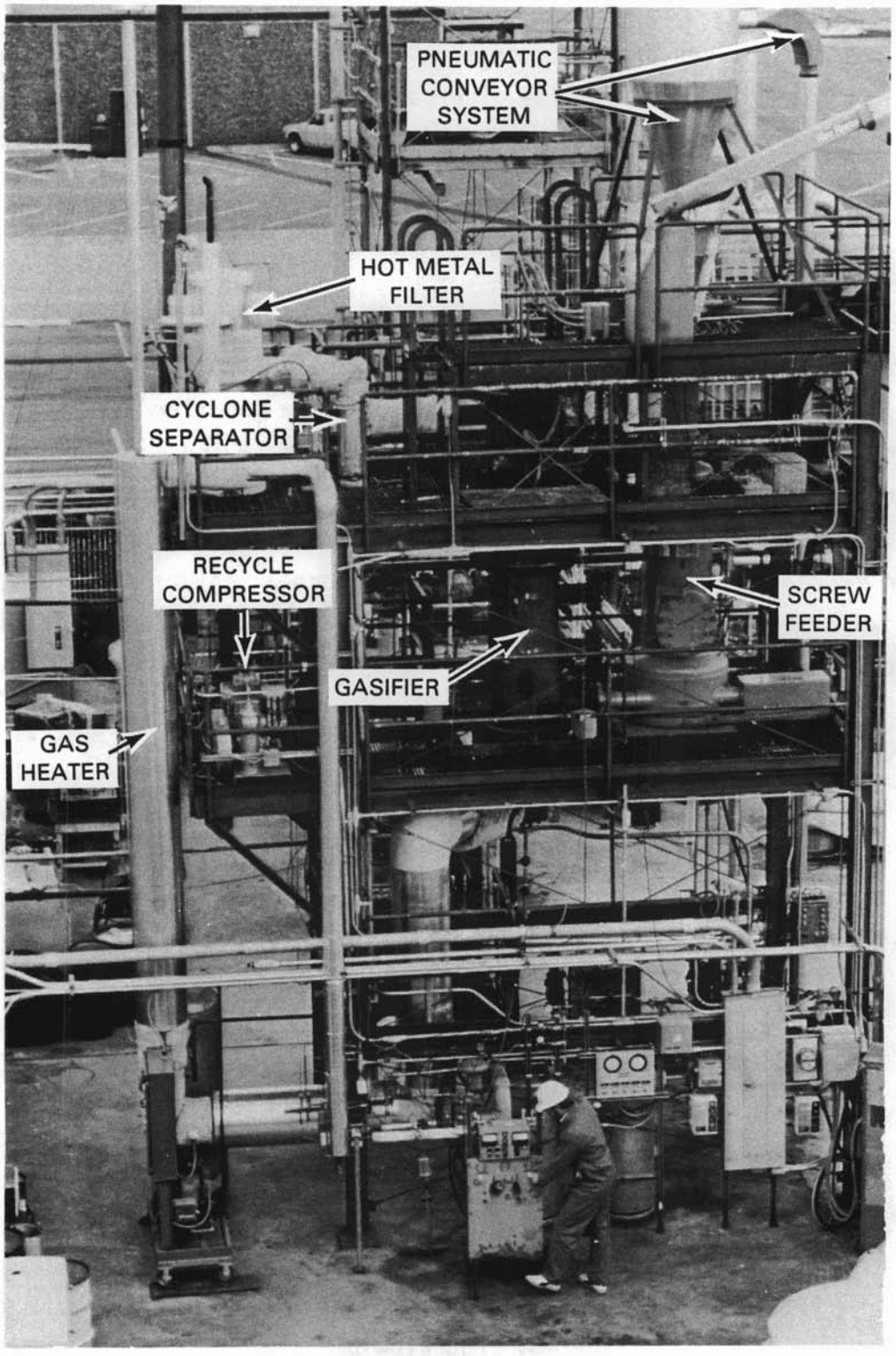

FIGURE 6. Fluid Bed Gasifier

Gasifier

The gasifier is a 0.61 m (24-in.) carbon steel pipe, 3 m (10 ft) long, with blind flanges at each end and a flanged connection near the middle of the vessel. A schematic is shown in Figure 7. The gasifier has ceramic fiber

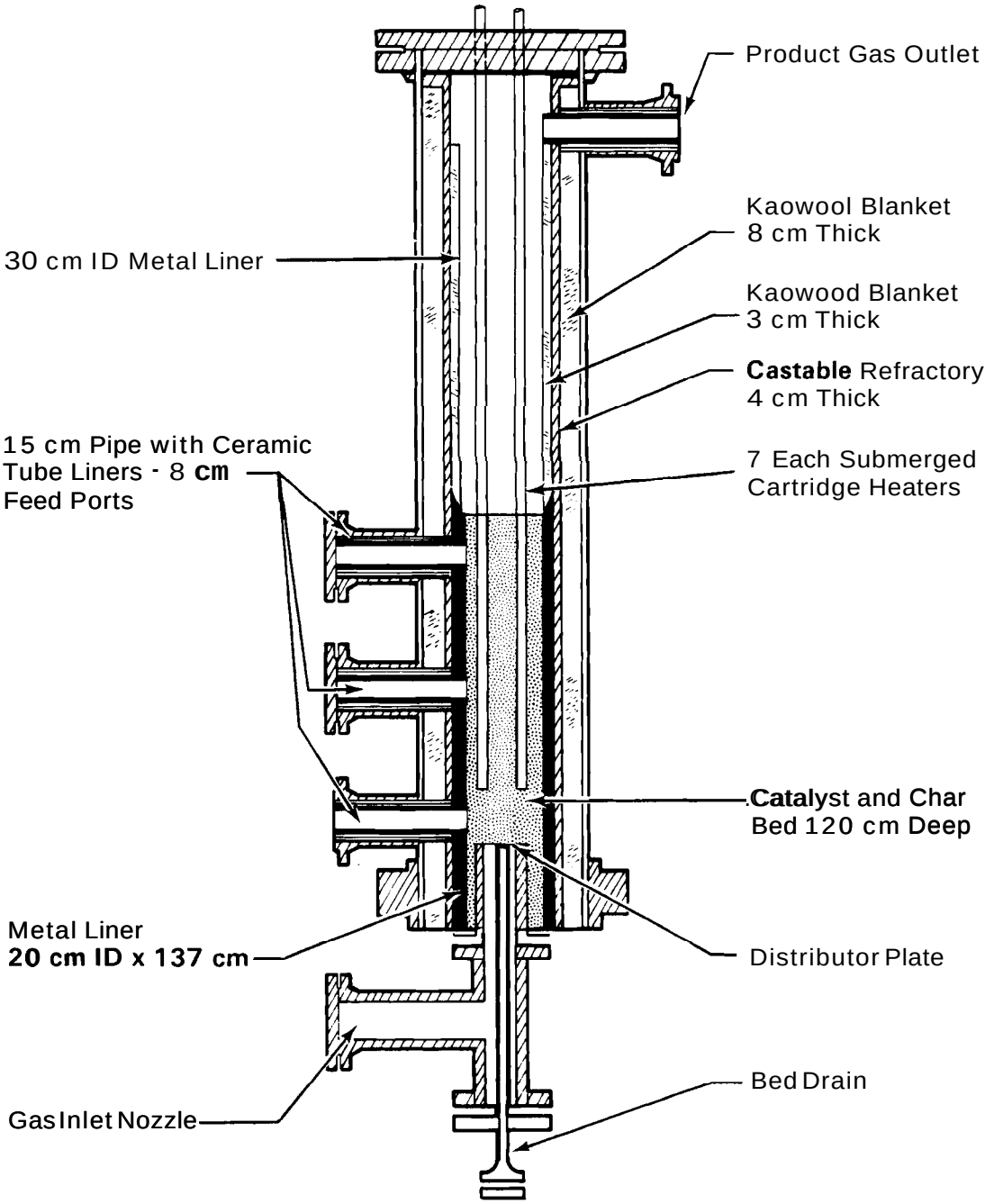

FIGURE 7. Schematic of Fluid Bed Gasifier

insulation [approximately 8 cm (3 in.) thick] and a cast refractory lining 4.25 cm (1.5 in.) thick. The reaction zone (bed) is in a 20 cm (8 in.) diameter tube of stainless steel. The bed is about 0.91 m (3 ft) deep when stagnant and 1.20 m (4 ft) deep when fluidized. Seven cartridge heater elements 1.9 cm (0.75 in.) diameter are inserted into the reaction zone. Their heated length is 1.20 m (4 ft) and they are capable of maintaining 870°C (1600°F) temperature. Total capacity is 127 MJ/hr (120,000 Btu/hr). These elements provide energy to heat reactants (bagasse) to the bed temperature and energy for the heat of reaction. Energy consumption by these elements is measured by a watt-hour meter. Our experience showed that the heat transfer from the elements to the bed limited the output from the heaters to about 80 MJ/hr (75,000 Btu/hr) output. The effective heat transfer coefficient is about 1600 kJ/hr-m^2°C (80 Btu/hr-ft^2°F).

Gas enters the gasifier through a refractory lined pipe tee. A distributor plate made of Incoloy 800H (trade name of Huntington Steel Corp.) is placed just below the wood feed inlet. The distributor plate has 144 holes of 1.3 mm (0.05 in.) diameter.

A freeboard disentraining area of 1.3 m (6 ft) is above the fluid bed. A 30 cm (12 in.) diameter stainless steel liner was added to 1) prevent spalling refractory from falling into the bed and destroying the fluidization and 2) inhibit steam diffusion to the gasifier wall.

Thermocouples and pressure probes are inserted via 6.4 mm (0.25 in.) tubing from the reactor top to the desired level in the gasifier.

Biomass Lockhopper and Feeder

The fuel feeding system is shown in Figure 8. The feeder is enlarged to handle a variety of feedstocks and is designed for high pressure. The lockhopper valves are self-cleaning-knife gate valves. The lockhopper is a 0.3 m (12 in.) diameter pipe, 1.22 m (4 ft) long.

The live bin is 0.61 m (2 ft) in diameter and about 1.8 m (6 ft) tall. It contains three metering screws of 15 cm (6 in.) diameter. These metering

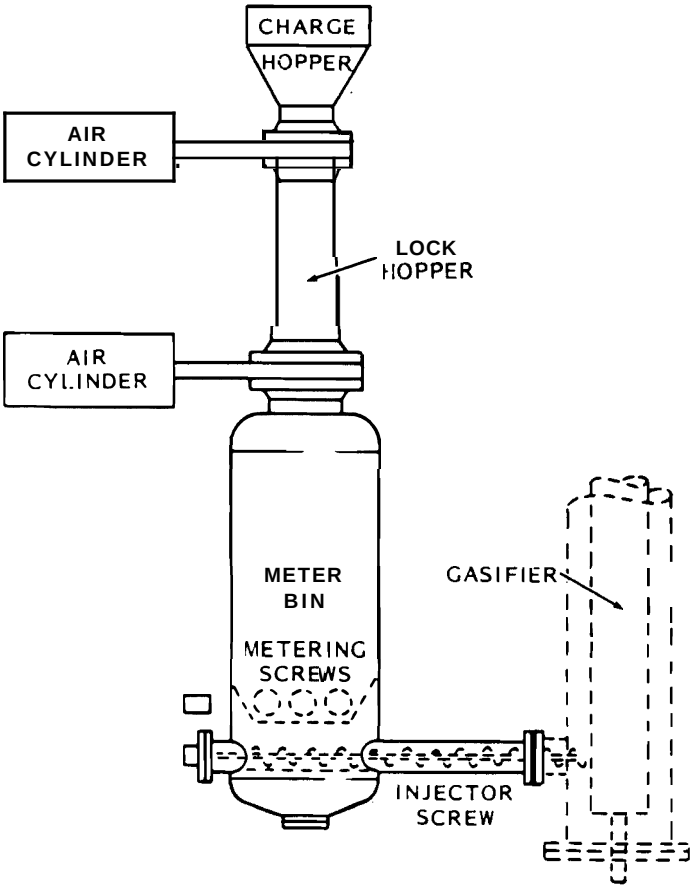

FIGURE 8. Fuel Feeding System

screws push biomass into an injector screw 8.9 cm (3.5 in.) diameter which rapidly conveys the fuel into the fluid bed. The injector screw typically operates at 150 rpm.

Levels of bagasse in the live bin are indicated by two types of sensors: a vibrating plate device and an energy transmission device. The antenna for the energy transmission devices were vertically oriented in the bin to obtain satisfactory calibration and operation.

A nitrogen purge in the feeder is used during all tests to prevent any flow of steam from the gasifier into the live bin. This technique was successful because bagasse removed from the live bin at the end of tests was never moist.

The feeder is driven hydraulically with manual controls for injector and metering screw speeds. The injector screw can be moved in and out while in operation. The 15 cm (6 in.) travel on the injector is used during startup when stagnant catalyst has settled on the screw and in the injector shaft.

The metering screws are electrically interlocked with the injector so they cannot operate without the injector. The feeder is interlocked with the data-logger to stop feeding in alarm conditions.

The lockhopper is manually operated with electrical interlocks which prevent both hopper valves from opening at the same time. Interlocks also prevent the top valve from opening without the pressure vent line open. A differential pressure switch prevents opening the bottom hopper valve until pressures in the live bin and the hopper are equalized.

Steam Superheater

The gas preheater heats steam and/or recycle gas to the desired bed temperature (or higher if possible). The preheater was designed to heat steam (27 Kg/hr) from 100°C (212°F) to 980°C (1800°F). This is a duty of 50 MJ/hr (48,000 Btu/hr). The heater is a propane fired, annular finned tube made of stainless steel and a cast chrome-nickel alloy.

The gas heater and the line between the heater and the reactor [7.5 cm (3 in.) 316 stainless steel] are both heavily insulated and waterproofed. Heat loss in the line can cause extremely slow heatups, so high temperature heat tapes are attached to the pipe, effectively eliminating heat losses in the line.

Controls on the gas heater include a temperature controller which activates a propane valve positioner. A high-temperature controller shuts down the heater if the finned tube temperature exceeds 810°C (1490°F). High and low pressure switches on the propane supply are additional safety features.

Product Gas Cleaning

When biomass of any kind is gasified in the PDU there are three distinct product categories: product gases, condensible liquids (tars, soluble

organics, and water), and chars. The efficient separation of these three is essential for the continued operation of the plant during a run and for determination of material balances.

Most of the char is separated from the gas in an efficient 7.5 cm (3 in.) diameter high temperature cyclone. The char which passes through the cyclone is separated using seven high-temperature sintered stainless steel filters 6.5 cm (2.6 in.) outside diameter by 1 m (3.3 ft) long. Filter pore size is 10 microns. The filter vessel houses the filters and is equipped with a differential pressure gauge to measure pressure drop across the filter elements. A nitrogen pulse backflush system is used to remove the char cake from the elements. The cyclone and filter vessel are insulated and are operable at 540°C (1000°F). Gas leaving the filters passes through a heat exchanger where steam and any organic compounds condense. The condensate is collected in a demister column with a 10 cm (4 in.) diameter teflon packed demister. Clean gas passing through the demister screen is either recycled or released to a flare via a pneumatically controlled letdown valve.

Product Gas Recycle System

Gas is recycled during startup, catalyst reduction, and operation by a piston type recycle compressor. The recycle flow is measured with an orifice meter and is manually controlled by a variable speed drive. When required recycle flow rates are below the compressor's minimum capacity, a valve in a bypass loop is manually opened and adjusted.

Instrumentation and Data Acquisition

Steam flow is measured by differential pressure across an orifice meter. The differential pressure is converted to a signal by a pneumatic differential pressure cell. The signal feeds a pneumatic controller inside the laboratory. The controller operates a 2.5 cm (1 in.) pneumatic control valve which fails to a closed position on loss of signal.

Temperature controllers are used on the in-bed cartridge heaters. The heater elements contain internal thermocouples which measure the skin

temperature. A high-temperature limit control is used on this circuitry. Silicon control rectifiers are used to maximize heater life and to provide steady energy input.

On 30 minute intervals operators take readings of controller indicators, meters, pressure gauges and differential pressure gauges. In addition, a datalogger continuously monitors 32 thermocouples, 5 pressure transducers, and the product gas differential pressure transducer. The datalogger prints on 15 minute intervals and plots important points on 2 minute intervals. Important data points are recorded on magnetic tape on 2 minute intervals. Once the test is completed, data from the magnetic tapes can be extracted and plotted automatically.

Gas Sampling

Gas samples are drawn from the filter vessel outlet or from the demister (see Figure 5). The gas is cooled in a small heat exchanger and filtered. Gas pressure is reduced to 115 kPa (2 psig). The low pressure lines convey gas to the laboratory gas analysis equipment.

The gas sample flow is continuous (about 0.1% of product flow) during PDU operation. Readings from continuous CO, CO_2, H_2, O_2, and CH_4 analyzers are recorded by operators every 20 minutes. A gas chromatograph with a thermal conductivity detector is also used for gas analysis. This unit determines concentrations of N_2, C_2H_4, C_2H_6, C_3H_8, acetylene and butane as well as the gases monitored by the continuous analyzers. Samples are automatically injected into the gas chromatograph for analysis on 20 minute intervals.

Product gas flow rate is measured with a differential pressure gauge and an electronic differential pressure transducer connected with an orifice meter.

Data Analysis

After a PDU test, preliminary results are determined using estimated compositions (moisture, ash, C, H, O, and total organic content). Then, when chemical analyses are complete, a final analysis of the test is made using measured values. This analysis is programmed on a microcomputer. The calculated results include elemental mass balances, energy balances, conversions, and efficiencies.

PRELIMINARY FEEDING TESTS

Three types of bagasse were supplied by Sugar Research Institute; Loose (as dried), baled, and pelletized (1.3 cm and 1.6 cm pellets) (Figure 9). The loose bagasse consisted of fiberous chaff similar to long grass clippings. Occasional long fibers (up to 60 cm) were present. This loose material was shipped in large (~3 m^3) burlap bags. Forty bales of compacted material were received. The baled material was full of fines as well as long fibers. Pelletized material had been milled prior to pelletizing. Even the pelletized material showed a visible amount of fines.

Prior to making a run with bagasse it was necessary to calibrate the screw feeder. An attempt was made to feed the loose material directly. The injector screw compacted the loose bagasse and completely bound up after a short time. Calibrations with pellets were much better. Repeatable calibrations were obtained with bagasse both plain and impregnated with 10% K_2CO_3 catalyst. Overall feed rates that were derived from each run agree closely with the original calibrations shown in Figure 10.

Loose material might be fed to a larger gasifier, but a novel approach possibly employing a different type of screw or a sloped injector would be required. Until further research can be done on roll compaction and other methods of densification and feeding it is recommended that pelletized material be used. When the pellets were impregnated with K_2CO_3 some breakdown of the pellets occurred but they still fed uniformly indicating size reduction may be as important as pelletization. With a specially designed lock hopper and screw feeder it may be possible to feed loose material which has been reduced in size.

PDU GASIFICATION TESTS WITH NICKEL CATALYSTS

Extensive laboratory testing with wood showed nickel based catalysts to be the most effective for production of synthesis gas. With an active catalyst the product consists primarily of H_2, CO and CO_2. Nearly all of the hydrocarbon gases, tars, and oils are reformed. Long catalyst lifetimes were shown in laboratory tests but had not been verified in the PDU.

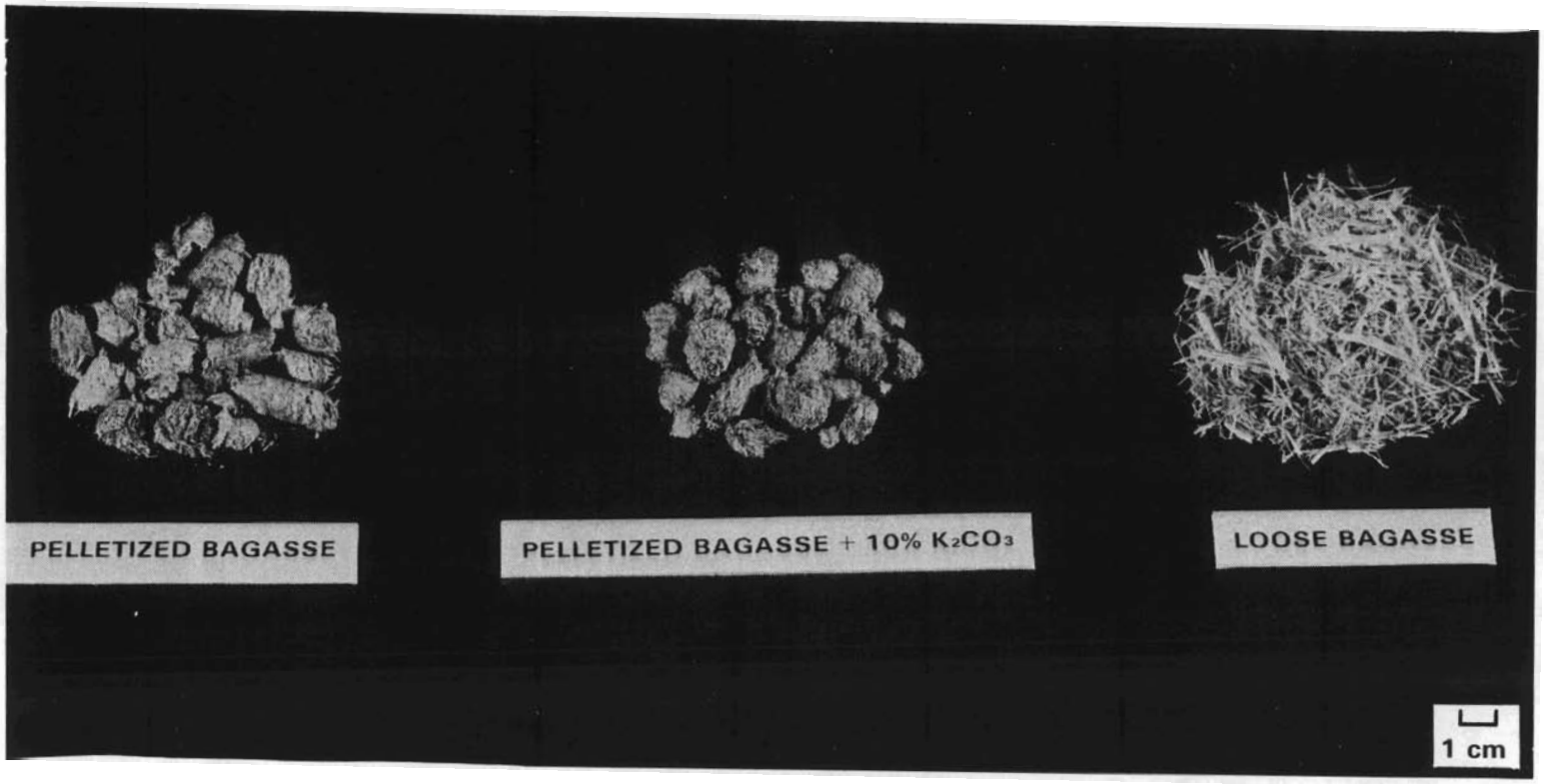

FIGURE 9. Feed Materials for PDU Tests

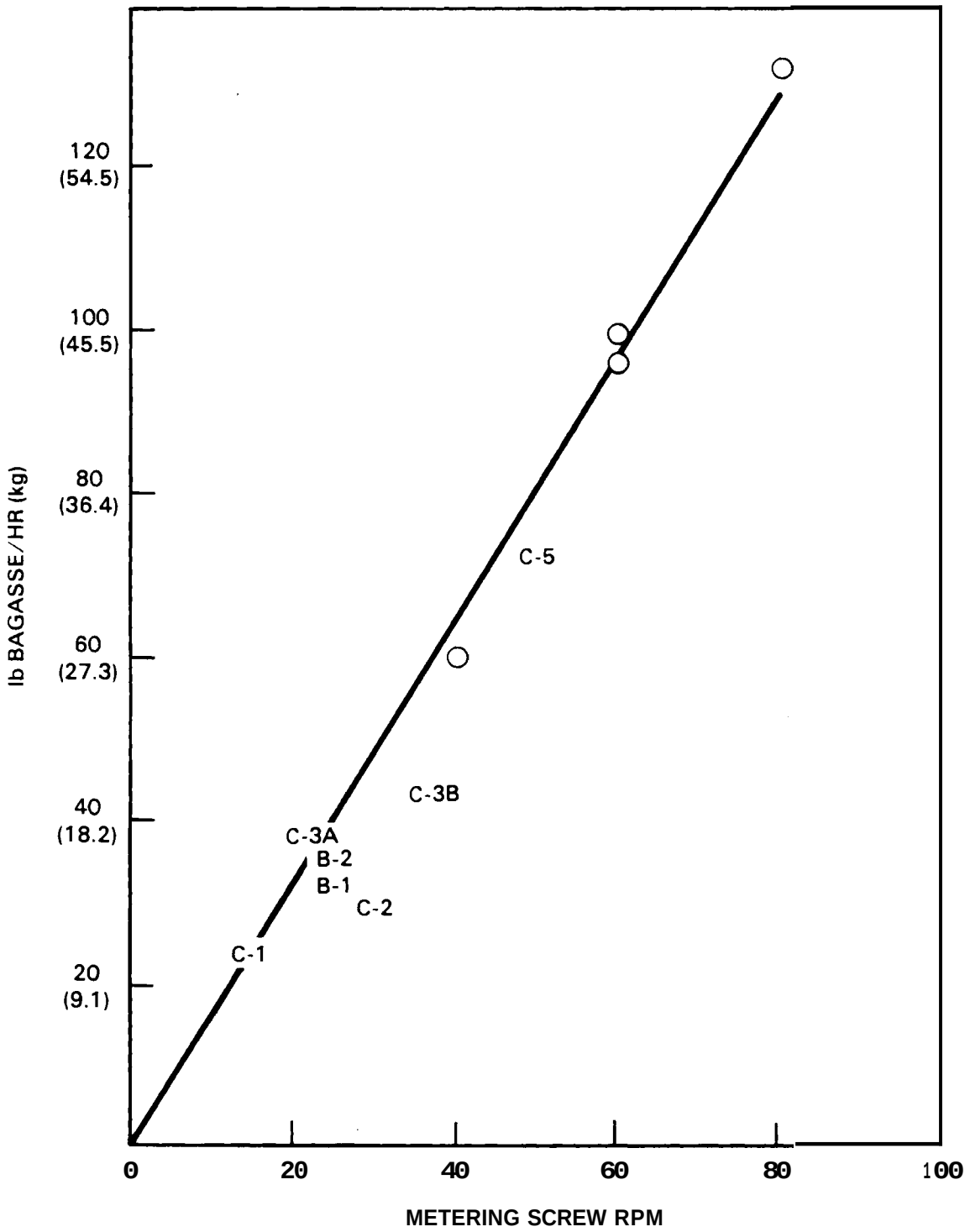

FIGURE 10. Calibration of Feed Metering Screw with Bagasse

The original plan for the PDU tests was for 1 or 2 short shakedown runs followed by a week-long-run with the $NiCuMo/SiO_2-Al_2O_3$ catalyst judged best for wood. A week-long-run would allow time for different process parameters to be tested and provide some indication of the amount of catalyst deactivation that could be expected.

Initial results from the PDU using nickel catalysts were quite encouraging. In the shakedown run and the early part of the week-long-run the results were similar to those achieved with wood. However, as the week-long-run progressed catalyst activity deteriorated rapidly and the run was eventually terminated after only one day. Catalyst deactivation was readily apparent from the CH_4 and H_2 content of the gas versus time as shown in Figure 11. Laboratory studies were initiated to determine the cause of catalyst deactivation and to search for other catalysts.

Five distinct PDU tests were made using both Ni on Al_2O_3 and Ni-Cu-Mo on $SiO_2-Al_2O_3$ catalysts. Runs B-1, B-2A, and B-2B all used reduced Ni/Al_2O_3

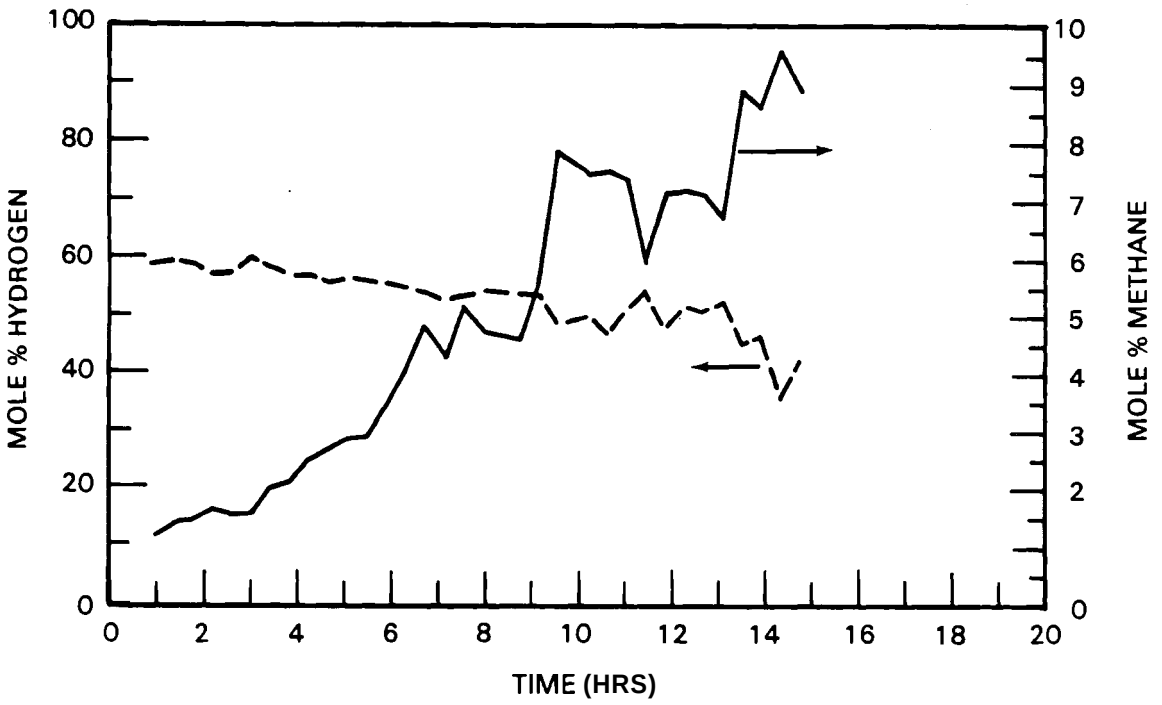

FIGURE 11. Product Gas Composition in Runs C1 and C2 Indicating Catalyst Deactivation

spheres. In Runs C-1 and C-2 we used a reduced trimetallic NiCuMo catalyst. Both of these catalysts are in the 40-70 mesh size range and are fluidizable at about 0.3 m/s superficial gas velocity.

A summary of the results of these runs is presented in Table 8. Additional information is located in Appendix A where complete mass balance calculations and computerized run summaries can be found. The following is a brief description of each run.

Run B-1

A Ni/Al_2O_3 hydrocracking catalyst was used in what best can be described as a "shakedown run". The primary purpose of Run B-1 was to provide a feel for how bagasse would behave in the PDU system. This run was very similar to those with wood as a feedstock. Steady state was achieved for approximately 3.5 hrs before an orderly shutdown. No problems were encountered, although there was some indication of catalyst deactivation near the end of the run.

Run C-1

After Run B-1, preparations were made for a continuous week-long-run. Our initial intention was to test a variety of operating conditions. The first conditions tested were steam/bagasse ratio of 1.0 a maximum reaction zone temperature (up to 780°C).

Steady state was reached and sustained for approximately eight hours. During this run the gas compositions continually deteriorated (see Figure 11). Methane concentration went from less than 1% to almost 5%, a sign of catalyst deactivation.

With a steam/bagasse ratio of 1.16 and a temperature around 775°C, the typical gas had a 2.6 H_2/CO ratio. This indicated a lower steam rate would be required to give a H_2/CO ratio about 2 which is considered optimum for methanol synthesis. Yields were very good at 1.54 m^3/kg (22.7 scf/lb) and are as good or better than with wood feedstocks.

TABLE 8. **Summary** of PDU Results with Nickel **Catalysts**

Run Number	B-1	C-1	C-2	B-2a	B-2b	Davy Base
Catalyst	Ni on Al_2O_3	Ni-Cu-Mo	Ni-Cu-Mo	Ni on Al_2O_3	Ni on Al_2O_3	Case (Wood)
Bed Temp, (°C)	740	775	780	740	740	750
Feed Rate (kg/hr dry bagasse)	14.0	10.2	12.0	14.5	14.5	
Steam Rate (kg/kg dry bagasse)	1.10	1.16	1.12	0.87	1.02	0.75
Gas Composition						
H_2	.566	.560	.512	.561	.568	.513
CO_2	.242	.198	.223	.218	.203	.174
CH_4	.041	.029	.072	.038	.021	.056
CO	.151	.210	.179	.183	.205	.257
C_2^+	0	.004	.14	0	.003	0
Dry nm^3/kg Dry Bagasse	1.66	1.54	1.04	1.90	1.82	1.57
Carbon Conversion (wt%)						
to gas	82.5	78.6	59.8	95.8	90.9	80
to char	12.9	8.0	9.5	6.8	12.9	20
to liquids	.1	0.1	1.5	.05	.05	0
Carbon Balance, %	95.3	86.6	70.8	102.6	103.8	100
Total Bagasse Fed (kg)	50	75	78	30	59	

(a) The results are averaged over the entire steady state operating period. For Runs C-1 and C-2 the gas compositions deteriorated steadily with time as shown in Figure 11. For Runs B-1, B-2a, and 8-Zb the gas compositions were fairly constant.

Run C-2

After Run C-1, the filter vessel was drained. Only 0.59 Kg (1.3 lbs) of char was removed. Blowback pulses of nitrogen were ineffective in dislodging any further char so Run C-2 was started.

Run C-2 was at steady state approximately 6.5 hrs. Reaction temperatures were again high (780°C) due to the low feed rate 11.8 kg/hr (26 lbs/hr). The Steam/bagasse ratio was 1.1.

During both C-1 and C-2 there was a continuous increase in the pressure drop across the system. The maximum pressure was approximately 240 kPa (20 psig). There are several possible causes for pressure buildup:

- char overload in the filter vessel
- tar deposition on the demister element
- tar deposition in the off gas line
- water vapor freezing in off gas line (ambient temperature -5°C)

11 kg (24 1bs) of compacted char was recovered from the filter vessel after Runs C-1 and C-2. We also found a large plug of ice downstream of the pressure letdown valve. Both char buildup and freezing contributed to the high pressure in the reactor.

Run C-2 was finally terminated after efforts to reduce the system pressure proved futile and gas compositions indicated the catalyst had become deactivated.

Runs B-2a,b

Since the Ni/Cu/Mo on Al_2O_3 catalyst used in runs C-1 and C-2 had deactivated so rapidly, we used Ni on Al_2O_3 in run R-2 to see whether it also deactivated. A mixture of two sizes of catalyst were used. A mixture of 19 kg of 40-70 mesh and 12 kg of 20-40 mesh constituted the required .025 m^3 (0.9 ft^3) of bed material.

The initial steam/bagasse ratio was .87. Increased gas flow needed far good fluidization forced the steam rate up after 2 hrs of steady state. This

second set of conditions is referred to as B-2b. Methane concentrations were down at 3% in B-2a with a H_2/CO ratio of just over 3. Gas production was very good at 1.76 nm^3/kg (26 scf/lb) bagasse.

In B-2b the steam/bagasse ratio was boosted to 1.02 to help fluidization. This appeared to help for a short while, then the fluidization problems reoccurred. The gas composition was good as methane was down to ~2% with these new conditions. The H_2/CO curiously went down to 2.7. Steady state was maintained for four hours before a voluntary shutdown was initiated. There was no positive indication of loss of catalyst activity.

Results and Discussion

The PDU results with nickel catalysts showed bagasse pellets to be a good feedstock for fluid-bed gasification. With an active nickel catalyst the results were as good or better than those achieved with wood. Particularly encouraging was the high carbon conversion to gas, about 90% on the average for five runs. Carbon conversion to liquid products was quite low, 0.1% or less when the catalysts were active. Unfortunately the NiCuMo catalyst quickly lost its activity and the yield of tars, oils, and light hydrocarbon gases (CH_4, C_2H_4, C_2H_6, etc) increased rapidly. The amount of the H_2 and CO produced was reduced accordingly. The Ni/Al_2O_3 maintained its activity longer in the PDU but laboratory tests (Run 24) showed that it also deactivated too fast to be considered for further tests.

Laboratory studies indicated the loss of activity was due to some combination of sulfur poisoning and carbon deposition. A brief catalyst screening study indicated alkali carbonate catalysts doped on the bagasse would be an effective catalyst system for bagasse and the rest of the PDU tests were made with this type of catalyst system.

PDU GASIFICATION TESTS WITH POTASSIUM CARBONATE CATALYST

Based on the results of laboratory testing a catalyst system of 10 wt% K_2CO_3 impregnated on the bagasse was selected for testing in the PDU. This catalyst had previously been successfully tested with wood (Mudge 1983). Use

of this catalyst in the PDU presented several technical problems compared to using supported catalysts in the bed including:

- preparing the feedstock,
- preventing agglomeration in the gasifier, and
- finding a suitable inert material for the fluid bed.

All of these problems were solved in the course of the PDU tests with K_2CO_3 and the results of these tests were quite encouraging.

Feed Preparation

In previous tests with wood the K_2CO_3 was dry mixed with wet wood chips. This worked quite well as the moisture in the wood absorbed the K_2CO_3. The chips were then dried prior to gasification. The bagasse pellets were too dry to use this method so a solution of K_2CO_3 had to be used. We were afraid that drying would result in pellet attrition so we had to use a minimum of water.

A 40 wt% solution of K_2CO_3 was mixed with the bagasse in a preliminary PDU impregnation test. Laboratory tests with this feed material (Runs 21 and 22) were good. A cement mixer and a spray wand were then used to impregnate about one ton of pellets with a 40 wt% solution of K_2CO_3. This resulted in a fairly even distribution on each pellet and a minimum of pellet attrition. The final composition of the pellets on a wet basis was about 9% K_2CO_3, 78% bagasse fibre, and 13% moisture. The impregnated bagasse is shown in Figure 9.

Run Summaries - Potassium Carbonate Catalysts

Four PDU runs were made with K_2CO_3 impregnated bagasse at four different bagasse feed rates which corresponded to four different steam/bagasse ratios. The steam rate was held constant at the rate required to fluidize the bed. The length of the runs was limited by accumulation of residue (char, ash, catalyst) in the filter vessel. The residue from these runs is finer and is produced at a higher rate (due to the K_2CO_3) than we have experienced in the past. This puts a heavier load on the filter vessel. The filter vessel is not designed to be emptied during a run. When we attempted to empty the filter vessel "while it was hot" the char ignited and caused a small fire.

Deactivated $NiCuMo/SiO_2-Al_2O_3$ catalyst was used as an inert bed in Runs C-3a, C-3b, and C-4. A laboratory test with this material confirmed that

it had no catalytic activity. In Run C-5 we used sintered bauxite obtained from the Norton Co. for the inert bed. It is a low surface area, high alumina material, 0.2 to 0.4 mm in size with a density of 2.2 g/cm^3.

A summary of the results of each run is given in Table 9. Detailed run summary sheets are in Appendix A. Following is a brief summary of each run.

Run C-3a

This was probably the smoothest run we have ever had in the PDU. Steady state was achieved almost immediately and continued for 8 complete hours.

TABLE 9. Summary of PDU Results with 10% K_2CO_3 Impregnated Australian Bagasse Pellets(a)

Run Number	C-3a	C-3b	C-4	C-5
Bed Temp., (°C)	760	760	730	720
Feed Rate (kg/hr dry bagasse + K_2CO_3)	14.3	19.6	24.5	30.6
Steam Rate (kg/kg dry bagasse)	0.95	0.70	0.62	0.58
Gas Compositions (mole %)				
H_2	54.5	55.3	51.4	46.8
CO_2	24.0	24.5	24.2	24.8
CH_4	3.6	4.0	3.0	3.9
CO	11.0	15.1	20.2	22.6
C_2^+	0.9	1.1	1.2	1.9
Dry nm^3 gas/kg dry bagasse	1.47	1.47	1.44	1.14
Carbon Conversion (wt%)				
Gas	83.0	81.7	87.4	76.5
Solid.char	22.0	21.4	18.3	23.1
Liquids	0.19	0.05	0.57	0.74
Total	105.3	103.1	106.2	100.2
Total Bagasse Fed (kg)	96.5	46.2	102.3	91.8

(a) The results are averaged over the entire run.

Temperatures in the bed did not vary more than a few degrees from 760°C as shown in Figure 12. This indicates good fluidization was achieved and no agglomeration or clinkering was occuring. The results of the run were somewhat better than laboratory tests and nearly as good as achieved with active nickel catalysts. Gas compositions were nearly constant throughout the run.

Run C-3b

This run is a direct continuation of C-3a. The only difference is the feed rate was increased from 14.3 kg/hr(31.5 1b/hr) to 19.6 kg/hr(43.2 1b/hr). This is the weight of bagasse plus K_2CO_3 catalyst. Very little change can be seen in the gas compositions between runs. The product flow rate did rise from .337 nm^3/min(11.9 scfm) to .447 nm^3/min(15.95 scfm). This is a 33% rise in gas production for a 37% increase in feedrate.

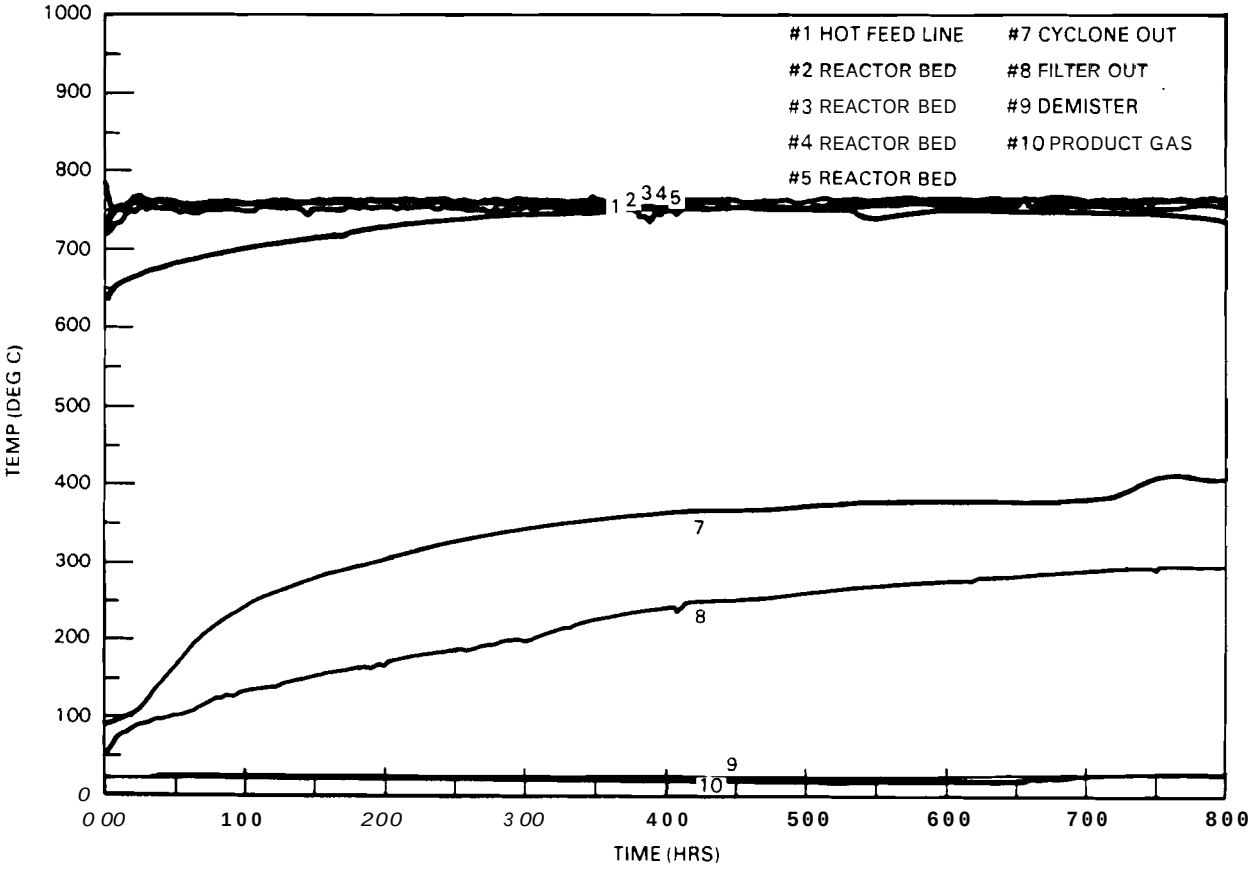

FIGURE 12. PDU System Temperatures - Run C-3a

This test was terminated when pressure drop across the filter vessel became excessive. The maximum pressure reached in the reactor was just over 239 kPa(20 psig). The plant was "idled" and the filter vessel was emptied. A small fire resulted when hot, fine char particles contacted the atmosphere. We had previously noticed that char removed from the cyclone during the run would burn slowly in the buckets used to collect it. This was a continual problem throughout the PDU runs with K_2CO_3 catalyst. After C-3b we decided that the filter vessel could not be safely emptied while it was hot and the plant would have to be shutdown each time the filter vessel had to be emptied.

Run C-4

After the filter vessel was emptied from Runs C-3a and C-3b we increased the feedrate to 24.5 kg/hr (54 lb/hr). The steam/bagasse ratio dropped to 0.62 and the product flowrate rose further to .55 m^3/min (19.5 scfm). Gas compositions changed slightly. The hydrogen/CO ratio predictably dropped from 3.5 to 2.5.

Toward the end of the run we experienced fluidization problems as the system pressure rose. Successive circuit breaker trips caused large fluctuations in the reactor core temperature (Figure 13). Immediately following these fluctuations a layer of nonfluidizable material formed just over the distributor plate. Thermocouple number 2 (Figure 13) clearly shows the thermocline.

After Run C-4 the reactor and off-gas system were dismantled and cleaned. There was a large accumulation of caked potassium carbonate and bed material in the bottom of the reactor. Almost all of the holes in the distributor plate were plugged. With the poor fluidization and low temperature in the bottom of the bed that were noted toward the end of C-4, we anticipated something like this when we opened up the vessel.

Run C-5

The three primary objectives of Run C-5 were: 1) operate without agglomeration or clinkering, 2) increase the bagasse/steam ratio to produce a gas with a H_2/CO ratio of 2:1, and 3) test the new sintered bauxite inert bed material.

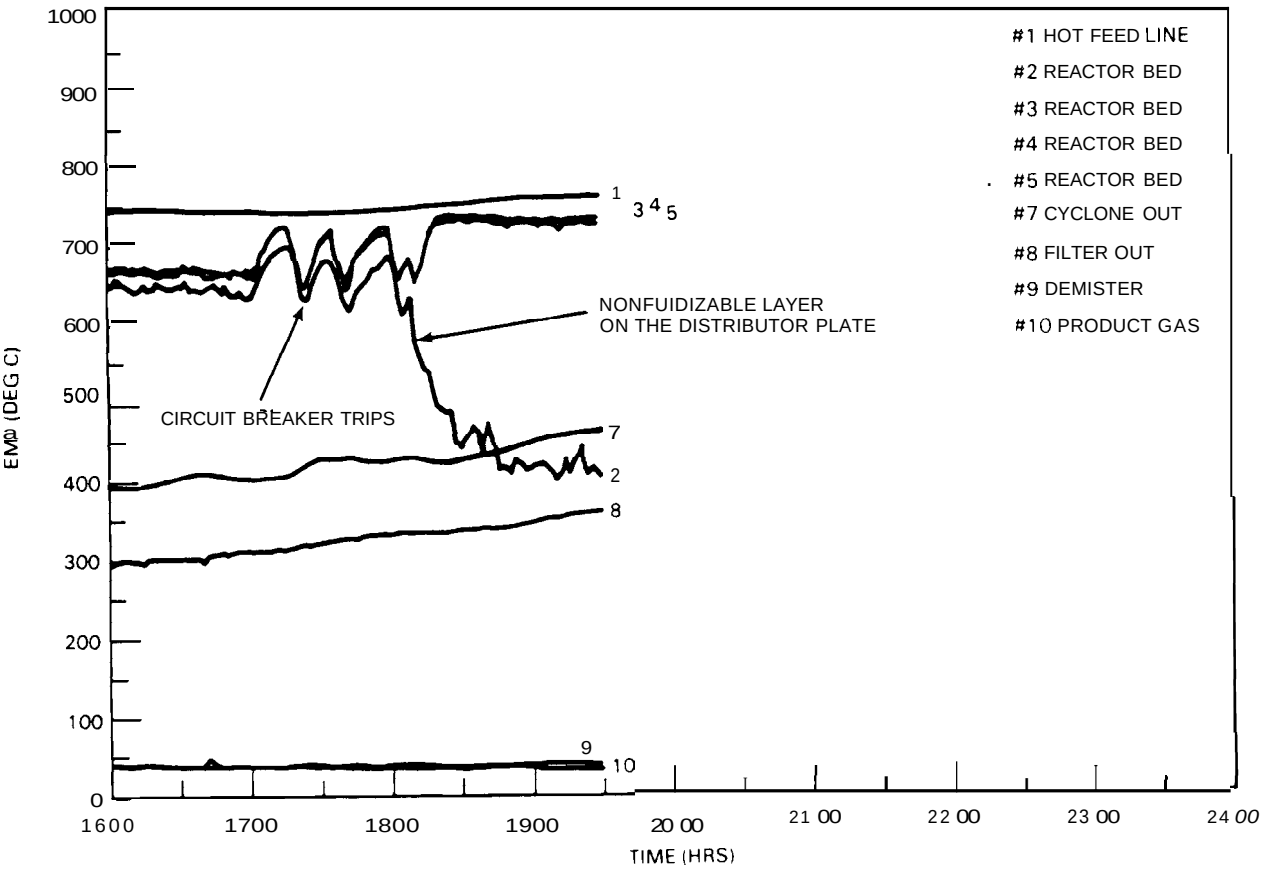

FIGURE 13. PDU System Temperatures - Run C-4

Run C-5 was a very steady run. The feed rate was over 75 lb/hr (34.1 kg/hr) and was the highest sustained rate that we have ever run in this PDU. Fluidization was poor at first but improved as feed was introduced to the bed. The fine char may act like a lubricant to improve fluidization,

Temperatures in the reactor averaged about 720°C during the time when bagasse was being fed. The steam/bagasse ratio was 0.58 and the H_2/CO ratio turned out to be 2.1. The material balance was quite good for this run (99%). Gas rates (.51-.57 m^3/min) and carbon conversion to gas were less than anticipated based on Runs C-3 and C-4. We attribute this to the increased gas velocity in the bed necessary to fluidize the new bed material. This reduces the char residence time, and thus reduces carbon conversion to gas. The temperature was somewhat lower due to the higher flow rates. This lower temperature also reduces carbon conversion.

During this run there was no sign of clinker formation. Examination of the bed material when it was removed showed no large clinkers. Over 99% of the potassium impregnated on the bagasse was accounted for in the cyclone solids (54%), the filter solids (17%) and in the bed material (28%).

Table 10 shows screen analysis of the bed material after Run C-5. There was a small increase in the particle size distribution but of the larger particles almost all were pellet skeletons, i.e. unreacted bagasse, and not clinkers. Very little attrition was noticeable indicating sintered bauxite is a good material for fluid-bed applications.

Results and Discussion

The tests with K_2CO_3 impregnated bagasse in the PDU went quite well and showed the feasibility of using this type of catalyst system in a larger scale. The results were as good or better than laboratory gasifier results with the same catalyst and nearly as good as laboratory and PDU results with active nickel catalysts.

Carbon conversion to gas was good (77-87%) although not as high as achieved with nickel catalysts. This is consistent with results with wood in the PDU (Mudge 1983). Although K_2CO_3 is an effective catalyst for the

TABLE 10. Particle Size Distribution of Bed Material After Run C-5[a]

US Screen Size	Sieve Opening mm	% Retained
8	2.38	1.3
16	1.19	1.3
20	.841	0.5
40	.420	14.2
70	.210	80.7
100	.105	1.0
pan	--	1.0

(a) Starting material was 90% minimum between 40 and 70 mesh.

steam/char reaction (which should increase carbon conversion to gas), at short residence times (less than 10 minutes) laboratory tests show an increase in char production with alkali carbonate catalysts (Sealock, 1982; Mudge 1981). This char is more reactive and will gasify faster than uncatalyzed char if sufficient time is provided. The char residence time in the gasifier is difficult to determine but the fact that carbon conversion is lower with K_2CO_3 indicates it probably is fairly short.

Carbon conversion appears to be more a function of gas velocity in the gasifier bed. Increasing the gas velocity as was done in Run C-5 increases the size of char particles which are carried out of the bed. This reduces their residence time and reduces carbon conversion.

A steam rate of about 0.5 kg/kg dry bagasse is required to produce a gas with a H_2/CO ratio of 2. Alkali carbonates appear to be a better shift catalyst than nickel catalysts so a lower steam rate can be used.

Clinker formation can be a problem with alkali carbonate catalysts; however, if good fluidization is maintained in the bed through the run and during the cooldown phase of shutting down, clinker formation can be avoided. The melting point of K_2CO_3 is 891°C and the K_2CO_3 is quite mobile at 750°C. Particles of K_2CO_3 agglomerate and gradually get larger until chunks like those found at the end of Run C-4 are formed. Maintaining good fluidization prevents agglomeration and minimizes hot spots where agglomeration would be more severe.

Prior to Run C-4 the steam flow was stopped but the reactor kept hot when the filter vessel was emptied. We believe the clinkers found after Run C-4 were formed primarily during this time period. During Run C-5 good fluidization was maintained throughout the run. We were careful to maintain fluidization after shutdown until the reactor had cooled to below 500°C. No clinker formation was observed.

PROCESS EVALUATION

Based on the results of the laboratory and PDU gasification tests a design basis for economic calculations was developed. A process flow diagram was developed, and heat and material balances were performed to determine the ultimate yield of methanol from bagasse. This was combined with cost information, supplied primarily by Davy McKee Engineers and Constructors for wood based plants, to calculate the selling price of methanol. The effect of variables such as plant size, capital cost, financing method, and bagasse cost on the final cost of methanol were evaluated.

PROCESS DEVELOPMENT

The process developed to convert bagasse to methanol is based on:

- PDU and laboratory investigations of catalytic gasification of bagasse for the production of methanol synthesis gas, and
- a detailed feasibility study of methanol production via catalytic gasification of wood done by Davy McKee Engineers and Constructors for PNL (Mudge 1981).

The design basis is shown in Table 11. The gasifier operates at 750°C and 1000 kPa (150 psia) and 90% of the carbon in the bagasse is converted to gas. Operation of the gasifier at 1000 KPa significantly reduces downstream compression costs. In PDU tests with wood carbon conversion and gas yields were slightly higher at 1000 kPa compared to atmospheric tests. Gas compositions were similar when a catalyst was used (Mudge 1983). Carbon conversion to gas ranged from 77-87% for the bagasse PDU tests. With a properly designed system operating at 1000 kPa (150 psia) we believe 90% carbon conversion to gas can be achieved with bagasse and this was used for the design basis.

A heat and material balance for the process is shown in Table 12. The yield, of methanol is 0.47 kg/kg dry bagasse (0.49 kg/kg MAF bagasse), and the overall thermal efficiency is 56%.

TABLE 11. Design Basis for Conversion of Bagasse to Methanol

Plant Capacity- 800 ton/day (727 t) dry bagasse
Location - U.S. Gulf Coast
Storage Required - 8 months feed
Operating Factor - 330 days/year

Feedstock - Bagasse

Ultimate Analysis (wt%) dry basis	
Moisture	50
C	44
H	6
O	44
N	1
S (ppm)	300
Ash	5
Total	100
MJ/kg	9.11

Gasifier Operation

Pressure - 1000 kPa (10 atm)
Temperature - 750°C
Steam Rate - 0.4 kg/kg dry bagasse (assumes ~10% moisture in bagasse feed
Catalyst - 10 wt% K_2CO_3 (0.11 kg/kg dry bagasse)

Gas Production (dry basis) - 1.4 nm^3/kg dry bagasse
Gas Composition - (vol %)

H_2	49.5
CO	24.5
CO_2	19.0
CH_4	5.0
C_2^+	2.2
N_2	0.5
H_2S (ppm)	13.0

Residue Production - 0.2 kg/kg dry bagasse (90% carbon conversion to gas)

Catalyst Recovery - 80% with 2ℓ H_2O/kg residue

ΔH_{rxn} = 308 kcal/kg dry bagasse

Total Heat Input Required to the Gasifier - 787 kcal/kg dry bagasse

TABLE 12. Heat and Material Balance for 800 ton (727 t)/day Plant

In	MJ/kg	kg/hr	MJ/hr x 10^3
Bagasse	9.11	60,600	552
Water for Steam Generation	-	11,700	-
Catalyst K_2CO_3	-	530	-
Air	-	362,200	-
Electricity (8756 KW)	-	-	30
Diesel Fuel (21.2 ℓ)	-	-	1
Total		435,000	583

Out	MJ/kg	kg/hr	MJ/hr x 10^3
Methanol	22.8	14,300	326
Char/Ash Residue	8.2	3,800	25
Acid Gas	-	10,100	
Process Waste Water	-	11,700	
Vent Gases (Dryer, Boiler, Reformer Gasifier Heater)	0.262	396,100	105
Cooling Tower, Mechanical, and other Losses	-	-	127
Total		435,000	583

Methanol Yield = 0.47 kg/kg dry bagasse (.49 kg/kg MAF bagasse)

$$\text{Thermal Efficiency} = 100 \times \frac{\text{methanol, HHV}}{\text{Bagasse, HHV} + \text{Electricity} + \text{Diesel}} = 56\%$$

Figure 14 is a process flow diagram for converting bagasse to methanol. The flow rate and composition of the major streams are detailed in Table 13 for a plant feeding 800 ton (727 t)/day of dry bagasse. Following is a more detailed description of the individual sections of the plant.

Bagasse Storage and Preparation

Storage for an eight month supply of feedstock for the gasifier is provided. This makes it possible to operate the plant on a year round basis. Nylon tarpaulins are used to keep the bagasse dry during the wet season.

The bagasse is conveyed to the plant and about 20% is split off to fuel the bagasse dryer. The remainder is hammermilled to smaller than 1 cm and then mixed with a solution of potassium carbonate. The impregnated feedstock is then dried to about 10 wt% moisture and conveyed to the gasifier feed bins. We

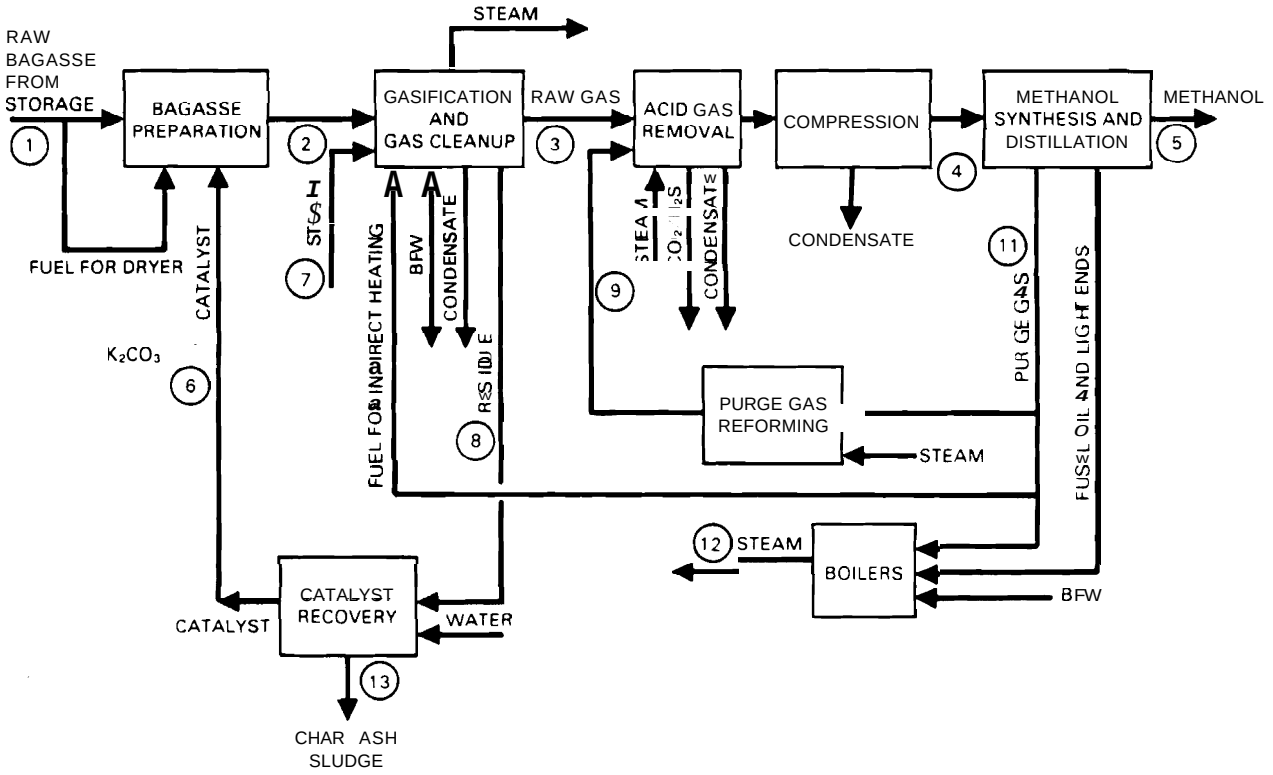

FIGURE 14. Process Flow Diagram for Converting Bagasse to Methanol

investigated the use of pelletized feed, but cost information from Davies Hamakua Sugar indicated pelletizing would be quite expensive. Indications from the PDU were that the major advantage of the pelletized feed was the reduced length of the fibers, so size reduction is all that is included in the process design.

Gasification and Gas Cleanup

Bagasse from the feed bins is continuously fed by screw conveyors to the fluidized bed gasifiers. For 800 ton (727 t)/day of dry bagasse, two gasifiers with inside diameters of 3 m would be required. This corresponds to a processing rate of about 1630 kg dry bagasse/m^2/hr. The fluid bed in the gasifiers is an inert material in the size range of 0.2 to 0.8 mm and is about 3 m deep. The gasifiers operate at 1000 kPa (150 psia) and 750°C.

Purge gas is burned in a heater to supply the necessary heat input to the gasifier, by circulating hot combustion gases through tube bundles in each gasifier bed.

TABLE 13. Flow Rate and Composition for Major Streams

Solids

Stream No. / Name	1 Raw Bagasse	2 Bagasse to Gasifier	8 Gasifier Residue	6 Catalyst Make Up	13 Char/Ash Sludge
Composition, kg/hr					
Carbon	13.330	10,540	1,050		1,050
Hydrogen	1,820	1,430			
Oxygen	13.330	10,540			
Nitrogen	290	240			
Sulfur	10	8			
Ash	1,520	1,190	1,190		1,190
Water	30,300	2,650			
K_2CO_3	-	2,650	2,650	530	530
Total	60,600	29,300	4,890	530	2,770

Gases (dry basis)

Stream No. / Name	3 Raw Synthesis Gas	9 Reformed Purge Gas	4 Synthesis Gas to Methanol	11 Purge Gas	10 CO_2, H_2S
Flow Rate, $nM^3/hr \times 10^3$	33.5	13.9	42.2	9.0	5.1
Composition, vol %					
H_2	49.0	74.0	63.6	47.2	
CO	24.5	16.8	25.1	18.7	
CO_2	19.0	9.2	5.8	4.2	99.9
CH_4	5.0		3.9	17.9	
C_2^+	2.2			8.0	
N_2	0.5			4.0	
	130 ppm				0.1
Water, vol % (wet basis)	10.0	49.7	0.1	-	11.3

Others

Strew No. / Name	5 Methanol	7 Steam to Gasifier	12 Steam from Boilers
Composition, kg/hr			
Water		9,590	11,700
Methanol	14.340		

Raw synthesis gas leaves the gasifier at 750°C and is cooled in a waste heat boiler. It then passes through a cyclone and a bag filter arrangement to remove the char/ash/catalyst residue. It is then further cooled in a second waste heat boiler and a water cooler.

The char/ash/catalyst residue is washed with water to recover the K_2CO_3 catalyst. The catalyst solution is recycled and the remaining residue disposed.

Acid Gas Removal

The synthesis gas from the gasifier combines with reformed purge gas and enters the Benfield process to remove the H_2S and some of the CO_2 from the synthesis gas. The process uses an aqueous solution of potassium carbonate (K_2CO_3) to absorb the acid gases. The acid gases are stripped from the solution with steam and the K_2CO_3 solution is recycled. The process design for the wood based plant called for venting the CO_2. This may be environmentally unacceptable because of the sulfur content of bagasse, and other disposal methods may have to be considered.

Compression

Synthesis gas is compressed to 7022 kPa (1000 psig) using a multistage centrifugal booster compressor. Following the compressor the gas passes through a chloride guard bed, a sulfur guard bed, and an ethylene hydrogenation bed. These are all catalyst poisons for the methanol catalyst and must be removed or, in the case of ethylene, converted to ethane.

Methanol Synthesis and Distillation

The ICI low pressure methanol process is used to convert the synthesis gas to methanol. It operates at 7022 kPa (1000 psig) and uses a proprietary copper based catalyst. The reaction to produce methanol are represented by the following equations:

$$2H_2 + CO \rightleftharpoons CH_3OH$$
$$3H_2 + CO_2 \rightleftharpoons CH_3OH + H_2O$$

The proper synthesis gas composition for the ICI process is 2.5 H_2/CO and 4.3 CO/CO_2. Other components such as CH_4, C_2H_6, and N_2 are inert.

The synthesis loop consists of the following items: circulator (centrifugal compressor), converter, heat exchanger, heat recovery exchanger, cooler, and separator. The methanol and water vapor produced are condensed and separated. The unconverted synthesis gas is recycled and a portion of the recycle is purged to remove inerts. The crude methanol is distilled in one column to produce fuel grade methanol.

Purge Gas Reforming

Part of the purge gas from the methanol synthesis unit is used as fuel, the remainder is reacted with steam in a reforming furnace to produce a hydrogen rich gas. The reforming furnace consists of alloy tubes packed with nickel catalyst. A portion of the purge gas is used to fire the furnace.

Boilers and Other Utilities

Steam for the process is generated by a package boiler at 4255 kPa (600 psig) and 400°C. It is fired with purge gas and fusel oils from the methanol synthesis unit. Steam is also generated in waste heat boilers in the gasification section.

All liquid wastewater streams in the plant are treated in a neutralizing basin and a subsequent three stage biological treatment system before being released to drainage.

Also included in the design are an instrument air system, an inert gas system, and a flare system.

COST STUDIES

Evaluating the economics of producing methanol from bagasse was accomplished by adapting previous studies utilizing wood as a feedstock (Mudge et al., 1981, 1983). We have computerized the entire economic analysis onto a large spreadsheet type program. Many of the process variables (bagasse cost, plant size, yields, labor costs, etc.) can be changed and a complete recalculation using these new values can be done rapidly. The simulation predicts the final selling price of methanol from bagasse required to make the process economically attractive.

Methodology

The base capital and operating cost data were developed for PNL's Catalytic Gasification process in 1980 by Davy McKee Engineers and Constructors (now DM International) and reported as appendicies to Mudge et al. (1981). They determined the final selling price for methanol based on a method outlined in Coal Gasification - Gas Cost Guidelines (Skamser 1978) which is the standard cost estimating method used by the U.S. Department of Energy. Davy costed plants feeding 2000 tons/day (1818 t) and 200 tons/day (182 t) of wood. We incorporated this information into a spreadsheet type program (VisiCalc$_{TM}$) which allowed us to calculate the effect of key variables and to determine the costs for plant sizes in between 200 and 2000 tons/day using appropriate scaling factors. The costs can be updated by inputting current cost indices.

The process design for converting bagasse to methanol varies somewhat from the wood based process, so a new basis was developed. The differences are primarily in storage, feed preparation, and catalyst recovery. For storage and feed preparation, cost information obtained from Sugar Research Institute and equipment vendors was combined with the information on wood developed by Davy. Catalyst recovery costs were based on a cost estimate by Exxon for their catalytic coal gasification process which uses K_2CO_3 (Fant 1980; Furlong 1978). The rest of the equipment from the gasifier downstream to and including the methanol synthesis unit was costed based on the Davy study. Relative sizes of the equipment were adjusted to account for the small differences between bagasse and wood (e.g. less steam is required for bagasse).

Economic Evaluation

Once the plant capital requirements and operating costs have been determined the required methanol selling price can be calculated using the method outlined in Skamser (1978) for utility and private financing. The following economic basis was used:

Utility Financing

- project life - 20 years
- depreciation method - 5% straight line
- debt/equity ratio - 75/25
- income tax rate - 48%

- interest - 10%
  return on equity - 15%

Equity Financing

- project life - 20 years
- depreciation method - 16 yr sum of the digits
- equity capital - 100%
- rate of return - 12% DCF
- income tax rate - 48%

Based on methods presented by Skamser (1978) and used by Mudge et al (1981, 1983), this information was used to develop equations relating net methanol cost in US $/gal to the following:

- total plant investment
- start up costs
- working capital
- net operating costs
- annual MeOH production
- annual MeOH revenue.

These equations, were then input to the spreadsheet program to perform the calculations. A sample computer output and the economic analysis is included in Appendix B.

There are fourteen input cost variables which are easily altered at each discount rate. They are:

- plant capacity (dry short tons)
- Chemical Engineering plant cost index
- chemical engineering construction index (labor)
- bagasse cost (US $/dry ton)
- water cost (US $/M gal)
- electricity cost (US $/kwhr)
- diesel fuel (US $/gal)
- gasifier catalyst (US $/lb)
  shift catalyst (US $/$ft^3$)
  chloride guard catalyst (US $/$ft^3$)
- sulfur guard catalyst (US $/$ft^3$)

- methanol catalyst (US $/lb)
- reformer catalyst (US $/ft^3)
- process labor (US $/hr).

For this study we have varied only plant capacity and feedstock costs.

In the United States most of the sugar cane is grown in Hawaii, Florida, Louisiana, and Puerto Rico in that order. The average size suger mill produces anywhere from 50 to 800 ton/day of dry fiber (Boyd 1980). Four cases, representative of the bagasse supply, have been evaluated to determine the price of methanol necessary to make this process competative. Plant capacities range from 200 to 800 ton/day of dry fibre. Bagasse costs were varied between 10 and 40 $/ton. Table 14 shows the total installed capital cost and the calculated required methanol selling price for each case considered. These results are also shown in graphical form in Appendix B (Figures R.l and B.2).

The capital cost estimates are based on Newport, Oregon location (the site for the wood based plants). Location factors we have found (Guthrie 1974) indicate Gulf coast USA installed costs would be about 6% higher. Given the original estimate was only accurate to ±25% making such small corrections may not be needed. Each % error reflects a corresponding $.01/gal rise or fall in the required price of methanol.

Table 15 shows a breakdown of the capital cost for a plant feeding 800 ton day of dry bagasse. Bagasse storage, gasification, and methanol synthesis are the three largest cost areas. Capital costs are the most significant contributor to the final methanol selling price. The capital costs are higher for bagasse than wood primarily due to the large amount of storage required (8 months) and additional costs for catalyst recovery because the K_2CO_3 catalyst was used.

TABLE 14. Summary of Economic Evaluation

Case	Plant Size Ton/Day	Capital Cost $ million	Bagasse Cost $/dry ton	Methanol Cost	
				Utility	Private
1	800	97.5	10	0.91	1.22
2	800	97.5	40	1.13	1.44
3	200	32.1	10	1.30	1.75
4	200	32.1	40	1.52	1.97

TABLE 15. Total Capital Required for a Plant Feeding 800 ton (727 t)/day Dry Bagasse Fibre (dry basis)

Cost Component	US $ Millions
Bagasse Storage and Preparation	20.1
Bagasse Drying	5.3
Gasification and Gas Cleanup	14.1
Shift Conversion	0.6
Acid Gas Removal	3.5
Compression	3.5
Methanol Synthesis and Distillation	9.9
Purge Gas Reforming	0.5
Utilities, Offsites, Miscellaneous	8.5
Direct Equipment Costs (DEC)	66.1
Field Indirects	13.9
Professional Services	15.9
Other	1.6
Total Installed Cost (TIC)	97.5
Funds During Construction	11.0
Start-Up Costs	4.0
Working Capital	2.4
Total Capital Required	114.9

The rest of the capital costs are actually somewhat less for bagasse. The large storage costs will be a problem for any year-round process using bagasse as a feedstock.

Table 15 lists the operating costs for the same plant. Operating costs other than the cost of bagasse are based on the original Davy Study with wood. Labor costs are the largest single operating costs. If the cost of bagasse exceeded $20/ton it would become the dominant operating cost. As shown in Table 14 the plant size also has a significant effect on the final cost of methanol. This results from economies of scale primarily for the downstream equipment after the gasifier, particularly compression and methanol synthesis.

TABLE 16. Projected Annual Operating Costs for a Plant Feeding 800 ton (727 t)/day Dry Bagasse Fibre

Cost Component	US $ 1000/yr
Bagasse @ $ 10/dry ton	2,640
Gasification Catalyst	1,824
Other Catalysts and Chemicals	437
Utilities	2,261
Labor	4,879
Administration and General Overhead	2,928
Supplies	2,354
Taxes and Insurance	2,700
Total Gross Operating Costs	20,026

A central location receiving bagasse from several mills is more economic than a small plant at a single mill even if transportation costs significantly increase the cost of bagasse.

The calculated cost of methanol produced from bagasse is significantly higher than the current cost of methanol. A previous economic evaluation with wood showed the required selling price of methanol was competitive with the market price of methanol (Mudge 1981). The following factors make this latest analysis less favorable:

- Capital costs are higher
  - Construction costs have inflated 20% since 1980
  - Capital costs for bagasse are somewhat higher than for a wood based plant,
- The previous evaluation used a larger plant size than is feasible with bagasse which resulted in significant economies of scale.
- The market price of methanol in the U.S. has dropped 44% in the last six months to $0.45/gal ($0.12/ℓ). (Current as of April 1984)

Production of methanol from bagasse may still be profitable in the long term if a shortage of transportation fuels occurs and prices increase

significantly. The cost of bagasse is primarily a function of capital costs and so the future price will be determined primarily by the inflation of the cost of materials and labor for construction.

REFERENCES

Baker, E. G., L. K. Mudge, and W. A. Wilcox. 1982. "Catalysis of Gas Phase Reactions in Steam Gasification of Biomass." Presented at the Fundamentals of Thermochemical Biomass Conversion: An International Conference, Estes Park, Colorado, October 18–22, 1982.

Boyd, S. S. 1980. The Gilmore Sugar Manual. Sugar Publications, Fargo, North Dakota.

Bridgewater, A. V. 1979. "International Construction Cost Location Factors." Chemical Engineering, November 5, pp. 119–121.

Cox, J. L. and L. J. Sealock, Jr. 1974. Prepr. Am. Chem. Soc., Div. Fuel Chem. 19 (1):64-78.

Fant, B. T. and C. A. Euker, Jr. 1980. "Exxon's Catalytic Coal Gasification Process." Proceedings of the First International Gas Research Conference, Gas Research Institute, Chicago, Illinois.

Furlong, L. E. and N. C. Nahas. 1978. "Catalytic Coal Gasification-Process Research and Development." Proceedings of the Tenth Synthetic Pipeline Gas Symposium. Gas Research Institute, Chicago, Illinois.

Guthrie, K. M. 1974. Process Plant Cost Estimating Evaluation and Control. Craftsman Book Company of America, Solana Beach, California.

Hawley, M. C., et al. 1983. "Gasification of Wood Char and Effects of Intra-particle Transport." Fuel. 62 (2):213–216.

Houghtby, W. E., et al. 1982. Sulfur-Tolerant Fuel Processors for Fuel Cell Power Plants, EPRI EM-2686. United Technologies Corporation, South Windsor, Connecticut.

Johnson, J. L. 1981. "Fundamentals of Coal Gasification." In Chemistry of Coal Utilization, pp. 1491–1598. John Wiley & Sons, New York.

McCarty, J. G., et al. 1981. Hydrocarbon Reforming for Hydrogen Fuel Cells. DOE/MC/11323-T2. SRI International, Menlo Park, California.

McCoy, L. R. 1983. Investisations of Coal Gasification Catalysis Reaction Mechanisms. DOE/MC/14592-1419. Energy Systems Research Group, Albany, New York, and Rockwell International Corp., Thousand Oaks, California.

McKee, D. W., et al. 1983. "Catalysis of Coal Char Gasification by Alkali Metal Salts." Fuel. 62 (2):217–220.

Mudge, L. K., L. J. Sealock, Jr., and S. L. Weber. 1979. "Catalyzed Steam Gasification of Biomass." Journal of Analytical and Applied Pyrolysis. 1:165–175.

Mudge, L. K., S. L. Weber, D. H. Mitchell, L. J. Sealock, Jr., R. J. Robertus. 1981. Investigations on Catalyzed Steam Gasification of Biomass (Including Appendix A, B, C, and D), PNL-3695, Pacific Northwest Laboratory, Richland, Washington.

Mudge, L. K., E. G. Baker, D. H. Mitchell, R. J. Robertus, and M. D. Brown. 1983. Catalytic Gasification Studies in a Pressurized Fluid Bed Unit. PNL-4594, Pacific Northwest Laboratory, Richland, Washington.

Nahas, N. C. 1983. "Exxon Catalytic Coal Gasification Process." Fuel. 62 (2):239-241.

Rostrup-Nielsen, J. R. 1975. Steam Reforming Catalysts. Danish Technical Press, Copenhagen.

Sealock, L. J., D. C. Elliott, and R. T. Hallen. 1982. Kinetics and Catalysis of Producing Synthetic Gases from Biomass. GRI-82/0038. Battelle, Pacific Northwest Laboratories, Richland, Washington.

Shafizadeh, F. 1982. "Introduction to Pyrolysis of Biomass." J. Anal. Appl. Pyrolysis. 3: 283-305.

Skamser, R. 1978. Coal Gasification, Commercial Concepts, Gas Cost Guidelines. FE-2240-100. C. t. Braun & Co., California.

Stern, E. W. 1982. Bench-Scale Development of Catalysts for Reforming Aromatic and Heterocyclic Hydrocarbons. DOE/ET/11029-1191. Engelhard Corporation, Edison, New Jersey.

Taylor, H. S. and H. A. Neville. 1921. J. Am. Chem. Soc. 43:2065.

Walker, P. L., et al. 1983. "Catalysis of Coal-Derived Cokes and Chars." Fuel. 62 (2):140-149.

APPENDIX A

DESCRIPTION OF CATALYSTS USED FOR GASIFICATION TESTS AND PDU RUN SUMMARIES

PDU RUN SUMMARIES

Following are summaries for all of the PDU tests with bagasse. Most of the terms from the computer output need no further explanation. The following list is a key to terms that may need further definition. All stream flows are in pounds.

BED - solids from the reactor

BED GAS VELOCITY - superficial linear velocity in ft/sec at reactor conditions (based on an empty reactor)

BURNER TRAP - condensate liquid collected from the product gas downstream from the demister

CART HEAT COEFF - overall heat transfer coefficient from heaters in Btu/hr-ft^2-°F

CCHAR - char from the cyclone, also called CYCLONE SOLIDS

DRY MOL WT - molecular weight of the dry product gas

FCHAR - char from the filter; also called FILTER SOLIDS

METHANOL #/# WOOD - total possible yield in 1b methanol/1b dry wood if (1) all H_2+CO were converted and (2) if all H_2+CO+CH_4 were converted

N2-02 FREE MW - molecular weight of dry, O_2-N_2 free product gas

RECYCLE - refers to the recycle gas stream

HEATS OF COMBUSTION - standard heat of combustion in Btu/lb

STEAM - refers to steam from generator; heated to reactor inlet temperature in the gas preheater

TAR - tars from the demister

TOC-DEMISTER (PPM) - total organic carbon in the demister liquid (aqueous) in ppm by weight

WET MOL WT - molecular weight of product gas saturated with water (used to calculate flow rate)

WT FRAC CAT-CCHAR - weight fraction of catalyst in the cyclone char

WT FRAC CAT-FCHAR - weight fraction of catalyst in the filter char

YH20 - mole fraction of water determined by assuming saturation in the gas stream

INST SCF/# DRY FEED - instantaneous gas production based on current gas flow (standard ft^3 per lb of dry feedstock)

AVG SCF/# DRY FEED - integrated average gas production (over the entire run)

TOTAL FLOW (SCF) - total dry gas production (standard cubic feet)

TOTAL BAGASSE FED - total bagasse in pounds

THEORETICAL ASH IN CHAR - % Ash in the char if all the ash in the bagasse gasified stays in the char

CAT LOSS THIS RUN - Pounds of catalyst entrained out of the gasifier

STEAM BAGASSE RATIO - lb/hr of steam divided by lb/hr of dry bagasse

ELECTRIC INPUT (KWHR/HR) - rate of electric input to the cartridge heaters

KWHR/LB DRY BAGASSE - electric heat required to gasify 1 lb to dry bagasse

ELECTRIC BTu/LB DRY - conversion of above

HEAT GAS/HEAT BAGASSE - conversion efficiency (not including heat input)

GAS BTU/LB BAGASSE - efficiency (as above)

RUN#: B1 TIME: 1630
DATE: 120382 TOTAL RUN TIME(HRS) :3.75
FEEDSTOCC::: BAGASSE PELLETS CATALYST: Ni on A1203(FINE)

PRODUCT GAS COMPOSITIONS
(MOLE FRACS)

		N2,02 FREE
H2	.562	.566
CO2	.240	.242
C2H4	0.000	0.000
C2H6	0.000	0.000
02	.004	0.000
N2	.003	0.000
CH4	.041	.041
CO	.150	.151
C3H6	0.000	0.000
C3H8	0.000	0.000
YH20	.017	.017

TEMPERATURES
(DEG C)

REACTOR INLET	751
CATALYST BED	749
REACTOR OUTLET	693
STEAM	127
CARTR▮DGE HTRS	828
FHODUCT METER	16
DEM▮STER	23

DIFFERENTIAL PRESSURES

FRODUCT 9.1 IN H20
STEAM 20 IN H20
RECYCLE 0 IN H20

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.9	.50
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	25
DEMISTER	25
REACTOR	20
PRODUCT	.5

** SOLIDS **
** COMPOSITIONS **

	CARBON	HYDROGEN	OXYGEN	ASH	MOISTURE
EAGAS	.445	.059	.441	.055	.100
CCHAR	.450	.010	.040	.500	.050
FCHAR	.450	.010	.140	.400	.005
BED	.820	.030	.130	.020	0.000
TAR	.840	.030	.130	.001	0.000

HEATS OF COMBUSTION

BAGASSE	8500
CYCLONE CHAR	6900
FILTER CHAR	6900
BED	10000
TAR	15000

****
*** INPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED (DRY)	30.60	13.62	1.81	13.49	1.68
FEED MOISTURE	3.40	0.00	.38	3.02	0.00
STEAM	33.59	0.00	3.73	29.86	0.00
TOTAL INPUTS	67.59	13.62	5.92	46.37	1.68

****
*** OUTPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	35.91	11.23	2.80	22.16	0.00
MOISTURE IN PRODUCT	.69	0.00	.08	.61	0.00
CYCLONE SOLIDS	2.80	1.21	.03	.21	1.35
FILTER SOLIDS	1.20	.94	.01	.17	.48
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DEMISTER LIQUIDS	24.80	.00	2.76	22.04	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
BURNER TRAP	1.10	.00	.12	.98	0.00
TOTAL OUTPUTS	66.50	12.98	9.79	46.07	1.83

****
** PRODUCT GAS (N2-O2 FREE) **
****

SCFM(DRY)	13.71
LR GAS/LB DRY FEED	1.17
INST SCF/# DRY FEED	26.70
AVG SCF/# DRY FEED	29.35
TOTAL FLOW(SCF)	3212.36
TOTAL BAGASSE FED	109.44
YH2O AT PRODUCT	.02
YH2O AT REACTOR OUT	.41
BTU/SCF	272.31

****
MISCELLANEOUS TIDBITS
****

WT FRAC CAT-CCHAR	.04
WT FRAC CAT-FCHAR	0.00
#CAT LOSS THIS RUN	.11
TOC-DEMISTER(PPM)	55.00
STEAM/BAGASSE RATIO	1.10
DRY MOL WT	16.75
WET MOL WT	16.77
N2-O2 FREE MW	16.66
ELEC INPUT(KWH/HR)	31.58
(based on 7 elements)	
KWHR/LB DRY BAGASS	1.03
ELECTRIC BTU/LB DRY	3523.09

CONVERSIONS & BALANCES

HEAT GAS/HEAT BAGASS	.86
GAS BTUS/LB BAGASS	7270

CARBON CONVERSION

TO GAS(%)	82.45
TO SOLID(%)	12.86
TO LIQUID(%)	.01
SUM CONVERSIONS	95.32

METHANOL #/#BAGASSE .592335426662 TO .669273478406
THEORETICAL ASH IN CHAR(%) 49.54

DESIGN INFORMATION

BED GAS VELOCITY	.91
CART HEAT COEFF	137.97

RUN#: C1 TIME: 0130
DATE: 120682 TOTAL RUN TIME(HRS) :7.5
FEEDSTOCK: BAGASSE PELLETS CATALYST: GOOD NI-CU-MO

PRODUCT GAS COMPOSITIONS
(MOLE FRACS)

		N2,02 FREE
H2	.556	.560
CO2	.196	.198
C2H4	.002	.002
C2H6	.002	.002
02	.005	0.000
N2	.003	0.000
CH4	.029	.029
CO	.208	.210
C3H6	0. 000	0.000
C3H8	0.000	0.000
YH20	.006	.006

TEMPERATURES
(DEG C)

REACTOR INLET	745
CATALYST BED	775
REACTOR OUTLET	715
STEAM	123
CARTRIDGE HTRS	845
PRODUCT METER	0
DEMISTER	13

DIFFERENTIAL PRESSURES

PRODUCT 4 IN H20
STEAM 16 IN H20
RECYCLE 0 IN H20

#*ORIFICE SIZES AND CONSTANTS**

ORIFICE	K	SIZE
PRODUCT	110.9	.50
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	15.2
DEMISTER	14
REACTOR	14
PRODUCT	0

** SOLIDS **
** COMPOSITIONS **

	CARBON	HYDROGEN	OXYGEN	ASH	MOISTURE
BAGAS	.445	.059	.441	.055	.100
CCHAR	.450	.100	.040	.500	.050
FCHAR	.450	.100	.140	.400	.005
BED	.820	.030	.130	.020	0.000
TAR	.840	.030	.130	.001	0.000

HEATS OF COMBUSTION

BAGASSE	8500
CYCLONE CHAR	6900
FILTER CHAR	6900
BED	10000
TAR	15000

****
*** INPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED (DRY)	22.50	10.01	1.33	9.32	1.24
FEED MOISTURE	2.50	0.00	.28	2.22	0.00
STEAM	26.20	0.00	2.91	23.29	0.00
TOTAL INPUTS	51.20	10.01	4.52	35.44	1.24

****
*** OUTPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	24.05	7.87	1.85	14.56	0.00
MOISTURE IN PRODUCT	.16	0.00	.02	.14	0.00
CYCLONE SOLIDS	2.00	.82	.18	.07	.91
FILTER SOLIDS	.50	.23	.05	.07	.20
BED BUILDUP	.07	.05	.00	.01	.01
DEMISTER LIQUIDS	21.60	.02	2.40	19.20	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
BURNER TRAP	1.00	.00	.11	.89	0.00
TOTAL OUTPUTS	49.38	8.98	4.62	34.94	1.12

****
** PRODUCT GAS (N2-02 FREE) **
****

SCFM!DRY)	9.41
LB GAS/LB DRY FEED	1.07
INST SCF/# DRY FEED	24.92
AVG SCF/# DRY FEED	22.74
TOTAL FLOW(SCF)	3749.10
TOTAL BAGASSE FED	164.88
YH20 AT PRODUCT	.01
YH20 AT REACTOR OUT	.46
BTU/SCF	282.42

****
MISCELLANEOUS TIDBITS
****

WT FRAC CAT-CCHAR	.09
WT FRAC CAT-FCHAR	0.00
#CAT LOSS THIS RUN	.18
TOC-DEMISTER(PPM)	800.00
STEAM/BAGASSE RATIO	1.16
DRY MOL WT	16.36
WET MOL WT	16.37
N2-02 FREE MW	16.25
ELEC INPUT(KWH/HR)	20.22
(based on 7 elements)	
KWHR/LB DRY BAGASS	.90
ELECTRIC BTU/LB DRY	3068.39

XXCONVERSIONS & BALANCESXX

HEAT GAS/HEAT BAGASS	.83
GAS BTUS/LB BAGASS	7039

CARBON CONVERSION

TO GAS(%)	78.56
TO SOLID(%)	10.93
TO LIQUID(%)	.18
SUM CONVERSIONS	89.68

METHANOL #/#BAGASSE .492709170432 TO .534738365125
THEORETICAL ASH IN CHAR(%) 40.55

XXDESIGN INFORMATION**

BED GAS VELOCITY	.88
CART HEAT COEFF	99.71

RUN#: C2 TIME: 1540
DATE: 120782 TOTAL RUN TIME(HRS) :6.5
FEEDSTOCK: BAGASSE PELLETS CATALYST: NI-CU-MO 2ND RUN

**PRODUCT GAS CONPOSITIONSXX
(MOLE FRACS)

		N2,02 FREE
HZ	.509	.512
CO2	.221	.223
C2H4	.010	.010
C2H6	.004	.004
02	.005	0.000
N2	.002	0.000
CH4	.071	.072
CO	.178	.179
C3H6	.000	.000
C3H8	0.000	0.000
YH2O	.008	.008

TEMPERATURES
(DEG C)

REACTOR INLET	730
CATALYST BED	780
REACTOR OUTLET	725
STEAM	150
CARTRIDGE HTRS	850
PRODUCT METER	4
DEMISTEK	70

DIFFERENTIAL PRESSURES

PRODUCT 2.5 IN H2O
STEAM 20 IN H2O
RECYCLE 0 IN H2O

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.7	.50
STEAM	32.0	.30
RECYCLE	116.0	.90

PRESSURES
(PSIG)

STEAM	16
DEMISTER	16
REACTOR	16
PRODUCT	1

**** SOLIDS ****
**** COMPOSITIONS ****

	CARBON	HYDROGEN	OXYGEN	ASH	MOISTURE
BAGAS	.445	.059	.441	.055	.100
CCHAR	.450	.100	.040	.500	.050
FCHAR	.450	.100	.140	.400	.005
BED	.820	.030	.130	.020	0.000
TAR	.840	.030	.130	.001	0.000

****HEATS OF COMBUSTION****

BAGASSE	8500
CYCLONE CHAR	6900
FILTER CHAR	6900
BED	10000
TAR	15000

****
***** INPUT STREAMS *****
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED(DRY)	26.28	11.69	1.95	11.59	1.45
FEED MOISTURE	2.92	0.00	.32	2.60	0.00
STEAM	29.43	0.00	3.27	26.16	0.00
TOTAL INPUTS	58.63	11.69	5.14	40.34	1.45

****
***** OUTPUT STREAMS *****
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGGSES	20.19	7.00	1.60	11.78	0.00
MOISTURE IN PRODUCT	.16	0.00	.02	.14	0.00
CYCLONE SOLIDS	2.20	.89	.20	.08	.99
FILTER SOLIDS	.50	.23	.05	.07	.20
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DEMISTER LIQUIDS	28.10	.08	3.12	24.98	0.00
DEMISTER TAR	.10	.08	.00	.01	0.00
BURNER TRAP	2.00	.01	.22	1.70	0.00
TOTAL OUTPUTS	53.25	8.29	5.21	38.83	1.19

****
** PRODUCT GAS (N2-O2 FREE) **
****

SCFM(DRY)	7.37
LB GAS/LB DRY FEED	.77
INST SCF/# DRY FEED	16.74
AVG SCF/# DRY FEED	16.67
TOTAL FLOW(SCF)	2851.25
TOTAL BAGASSE FED	171.00
YH2O AT PRODUCT	.01
YH2O AT REACTOR OUT	.S9
BTU/SCF	318.45

****
MISCELLANEOUS TIDBITS
****

WT FRAC CAT-CCHAR	.10
WT FRAC CAT-FCHAR	0.00
#CAT LOSS THIS RUN	.22
TOC-DEMISTER(PPM)	3000.00
STEAM/BAGASSE RATIO	1.12
DRY NOL WT	17.49
WET MOL WT	17.49
N2-O2 FREE MW	17.40
ELEC INPUT(KWH/HR)	22.79
(based on 7 elements)	
KWHR/LB DRY BAGASS	.87
ELECTRIC BTU/LB DRY	2961.25

CONVERSIONS & BALANCES

HEAT GAS/HEAT BAGASS	.63
GAS BTUS/LB BAGASS	5330

CARBON CONVERSION

TO GAS(%)	99.84
TO SOLID(%)	3.54
TO LIQUID(%)	1.49
SUM CONVERSIONS	70.87

METHANOL #/#BAGASSE .324421334896 TO .40019962865
THEORETICAL ASH IN CHAR(%) 21.64

#*DESIGN INFORMATION**

BED GAS VELOCITY	.93
CART HEAT COEFF	112.40

RUN#: B1-A TIME: 09:17
DATE: 011483 TOTAL RUN TIME(HRS) :2
FEEDSTOCK: BAGASSE CATALYST: NI ON AL203

PRODUCT GAS COMPOSITIONS
(MOLE FRACS)

		N2,02 FREE
H2	.550	.561
C02	.214	.218
C2H4	0.000	0.000
C2H6	0.000	0.000
02	.004	0.000
N2	.015	0.000
CH4	.037	.038
CO	.179	.183
C3H6	0.000	0.000
C3H8	0.000	0.000
YH20	.006	.006

TEMPERATURES
(DEG C)

REACTOR INLET	740
CATALYST BED	740
REACTOR OUTLET	700
STEAM	122
CARTRIDGE HTRS	850
PRODUCT METER	0
DEMISTER	22

**DIFFERENTIAL PRESSURESXX

PRODUCT 12 IN H20
STEAM 22 IN H20
RECYCLE 0 IN H20

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.9	.50
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	10
DEMISTER	10
REACTOR	8
PRODUCT	1

** SOLIDS **
aX COMPOSITIONS **

	CARBON	HYDROGEN	OXYGEN	ASH	NOISTURE
BAGAS	.445	.059	.441	.055	.100
CCXAR	.250	.100	.040	.700	.050
FCHAR	.250	.100	.140	.700	.005
BED	.820	.030	.130	.020	0.000
TAR	.840	.030	.130	.001	0.000

HEATS OF COMBUSTION

BAGASSE	8500
CYCLONE CHAR	4000
FILTER CHAR	4000
RED	10000
TAR	15000

****
*** INPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED(DRY)	32.04	14.26	1.89	14.13	1.76
FEED MOISTURE	3.56	0.00	.40	3.16	0.00
STEAM	27.96	0.00	3.11	24.86	0.00
TOTAL INPUTS	63.56	14.26	5.39	42.15	1.76

****
*** OUTPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	42.65	13.66	3.30	26.06	0.00
MOISTURE IN PRODUCT	.27	0.00	.03	.24	0.00
CYCLONE SOLIDS	5.00	.88	.35	.14	2.45
FILTER SOLIDS	.50	.09	.04	.05	.25
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DEMISTER LIQUIDS	14.00	.00	1.56	12.44	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
BURNER TRAP	1.00	.00	.11	.89	0.00
TOTAL OUTPUTS	63.42	14.62	5.38	39.82	2.70

****
** PRODUCT GAS (N2-O2 FREE) **
****

SCFM(DRY)	16.68
LB GAS/LB DRY FEED	1.33
INST SCF/# DRY FEED	30.67
AVG SCF/# DRY FEED	26.61
TOTAL FLOW(SCF)	1724.06
TOTAL BAGASSE FED	64.80
YH20 AT PRODUCT	.01
YH20 AT REACTOR OUT	.25
BTU/SCF	277.52

****
MISCELLANEOUS TIDBITS
****

WT FRAC CAT-CCHAR	.30
WT FCAC CAT-FCHAR	.30
#CAT LOSS THIS RUN	1.65
TOC-DEMISTER(PPM)	100.00
STEAM/BAGASSE RATIO	.87
DRY MOL WT	16.69
WET MOL WT	16.70
N2-O2 FREE MW	16.45
ELEC INPUT(KWH/HR)	29.75
(based on 7 elements)	
KWHR/LB DRY BAGASS	.93
ELECTRIC BTU/LB DRY	3169.99

CONVERSIONS & BALANCES

HEAT GAS/HEAT BAGASS	1.00
GAS BTUS/LB EAGASS	8511

CARBON CONVERSION

TO GAS(%)	95.80
TO SOLID(%)	6.75
TO LIQUID(%)	.01
SUN CONVERSIONS	102.56

METHANOL #/#BAGASSE .556921814298 TO .621016155014
THEORETICAL ASH IN CHAR(%)

DESIGN INFORMATION

BED GAS VELOCITY	1.15
CART HEAT COEFF	93.35

RUN#: B2-B TIME: 15:00
DATE: 011583 TOTAL RUN TIME(HRS) :4
FEEDSTOCK: BAGASSE CATALYST: NI ON AL203

PRODUCT GAS COMPOSITIONS
(MOLE FRACS)

		N2,02 FREE
H2	.559	.568
CO2	.199	.203
C2H4	.002	.002
C2H6	.001	.001
02	.003	0.000
NZ	.013	0.000
CH4	.021	.021
CO	.201	.205
C3H6	0.000	0.000
C3H8	0.000	0.000
YH20	.008	.008

TEMPERATURES
(DEG C)

REACTOR INLET	740
CATALYST BED	740
REACTOR OUTLET	700
STEAM	122
CARTRIDGE HTRS	850
PRODUCT METER	5
DEMISTER	25

DIFFERENTIAL PRESSURES

PRODUCT 11 IN H20
STEAM 30 IN H20
RECYCLE 0 IN H20

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.9	.50
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	10
DEMISTER	10
REACTOR	8
PRODUCT	1

** SOLIDS **
** COMPOSITIONS **

	CARBON	HYDROGEN	OXYGEN	ASH	MOISTURE
BAGAS	.445	.059	.441	.055	.100
CCHAR	.350	.100	.040	.600	.050
FCHAR	.350	.100	.040	.600	.005
BED	.820	.030	.130	.020	0.000
TAR	.840	.030	.130	.001	0.000

HEATS OF COMBUSTION

BAGASSE	8500
CYCLONE CHAR	6900
FILTER CHAR	6900
BED	10000
TAR	15000

****
*** INPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED (DRY)	32.04	14.26	1.89	14.13	1.76
FEED MOISTURE	3.56	0.00	.40	3.16	0.00
STEAM	32.65	0.00	3.63	29.03	0.00
TOTAL INPUTS	68.25	14.26	5.91	46.32	1.76

****
*** OUTPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	40.24	12.95	3.07	24.46	0.00
MOISTURE IN PRODUCT	.37	0.00	.04	.33	0.00
CYCLONE SOLIDS	6.00	1.68	.48	.19	2.88
FILTER SOLIDS	.50	.16	.05	.02	.27
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DENISTER LIQUIDS	18.00	.00	2.00	16.00	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
BURNER TRAP	1.00	.00	.11	.89	0.00
TOTAL OUTPUTS	66.11	14.79	5.74	41.89	3.15

```
*****************************
** PRODUCT GAS (N2-02 FREE) **
*****************************

SCFM(DRY)                15.92
LB GAS/LB DRY FEED        1.26
INST SCF/# DRY FEED      29.35
AVG SCF/# DRY FEED       23.64

TOTAL FLOW(SCF)        3064.33

TOTAL BAGASSE FED       129.60
YH20 AT PRODUCT            .01
YH20 AT REACTOR OUT        .31
BTU/SCF                 275.36


************************
**MISCELLANEOUS TIDBITS**
************************

WT FRAC CAT-CCHAR          .20
WT FRAC CAT-FCHAR          .10
#CAT LOSS THIS RUN        1.25

TOC-DEMISTER(PPM)       100.00

STEAM/BAGASSE RATIO       1.02

DRY MOL WT               16.42
WET MOL WT               16.43
N2-02 FREE MW            16.22

ELEC INPUT(KWH/HR)       25.96
(based on 7 elements)
KWHR/LB DRY BAGASS         .81
ELECTRIC BTU/LB DRY   2765.97


XXCONVERSIONS & BALANCES**

HEAT GAS/HEAT BAGASS       .95

GAS BTUS/LB BAGASS  8082

CARBON CONVERSION
-----------------

TO GAS(%)                90.86
TO SOLID(%)              12.89
TO LIQUID(%)               .01
SUM CONVERSIONS         103.76

METHANOL #/#BAGASSE .514194796426 TO .546230302963
THEORETICAL ASH IN CHAR(%)    95.11

**DESIGN INFORMATIONXX
BED GAS VELOCITY          1.35
CART HEAT COEFF          81.45
```

RUN#: C-3 TIME: 05:00
DATE: 061683A TOTAL RUN TIME(HRS) :7.8
FEEDSTOCK: BAGASSE W/10%K2CO3 CATALYST: NONE

PRODUCT GAS COMPOSITIONS
(MOLE FRACS)

		N2,02 FREE
H2	.525	.545
CO2	.231	.240
C2H4	.005	.006
C2H6	.003	.003
02	.003	0.000
N2	.033	0.000
CH4	.035	.036
CO	.163	.170
C3H6	.001	.001
C3H8	.000	.000
YH2O	.016	.016

TEMPERATURES
(DEG C)

REACTOR INLET	754
CATALYST BED	760
REACTOR OUTLET	721
STEAM	116
CARTRIDGE HTRS	845
PRODUCT METER	15
DEMISTER	23

DIFFERENTIAL PRESSURES

F'RODUCT 7 IN H2O
STEAM 25 IN H2O
RECYCLE 0 IN H2O

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.9	.SO
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	10
DEMISTER	10
REACTOR	5
PRODUCT	1

** SOLIDS **
** COMPOSITIONS **

	CARBON	HYDROGEN	OXYGEN	ASH	MOISTURE
BAGAS	.389	.051	.427	.133	.100
CCHAR	.446	.010	.036	.508	0.000
FCHAR	.470	.010	.012	.508	0.000
BED	.400	.010	.090	.500	0.000
TAR	.840	.030	.130	.001	0.000

HEATS OF COMBUSTION

BAGASSE	7050
CYCLONE CHAR	3550
FILTER CHAR	3550
FED	3550
TAR	15000

****
*** INPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED (DRY)	31.50	12.25	1.61	13.45	4.19
FEED MOISTURE	3.50	0.00	.39	3.11	0.00
STEAM	30.04	0.00	3.34	26.70	0.00
TOTAL INPUTS	65.04	12.25	5.33	43.26	4.19

****
*** OUTPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	31.45	10.17	2.33	19.13	0.[illegible]
MOISTURE IN PRODUCT	.55	0.00	.06	.49	0.00
CYCLONE SOLIDS	5.00	2.23	.05	.18	2.54
FILTER SOLIDS	1.00	.47	.01	.01	.51
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DEMISTER [illegible]	25.00	.02	2.78	22.22	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
BURNER TRAP	.50	.00	.06	.44	0.00
TOTAL OUTPUTS	63.50	12.90	5.29	42.48	3.05

****
** PRODUCT GAS (N2-O2 FREE) **
****

SCFM(DRY)	11.91
LB GAS/LB DRY FEED	1.00
INST SCF/# DRY FEED	21.9.3
AVG SCF/# DRY FEED	24.60
TOTAL FLOW(SCF)	5224.25
TOTAL BAGASSE FED	212.40
YH2O AT PRODUCT	.02
YH2O AT REACTOR OUT	.45
BTU/SCF	282.25

****
MISCELLANEOUS TIDBITS
****

WT FRAC CAT-CCHAR	0.00
WT FRAC CAT-FCHAR	0.00
#CAT LOSS THIS RUN	0.00
TOC-DEMISTER(PPM)	934.00
STEAM/BAGASSE RATIO	.95
DRY MOL WT	17.66
WET HOL WT	17.66
N2-O2 FREE MW	17.25
ELEC INPUT(KWH/HR)	24.98
(based on 7 elements)	
KWHR/LB DRY BAGASS	.77
ELECTRIC BTU/LB DRY	2707.21

CONVERSIONS & BALANCES

HEAT GAS/HEAT BAGASS	.88
GAS BTUS/LB BAGASS	6191

CARBON CONVERSION

TO GAS(%)	83.03
TO SOLID(%)	22.03
TO LIQUID(%)	.19
SUM CONVERSIONS	105.26

METHANOL #/#BAGASSE .494715121242 TO .551236766597
THEORETICAL ASH IN CHAR(%) 51.24

DESIGN INFORMATION

BED GAS VELOCITY	1.46
CART HEAT COEFF	101.43

RUN#: C-3B TIME: 10:00
DATE: 0617/83 TOTAL RUN TIME(HRS) :2.333
FEEDSTOCK: BAGASSE CATALYST: 10%K2CO3

PRODUCT GAS COMPOSITIONS
(MOLE FRACS)

		N2,02 FREE
H2	.548	.553
CO2	.243	.245
C2H4	.006	.006
C2H6	.004	.004
02	.004	0. 000
N2	.004	0.000
CH4	.040	.040
CO	.150	.151
C3H6	.001	.001
C3H8	.000	.000
YH20	.016	.016

XXTEMPERATURESXX
(DEG C)

REACTOR INLET	754
CATALYST RED	760
REACTOR OUTLET	721
STEAM	116
CARTRIDGE HTRS	845
PRODUCT METER	15
DEMISTER	23

DIFFERENTIAL PRESSURES

PRODUCT 12 IN H20
STEAM 25 IN H20
RECYCLE 0 IN H20

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.9	.50
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	10
DENISTER	10
REACTOR	5
PRODUCT	1

** SOLIDS **
** COMPOSITIONS **

	CARBON	HYDROGEN	OXYGEN	ASH	MOISTURE
BAGAS	.389	.051	.427	.133	.100
CCHAR	.446	.010	.036	.508	0.000
FCHAR	.470	.010	.012	.508	0.000
BED	.400	.010	.090	.500	0.000
TAR	.840	.030	.130	.001	0.000

**HEATS OF COMBUSTIONXX

BAGASSE	7050
CYCLONE CHAR	3550
FILTER CHAR	3550
BED	3550
TAR	15000

****
*** INPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED (DRY)	43.20	16.80	2.20	18.45	5.75
FEED MOISTURE	4.80	0.00	.53	4.27	0.00
STEAM	30.04	0.00	3.34	26.70	0.00
TOTAL INPUTS	78.04	16.80	6.07	49.41	5.75

****
*** OUTPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	42.54	13.72	3.29	25.84	0.00
MOISTURE IN PRODUCT	.72	0.00	.08	.64	0.00
CYCLONE SOLIDS	7.00	3.12	.07	.25	3.56
FILTER SOLIDS	1.00	.47	.01	.01	.51
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DEMISTER LIQUIDS	25.00	.01	2.78	22.22	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
BURNER TRAP	.50	.00	.06	.44	0.00
TOTAL OUTPUTS	76.76	17.32	6.29	49.42	4.06

```
*****************************
** PRODUCT GAS (N2-02 FREE) **
*****************************

SCFM(DRY)                15.85
LB GAS/LB DRY FEED         .98
INST SCF/# DRY FEED      21.85
AVG SCF/# DRY FEED       20.48

TOTAL FLOW(SCF)        2083.22

TOTAL BAGASSE FED       101.70
YH20 AT PRODUCT            .02
YH20 AT REACTOR OUT        .37
BTU/SCF                  286.24


************************
XXMISCELLANEOUS TIDBITS**
************************

WT FRAC CAT-CCHAR          0.00
WT FRAC CAT-FCHAR          0.00
#CAT LOSS THIS RUN         0.00

TOC-DEMISTER(PPM)       315.00

STEAM/BAGASSE RATIO        .70

DRY MOL WT               17.18
WET MOL WT               17.20
N2-02 FREE MW            17.08

ELEC INPUT(KWH/HR)       24.50
(based on 7 elements)
KWHR/LB DRY EAGASS         .57
ELECTRIC BTU/LB DRY    1936.46


**CONVERSIONS & BALANCES**

HEAT GAS/HEAT BAGASS       .89

GAS BTUS/LB BAGASS    6254

CARBON CONVERSION
-----------------

TO GAS(%)                81.66
TO SOLID(%)              21.37
TO LIQUID(%)               .05
SUM CONVERSIONS         103.08

METHANOL #/#BAGASSE .405809562544 TO .458190673589
THEORETICAL ASH IN CHAR(%)    47.40

**DESIGN INFORMATION**
BED GAS VELOCITY          1.46
CART HEAT COEFF          99.50
```

RUN#: C-4A TIME: 19:00
DATE: 0618/83 TOTAL RUN TIME(HRS) :5
FEEDSTOCK: BAGASSE CATALYST: 10WT%K2CO3

**PRODUCT GAS COMPOSITIONSXX
(MOLE FRACS)

		N2,02 FREE
H2	.511	.514
CO2	.241	.242
C2H4	.007	.007
C2H6	.004	.004
02	.003	0.000
N2	.003	0.000
CH4	.030	.030
CO	.200	.202
C3H6	.000	.000
C3H8	.000	.000
YH20	.049	.049

TEMPERATURES
(DEG C)

REACTOR INLET	759
CATALYST BED	730
REACTOR OUTLET	711
STEAM	123
CARTRIDGE HTRS	850
PRODUCT METER	34
DEMISTER	39

DIFFERENTIAL PRESSURES

PRODUCT 22 IN H20
STEAM 26 IN H20
RECYCLE 0 IN H20

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.7	.50
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	15
DEMISTER	15
REACTOR	10
PRODUCT	1

**** SOLIDS ****
**** COMPOSITIONS ****

	CARBON	HYDROGEN	OXYGEN	ASH	MOISTURE
BAGGS	.389	.051	.427	.133	.100
CCHAR	.446	.010	.036	.508	0.000
FCHAR	.470	.010	.012	.508	0.000
BED	.400	.010	.090	.500	0.000
TAR	.840	.030	.130	.001	0.000

****HEATS OF COMBUSTIONXX**

BAGASSE	7050
CYCLONE CHAR	3550
FILTER CHAR	3550
RED	3550
TAR	15000

****
***** INPUT STREAMS *****
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED(DRY)	54.00	21.01	2.75	23.06	7.18
FEED MOISTURE	6.00	0.00	.67	5.33	0.00
STEAM	33.29	0.00	3.70	29.59	0.00
TOTAL INPUTS	***93.29***	21.01	7.12	57.98	7.18

****
***** OUTPUT STREAMS *****
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	55.77	18.36	3.71	34.00	1.00
MOISTURE IN PRODUCT	2.88	0.00	.32	***2.56***	0.00
CYCLONE SOLIDS	7.00	3.12	.07	.25	3.56
FILTER SOLIDS	1.50	.71	.02	.02	.76
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DEMISTER LIQUIDS	25.00	.12	2.78	22.22	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
RURNER TRAP	.50	.00	.06	.44	0.00
TOTGL OUTPUTS	92.65	22.31	6.94	59.50	4.32

****
** PRODUCT GAS (N2-02 FREE) **
****

SCFM(DRY)	19.51
Lb GAS/LB DRY FEED	1.03
INST SCF/# DRY FEED	21.56
AVG SCF/# DRY FEED	22.18
TOTAL FLOW(SCF)	4989.8 1
TOTAL BAGASSE FED	225.00
YH20 AT PRODUCT	.05
YHZO AT REACTOR OUT	.34
BTU/SCF	280.70

****
MISCELLANEOUS TIDBITS
****

WT FRAC CAT-CCHAR	0.00
WT FRAC CAT-FCHAR	0.00
#CAT LOSS THIS RUN	0.00
TOC-DEMISTER(PPM)	4778.00
STEAM/BAGASSE RATIO	.62
DRY MOL WT	18.23
WET MOL WT	18.21
N2-02 FREE MW	18.15
ELEC INPUT(KWH/HR)	25.20
(based on 7 elements)	
KWHR/LB DRY BAGASS	.47
ELECTRIC BTU/LB DRY	1593.20

**CONVERSIONS & RALANCESXX

HEAT GAS/HEAT BAGASS	.86
GAS BTUS/LB BAGASS	6052

CARBON CONVERSION

TO GAS(%)	87.59
TO SOLID(%)	18.22
TO LIQUID(%)	.58
SUM CONVERSIONS	106. 18

METHANOL #/#BAGASSE .446902332038 TO .489496009771
THEORETICAL ASH I N CHAR(%) 613.91

**DESIGN INFORMATIONXX

BED GAS VELOCITY	1.25
CART HEAT COEFF	72.48

RUN#: C-5 TIME: 13:11
DATE: 07/18/83: T TOTAL RUN TIME(HRS) :3.0
FEEDSTOCK: BAGASSE W/K2CO3 CATALYST: 10% K2CO3 IMFREG

XXFRODUCT GAS COMPOSITIONS**
(MOLE FRACS)

		N2,02 FREE
HZ	.463	.468
CO2	.245	.248
C2H4	.009	.009
C2H6	.007	.007
02	.003	0.000
N2	.007	0.000
CH4	.038	.039
CO	.224	.226
C3H6	.003	.003
C3H8	.001	.001
YH20	.044	.044

XXTEMPERATURESXX
(DEG C)

REACTOR INLET	712
CATALYST BED	720
REACTOR OUTLET	698
STEAM	129
CARTRIDGE HTRS	815
PRODUCT METER	32
DEMISTER	40

XIDIFFERENTIAL PRESSURES**

PRODUCT 24 IN H20
STEAM 30 IN H20
RECYCLE 1 IN H20

ORIFICE SIZES AND CONSTANTS

ORIFICE	K	SIZE
PRODUCT	110.9	.50
STEAM	32.0	.30
RECYCLE	116.0	.50

PRESSURES
(PSIG)

STEAM	21
DEMISTER	21
REACTOR	17
PRODUCT	1

** SOLIDS **
** COMPOSITIONS **

	CARBON	HYDROGEN	OXYGEN	ASH	NOISTURE
BAGAS	.403	.051	.410	.136	.071
CCHAR	.425	.010	.080	.485	0.000
FCHAR	.507	.010	.095	.388	0.000
BED	.400	.010	.090	.500	0.000
TAR	.840	.030	.130	.001	0.000

HEATS OF COMBUSTION

BAGASSE	7050
CYCLONE CHAR	3550
FILTER CHAR	3550
BED	3550
TAR	15000

****
*** INPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
FEED(DRY)	67.35	27.14	3.43	27.61	9.16
FEED MOISTURE	5.15	0.00	.57	4.58	0.00
STEAM	38.91	0.00	4.32	34.59	0.00
TOTAL INPUTS	111.41	27.14	8.33	66.78	9.16

****
*** OUTPUT STREAMS ***
****

	TOTAL	CARBON	HYDROGEN	OXYGEN	ASH
DRYGASES	60.50	20.77	3.71	36.37	0.00
MOISTURE IN PRODUCT	2.61	0.00	.29	2.32	0.00
CYCLONE SOLIDS	9.90	4.21	.10	.79	4.80
FILTER SOLIDS	4.00	2.03	.04	.38	1.55
BED BUILDUP	0.00	0.00	0.00	0.00	0.00
DEMISTER LIQUIDS	29.50	.20	3.28	26.22	0.00
DEMISTER TAR	0.00	0.00	0.00	0.00	0.00
BURNER TRAP	0.00	0.00	0.00	0.00	0.00
TOTAL OUTPUTS	106.51	27.20	7.42	66.08	6.35

****
** PRODUCT GAS (N2-02 FREE) **
****

SCFM(DRY)	19.90
LB GAS/LB DRY FEED	.90
INST SCF/# DRY FEED	17.55
AVG SCF/# DRY FEED	15.67
TOTAL FLOW(SCF)	3166.44
TOTAL BAGASSE FED	202.06
YH20 AT PRODUCT	.04
YH20 AT REACTOR OUT	.37
BTU/SCF	297.02

****
MISCELLANEOUS TIDBITS
****

WT FRAC CAT-CCHAR	0.00
WT FRAC CAT-FCHAH	0.00
#CAT LOSS THIS RUN	0.00
TOC-DEMISTER(PPM)	6840. 00
STEAM/BAGASSE RATIO	.58
DRY MOL WT	19.50
WET MOL WT	19.43
N2-02 FREE MW	19.40
ELEC INPUT(KWH/HR)	32.67
(based on 7 elements)	
KWHR/LB DRY EAGASS	.49
ELECTRIC BTU/LB DRY	1655.83

CONVERSIONS & BALANCES

HEAT GAS/HEAT BAGASS	.74
GAS BTUS/LB BAGASS	5214

CARBON CONVERSION

TO GAS(%)	76.51
TO SOLID(%)	22.97
TO LIQUID(%)	.74
SUM CONVERSIONS	100.22

METHANOL #/#BAGASSE .306129945273 TO .344792804934
THEORETICAL ASH IN CHAR(%) 37.00

DESIGN INFORMATION

BED GAS VELOCITY	1.67
CART HEAT COEFF	118.69

APPENDIX B

ECONOMIC EVALUATION

APPENDIX B

ECONOMIC EVALUATION

Figures B.1 and B2 present the results of the economic evaluation in graphical form. Also included is an example of the output from the economic evaluation program.

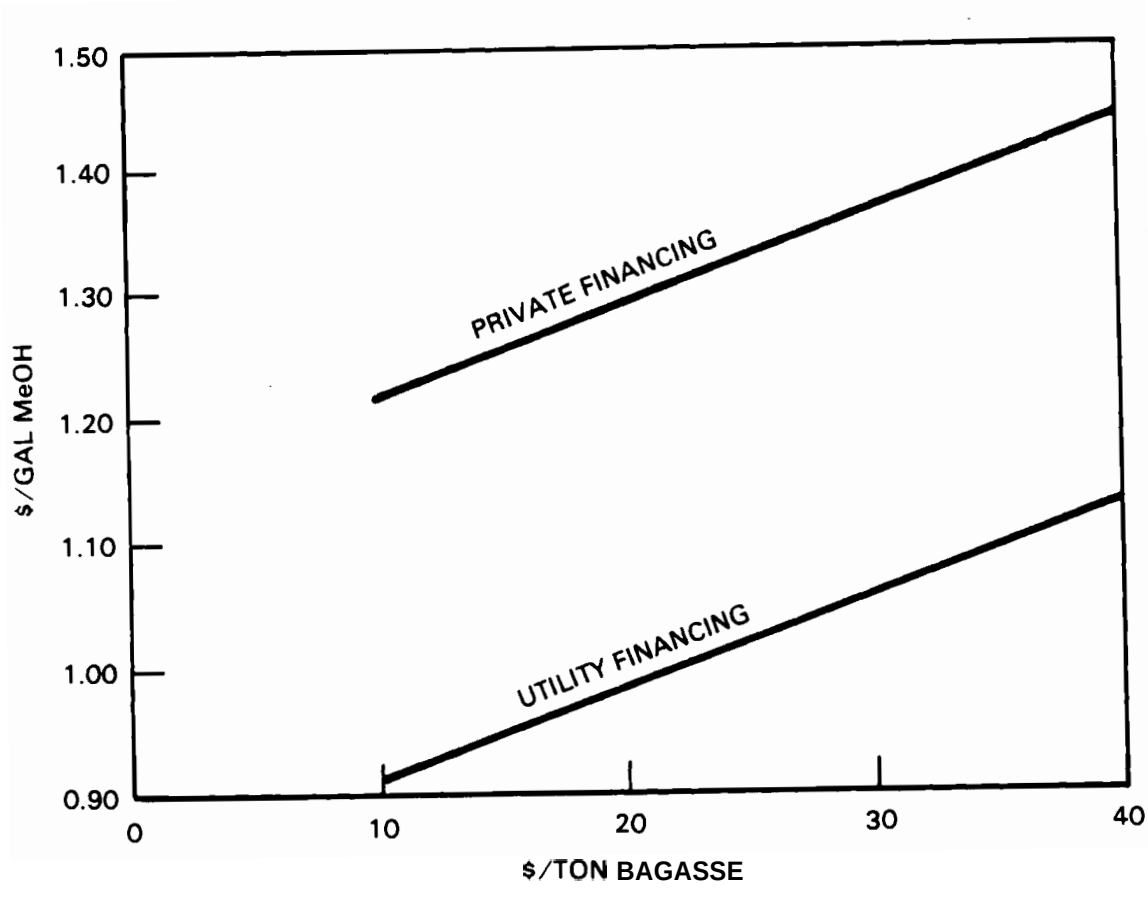

FIGURE B.1. Required Methanol Selling Price for a Plant Feeding 800 ton/day Dry Bagasse Fibre

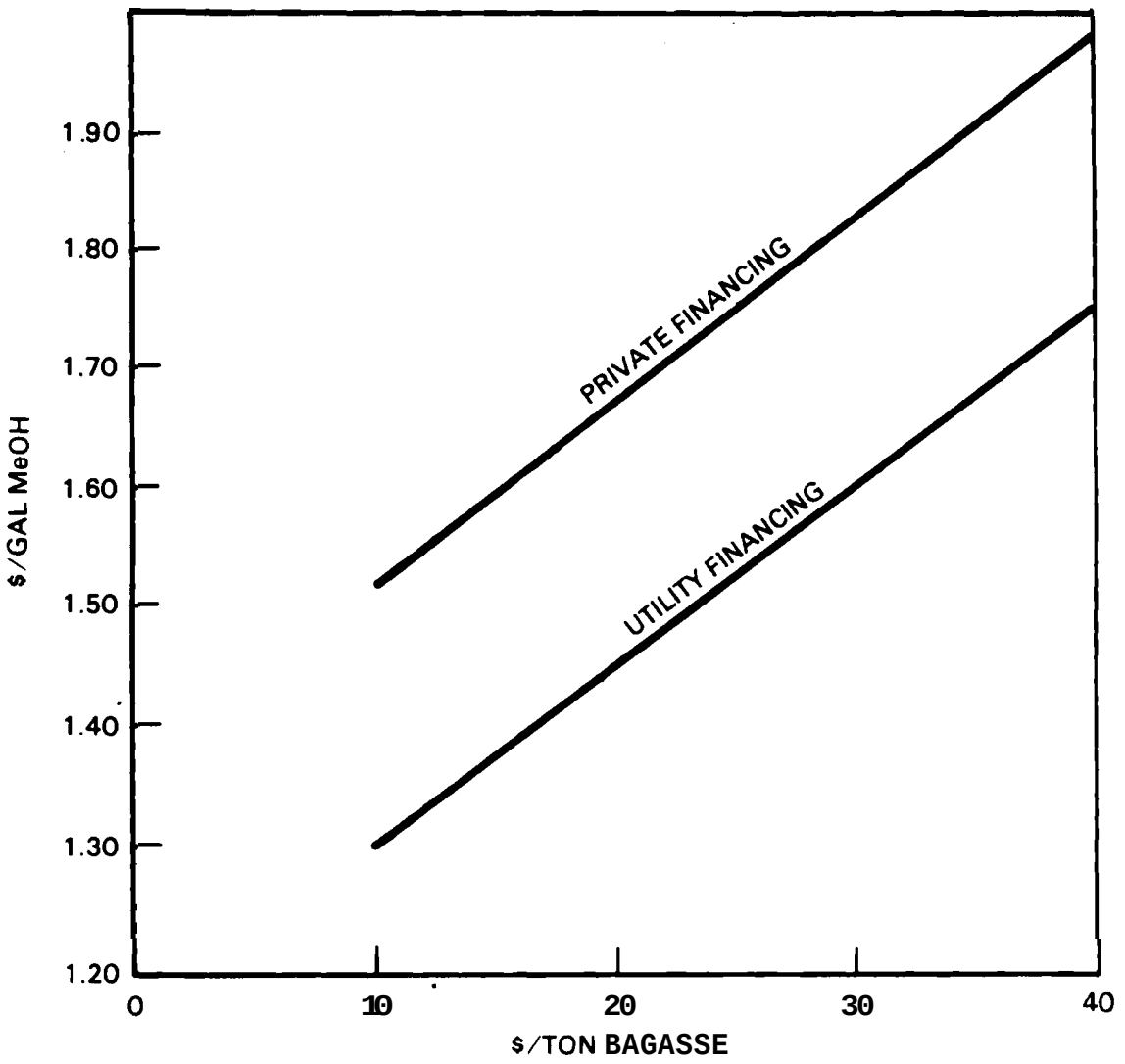

FIGURE B.2. Required Methanol Selling Price for a Plant Feeding 200 ton/day Dry Bagasse Fibre

300TPD BAGASSE TO MEOH
NEWPORT OREGON

TOTAL PLANT-DIRECT COST SUMMARY
AUG 83

CE PLANT COST INDEX FOR PRESWT MNTH 315.9
CE CONSTRUCTION INDEX = 270.2

CODE	DESCRIPTION	GRAND TOTAL $/1000	DIR PO MAT'L TOTAL $	SUBCONTRACT MATERIAL 3	SUBCONTRACT LABOR MAN HRS	SUBCONTRACT LABOR RATE	SUBCONTRACT LABOR TOT $	SUBCONTRACT GRAND TOTAL	DIRECT HIRE LABOR MAN HRS	DIRECT HIRE LABOR RATE	DIRECT HIRE LABOR TOTAL $
						38.79				16.62	
101	BAGASSE STORAGE&PREPERATION	20114	1226	9494	154.73		6002	15496	204.08		3393
102	BAGASSE DRYING	5305	0	3242	53.18		2063	5305	0.00		0
103	GASIFICATION	14078	8764	958	40.60		1575	2533	167.32		2782
104	SHIFT CONVERSION	565	485	9	0.79		30	40	7.27		121
105	ACID GAS REMOVAL	3528	0	1957	40.50		1571	3528	0.00		0
106	COMPRESSION	3523	2760	36	2.38		92	128	38.21		635
107	MEOH SYN &DISTILLATION	9894	9894	0	0.00		0	0	0.00		0
108	PURGE GAS REFORMING	497	373	12	1.18		43	55	4.18		69
109	WASTE WATER TREATING	1942	409	800	11.24		436	1236	17.89		297
110	RAW WATER TREAT & CLNG WATR	1627	683	374	6.92		268	643	18.09		301
111	BOILERS & BFW SYSTEM	2827	608	1201	16.24		630	1831	23.37		388
112	MISC. UTILITY SYSTEMS	1015	583	19	0.92		36	55	22.72		378
113	STR & LOG-PROD & UTILITIES	1185	120	641	8.22		319	960	6.31		105
	TOTAL	66100	25824	18743	336.82		13065	31808	509.43		8469

TOTAL MAINTENANCE = .078*DIR COST 5155.884 CAPACITY RATIO = 4 CE EQP IX 1.187594 CE LBR IX

EXECUTIVE SUMMARY
BAGASSE TO METHANOL

	COST M$	% TIC
EQUIPMENT		b.00
DIRECT PURCHASE MATERIAL	25823.72	26.49
SUBCONTRACT MATERIAL	18742.79	19.23
LABOR 336.8217M-MHRS	13064.86	13.40
DIRECT HIRE LABOR 509.4338M-MHRS	8468.679	8.69
		a.00
S/T DIRECT EQUIP COSTS (D.E.C)	66100	67.80TEMP VAL
FIELD INDIRECTS @ 21% D.E.C.	13881	14.24
PRO-SERVICES	15949	16.36
OTHER	1559	1.60
		0.00
		0.00
T.I.C SEED !!!	97489	100.00
ESCALATION PERCENT 9	8774	9.00
		0.00
ESCALATED T.I.C.	106263	109.00

EXCLUSIONS:

TOTAL CAPITAL REQUIRED	$/1000	
TOTAL INSTALLED COST (T.I.C.)	97489	
FUNDS DURING CONSTRUCTION (T.I.C. * 1.25YRS * .09)	10967	
START-UP COSTS (20% TOT ANN GROSS OP COSTS)	4005	
WORKING CAPITAL		
14 DA RAW MATL INVENTORY	112	
MATLS&SPLYS @ 9% T.I.C.	877	
NET RCVBLS @ 1/24 ANN REVENUE	1408	TEMP VAL
TOTAL CAPITAL REQD	114858	114858

ANNUAL DIRECT OPERATING COSTS

COST COMPONENT	ANNUAL USE	UNITS	$/UNIT	$1000/YR
RAW MATERIAL				
BAGASSE	264000	DRY TON	10	2640
UTILITIES				
WATER	210609	M GAL	.5	105
ELECTRICITY	70400000	KWHR	.03	2112
DIESEL FUEL	43560	GAL	1	44
CATALYSTS & CHEMICALS				
GASIFIER CATALYST	9120000	LBS	.2	1824
SHIFT CATALYST	60	FT**3	107	6
CL GUARD CATALYST	1320	FT**3	151	199
S GUARD CATALYST	1000	FT**3	75	75
MEOH CATALYST	18000	#????	8	144
REFORMER CATALYST	56	FT**3	235	13
LABOR				
PROCESS OPERATING	44	MAN YRS	10.7/HR	373
MAINTENANCE @ 60% TOT MAIN				3093
SUPERVISION @ 20% P.OP&MNT				813
ADMINISTRATIVE & OVERHEAD				
@ 60% OF TOTAL LABOR				2928
SUPPLIES				
OPERATING @ 30% PR OP LABR				292
MAINT @ 40% TOT MINT CST				2062
TAXES AND INSURANCE				
@ 2.77% OF T.I.C.				2700
TOTAL GROSS OP COSTS/YR				20026

METHANOL SELLING PRICE $/GAL		$/GAL	GAL/YR	EQUITY RETURN	DEBT INT %	UEPREC %	DEPREC YEARS	FED TAX
UTILITY FINANCING	SEED!!!!	0.91	37148000	15	10	.05	20	48
PRIVATE FINANCING		122		12				48

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UC-61D

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