The nominal feed gas rate for this unit is 1500 SCFH and this is supplied from a tube trailer via feed gas compressors. The reactor is 3.62" I.D. x 84" and the catalyst bed height can be varied from two to five feet. In six months of operation, through June 30, 1974, essentially all objectives of this phase of the development program were accomplished. Three phase fluidization data were obtained and correlated. This included the effect of liquid and gas velocities, catalyst particle size, and physical properties of both the liquid and gas phases. A gamma ray detector was utilized to measure actual bed densities under a variety of conditions.

Experimental runs were performed using Harshaw, CCI and Calsicat catalysts. Process variable scans were conducted and the kinetic model developed from bench scale results was corroborated. At the beginning of June, a continuous run was started. Plans were to obtain at least 600 hours of continuous operation. By June 30, 1974, 280 hours of on-stream time was acccumulated. Both catalyst activity and liquid stability were being monitored.

While the PDU program was proceeding, engineering contractors were being screened and evaluated for construction of the LPM pilot plant. In January, 1974, a subcontract for detailed engineering, procurement and construction of the pilot plant was awarded to Davy Powergas, Houston Engineering Division. This unit is being designed for a synthesis gas feed rate of 2 MM SCFD at 1000 psig. The LPM reactor is 2' O.D. x 15'. The synthesis gas feed is set by the maximum flow rate from the Institute of Gas Technology HYGAS plant located in Chicago, Illinois. The LPM pilot plant is being constructed as a skid-mounted unit. This will allow the decision on exact location to be deferred as late as posisble. In addition to the IGT pilot plant, the CO_2 Acceptor plant at Rapid City, South Dakota is also being considered. The design of the LPM pilot unit is compatible with the conditions there. The synthesis gas feed rate from the CO_2 Acceptor plant is 0.5 MM SCFD available at 100 psig. A feed gas compressor can boost one-third of the gas to 300 psig. It would be possible to test the unit at both locations or possibly other coal gasification pilot plant sites.

The objectives of the LPM pilot plant are to test the liquid phase methanation process with a synthesis gas feed actually produced in a coal gasification process and provide sufficient technical and engineering data to allow design of commercial sized (ca. 250 MM SCFD) plants.

Economic studies were completed during this reporting period based on experimental results obtained from both the bench scale unit and the PDU. In October, 1973, the LPM process was compared economically with IGT's recycle feed quench process. A pressure optimization study was completed shortly thereafter and strongly indicated the need for data at lower operating pressures, 300-500 psig. Recently, the economic justification for a polishing reactor following the LPM reactor was documented. An economic comparison of the LPM process with Lurgi's hot gas recycle process was also recently completed. This was based upon synthesis gas feed being produced in Lurgi gasifiers as proposed for El Paso's Coal Gasification Complex in Burnham, New Mexico.

Summary of Interim Report No. 3
Period of Performance, July 1, 1974 - June 30, 1975

Bench Scale Work

Experimental work in the 0.810" I.D. bench scale unit continued throughout the interim period. Complete process variable scans were performed on a variety of catalysts and process liquids. Several commercial and modified catalysts were tested including:

LaPorte-Davidson	CRG-A-R-11010
LaPorte-Davidson	Modified CRG-A-R-11020
Calsicat	Ni-230S
Harshaw	Ni-3266E
Engelhard	4037J

Various sizes and shapes of these catalysts were also investigated. The hydrocarbon liquids tested included:

Witco 40 Mineral Oil
Exxon 150 Aromatic Oil (C_9-C_{12} Polymethyl Benzenes)
Xylene-Biphenyl Mixture

The process variable scans were done to determine the effect of all variables on reaction rate and system productivity. From this work, the previously derived kinetic model was modified to incorporate terms for the H_2/CO molar feed gas ratio, thus, extending the range of the model to encompass compositions found in the combined methanation/shift reaction (from $3H_2$/CO to $1H_2$/CO feed ratios).

The effect of catalyst size was studied on the bench scale unit in the CRG-Witco 40 system. As expected for a first order reaction limited by pore diffusion, the reaction rate constant was inversely proportional to particle size. Compared to an equivalent sized Calsicat Ni-230S particle, the CRG-A catalyst was found to be about 20 percent less active. Two tests with Harshaw Ni-3266E showed sharp daily declines in catalyst activity. Simultaneously, liquid degradation and loss rates were monitored. Loss rates up to two times greater than found in the PDU were experienced. Systematic study of the problem led to the conclusion that the bulk of the oil lost is in the form of a fine micron mist formed during the rapid cooling and condensation of the product gas stream. The addition of a high pressure condenser and separator virtually eliminated this problem with all remaining losses attributable to mechanical losses associated with leakage through the pump seals.

During October through December, CRG-A (1/16" extrudates) were tested with Exxon 150, a xylene-biphenyl mixture, and Witco 40 mineral oil. Very high catalyst activities were obtained with the aromatic oil. Initial tests at 2:1 and 1:1 H_2/CO ratios (combined methanation/shift) were also promising.

During January, 1975, combined shift/methanation studies were performed on CRG-A/Witco and Calsicat Ni-230S/Exxon 150 catalyst-liquid pairs. In addition to standard process variable scans, the effect of direct water injection was investigated. The results indicated only a small effect when the H_2/CO ratio was above 1.0, even at water addition costs up to 0.14 moles H_2O/mole dry gas. Reaction rate studies on a tabulated CRG-A (1/8"Ø x 1/8") catalyst indicates almost an order of magnitude lower reactivity than the more porous extrudate form. In March, a modified CRG-A extrudate was tested with Witco 40 mineral oil and found to yield high activity levels with no deterioration and an activation energy of 18,400 cal/gm mole. During May, this system was further investigated at a $2H_2/CO$ and 0.95 H_2/CO feed gas ratio. Effluent gas compositions were carefully monitored to determine a typical analysis for polishing reactor studies. It was found that, for feeds lower than $1H_2/CO$, the reaction composition becomes hydrogen deficient at about 90 percent conversion, thus, requiring direct water injection.

Process Development Unit

A one month continuous run was completed on July 14, 1974 in the Process Development Unit (PDU). The Calsicat Ni-230S/Witco 40 system demonstrated its ability to maintain an acceptable activity to an extrapolated life of 8000 hours. A novel catalyst regeneration procedure was also tested and proven during this run. After completion of the life study, the Witco 40 was replaced with Exxon 150 and a process variable scan indicated that a 10 percent higher reactor productivity can be expected with this process liquid.

Life Test Studies

A bench scale life unit was set up for automatic long-term catalyst tests. Two successful runs of 1200-1400 hours duration with CRG-A catalyst and both Witco 40 and Exxon 150 oils were made during the year. These runs also verified long-term aspects of the LPM process.

Polishing Reactor Studies

Some preliminary reaction studies were initiated with a bench scale polishing reactor on a LPM reactor product from a $3H_2/CO$ mole ratio nominal feed gas. When the LPM product gas was passed through a high pressure condenser and separator in order to remove the product water and organic vapors prior to entering the polishing reactor, the final product met substitute natural gas requirements.

Pilot Plant

While the experimental program was proceeding, work on the engineering and construction of the LPM pilot plant was going ahead. By September, 1974, Davy Powergas, the engineering subcontractor selected for the project, had completed more than 95 percent of the detailed engineering. Most of the major equipment items were purchased. During November, Industrial Contractors, Texas City, Texas was selected as the skid fabricator and fabrication was begun in January, 1975. By the end of the interim period, the target completion date for the pilot plant was being held at July 15, 1975 although non-delivery of seventeen block valves by June 30 made it unlikely that the unit would be completed on schedule.

Summary of Work Performed During Final Contract Period, July 1, 1975 - September 30, 1976

Bench Scale Work

For simplicity this section includes a summary of both LPM and LPM/S work done on the bench scale unit in subsection (1). Subsections (2) and (3) include work performed on the bench scale life unit and polishing reactors, respectively.

1. Bench Scale Unit

During the third quarter 1975, the LPM bench scale unit examined the Calsicat Ni-230S (1/16" spheres)/ Witco 40 mineral oil system. Catalyst activity was equilibrated with a $3H_2/1CO$ feed gas. The reaction rate constant was determined to be 2.1×10^{-6} gm mole/gm cat-atm-sec at 650°F. Subsequent work with lower H_2/CO ratio feed gases had to be suspended due to mechanical problems with the unit. Leakage caused repeated depressurization of the reaction system while on standby overnight. The unit was shut down and the system piping was completely overhauled to insure a leak-free system. Following repairs, a new batch of Calsicat catalyst was loaded and reduced and a process variable scan performed. After the catalyst activity equilibrated with a $2H_2/CO$ feed gas, the feed gas composition was changed to a 1.4 H_2/CO ratio. The catalyst activity lowered somewhat probably due to the slower rate of the water gas shift reaction, limiting the methanation reaction. Even at high CO conversions, the effluent analysis indicated a H_2/CO ratio of less than 1, which is unsuitable as a polishing reactor feed.

2. Bench Scale Life Unit

In the early half of 1975, a bench scale life unit was set up for automatic long-term catalyst tests. Two successful runs of 1200-1400 hours duration with CRG-A catalyst and both Witco 40 and Exxon 150 oils were made and have been reported in Interim Report No. 3. These runs verified long-term aspects of the LPM process. The second of these runs was prematurely terminated in July, 1975 when the temperature controller failed in the energized state raising the reactor temperature to over 500°C, causing the reactor to plug. This eventually led to a ruptured line as the circulating oil pump continued to function. Repairs were made and the system was put back on-stream during the month of August. After 310 hours on-stream, the test was terminated due to another plug in the reactor inlet. A new test (Calsicat Ni-230S/Witco 40) was started and logged about 600 hours on ≥ $2H_2/1CO$ feed gas. The kinetic rate constant equilibrated at a value of 0.41×10^{-6} gm moles/atm-gm cat-sec.

During October, the life test unit was operated on a 1.40/1 H_2/CO feed gas. The activity level of the Calsicat Ni-230S/Witco 40 system was consistent with previous results. However, the run was terminated after 550 hours due to a plug in the reactor inlet. The probable cause of the plug was overheating of the liquid phase in a dead zone below the gas inlet. The reactor was modified to allow premixing of the gas and liquid streams. A second run on a 1.4 H_2/CO feed gas was terminated after 235 hours due to a power failure. A backup temperature limit control system was installed to prevent a recurrence of this problem. A 760 hour run was completed using a Calsicat Ni-230S catalyst and Witco 40 mineral oil in the first quarter, 1976. The feed gas contained a H_2/CO ratio of 1.4/1. The catalyst activity stabilized at a value somewhat higher than that obtained in previous runs with the same catalyst. This was attributed to modifications made in the system design. This completed work in the life test unit and it was shut down.

3. Bench Scale Polishing Reactor

During the third quarter, 1975, the polishing reactor was put on-stream with CRG-A 1/8" x 1/16" catalyst using the following feed gas composition:

CO	1.9%
H_2	14.0%
CO_2	21.1%
CH_4	63.0%
	100.0%

This run was terminated after 1000 hours. At reaction conditions of 300°C salt bath, 500 psig, and a feed gas VHSV of ~ 11000, the hot spot temperature, located 1 cm into the 6 cm bed, slowly fell to 370°C from its initial value of 390-400°C. The product gas composition had remained constant at 77-79 percent CH_4 and 21-23 percent CO_2 with no detectable quantities of H_2 and CO.

A second polishing reactor life test was terminated as planned after being on-stream for 1000 hours. The product gas composition remained constant over the range 60-61 percent CH_4, 39-40 percent CO_2 with trace H_2 and CO. The process conditions during the run were:

Salt Bath Temperature	300°C
Hot Spot Location	1 cm into 6 cm bed (constant)
Initial Hot Spot Temperature	372°C
Hot Spot Temperature after 620 hours	358°C
Final Hot Spot Temperature	341°C
Pressure	500 psig
Vapor hourly space velocity	9000 hr^{-1}

Following this run, the polishing reactors were modified to allow down-flow operation with steam addition.

During the first quarter 1976, each of the two polishing reactors completed a successful 1000 hour run. Both reactors were loaded with CRG-A, 1/8" dia. by 1/16" catalyst. The process conditions during the run were:

Polishing Reactor	A	B
Gas Flow	Down	Down
Dry Feed Gas Composition, Vol. %		
H_2	6	8
CO	3	2
CO_2	36	24
CH_4	55	66
	100	100
Steam Injection, Volume % of Feed Gas	15%	0%
Salt Bath Temperature °C	313	313
Pressure, psig	500	500
Vapor Hourly Space Velocity, $hr.^{-1}$	10,100	9,600
Hot Spot Temperature °C	333	337-343
Hot Spot Location, cm. into 6 cm Bed	3	2-3
Effluent Composition	38% CO_2; 62% CH_4 Traces H_2 + CO	26-28% CO_2 72-74% CH_4

Upon completion of the run, the catalyst was removed and returned to the catalyst supplier, Grace Research Division, for detailed analysis and evaluation.

All four 1000 hour duration runs indicated a constant reactor performance throughout. However, only in the run where a low level of steam (~ 0.1 mole steam/mole dry gas) was used was there virtually no carbon laydown on the catalyst. This completed work in the polishing reactor unit and it was shut down.

C. Process Development Unit

During the second half of 1975, the Process Development Unit was not operated. It was overhauled and modified to allow liquid phase methanation/shift operation.

The PDU was started up in February, 1976. The first system investigated was a $2H_2/1CO$ feed gas with the Calsicat Ni-230S/Witco 40 mineral oil system. A process variable scan was completed and the results were entirely consistent with previous work in both the bench scale and process development units. In addition to the excellent results obtained, no make-up process oil was required during the duration of the process variable scan. This was attributed to the modifications made in the PDU system; the addition of a demister system to the product gas, and a double mechanical seal closed loop, seal oil system.

Additional runs were made during April. Synthesis feed gases having a H_2/CO molar ratio of 1.02 and 1.40 were investigated both with and without the direct injection of steam. For these series of runs, a reduced and stabilized form of the Calsicat Ni-230S catalyst was used as received, without any further activation procedure. The activity equilibrated within the first 40 hours on-stream at a lower than normal level ($\sim$25% less). This is believed due to the inability of all the stabilized catalyst sites to revert to their active state while in the presence of the liquid phase (Witco 40 mineral oil). The effluent gas composition was monitored, especially with respect to the molar H_2/CO at the high conversion levels, 90-99 percent, and found to be consistent in all respects with the results obtained in the smaller bench scale unit. For both feed gases, the addition of steam mildly increased the catalyst activity, while markedly increasing the effluent H_2/CO ratio.

This completed the program in the PDU and the unit was shut down, cleaned, and placed on stand-by.

D. Pilot Plant

In July, 1975, a meeting was held at Industrial Contractors, Texas City, Texas with representatives of Chem Systems, Institute of Gas Technology, Conoco, C.F. Braun, ERDA and Davy Powergas. The LPM process was reviewed with respect to the pilot plant design and how it will fit in with the existing gasification facilities. At a subsequent meeting between Chem Systems and Davy Powergas, the pilot plant completion date was projected for August 22, 1975 due to a delay in delivery of 17 Walworth valves.

The pilot plant construction continued during August. Industrial Contractors made considerable progress but did not complete the job on August 22 as promised. Seventeen Walworth valves did not arrive as scheduled and expediting proved difficult. Radiographic random inspection of piping welds was completed by Industrial Contractors. An outside testing laboratory was contacted to have an independent verification of the welding job.

The valves were finally received in early October and installed. Chem Systems arranged to have a 100 percent X-ray analysis performed on all of the girth butt pipe welds. Chem Systems' personnel were on-hand when the X-ray tests were initiated.

Radiographic inspection by Consolidated X-Ray Service Corporation of all piping butt welds on the LPM pilot plant was completed by the end of January, 1976. Approximately 40 percent of the 423 welds were classified as defects under the 100 percent X-ray examination criteria according to the ANSI-B-31.3 code. Davy Powergas and Industrial Contractors were shown the films.

Chem Systems completed a preliminary inspection of the LPM pilot plant and informed Davy Powergas of the deficiencies uncovered. A meeting was also held with the ERDA/AGA Operating Committee to update them on the overall situation, with specific attention focussing on the radiographic inspection of the welds.

During the second quarter 1976, negotiations were finalized with Davy Powergas on completing the LPM pilot plant. Work to be completed included ultrasonic testing of socket welds, repair and X-ray of all girth butt welds on pressure piping to 100 percent radiographic standards, hydrostatic testing of the unit, reinsulation, break-down and preparation of the component skids for shipping, and arrangement for crating and transport to a still-to-be designated coal gasification pilot plant site. The agreement with Davy Powergas also included the financial responsibilities of both parties. The agreement was submitted to ERDA and AGA for approval. Plans were made to begin the weld repair work while awaiting ERDA/AGA formal approval.

The ERDA/AGA Operating Committee approved the agreement reached between Chem Systems and Davy Powergas on weld repair and completion of the LPM pilot plant skid. A meeting was held on June 4 at the skid fabricator's shop in Texas City, Texas. The X-rays and back-up data taken by Consolidated X-Ray Corporation were loaned to Davy Powergas for their use in identifying and repairing defective welds.

The weld repair program which started on June 7 proceeded slowly due to several factors; a one week closure for annual vacation at the fabricator shop, some inclement weather which prevented outdoor work on the LPM skid, and lack of sufficient welders. Also, repair of 4" welds, which was done "in place" on the skid, was hampered by excessive rust within the piping, causing numerous repaired welds to fail to meet 100 percent radiography standards. A revised cleaning and welding procedure was adopted to overcome this problem and by August 6 all welds had been repaired and passed radiographic inspection.

Several vendors were interviewed during July with respect to their interest in supplying services for (a) preparing the pilot plant skid sections for shipping, (b) rigging of the skids onto trucks, (c) shipment from Texas City to the designated coal gasification site and (d) chemical cleaning of the piping and equipment.

Satisfactory quotations for crating, rigging, and truck shipment of the LPM pilot plant skids to one of the proposed demonstration sites were received by Chem Systems. Quotation for chemical cleaning of the unit was also received. The vendor recommended cleaning at the final destination after reassembly.

The LPM pilot plant successfully passed a hydrostatic test on August 10, witnessed by Chem Systems. The unit was inspected and certain minor deficiencies were noted to Davy Powergas. The major item was revision of the P & I drawings to reflect changes, additions,; i.e. the "as-built" conditions. Davy Powergas agreed to do this and transmit the revised drawings to Chem Systems by the first week of September. Reinsulation of the piping started on August 16 and was completed by the end of the month.

A thorough inspection of the pilot plant was carried out by Chem Systems' personnel on September 9-10, 1976. Other than some minor details which were corrected immediately, two items were found incomplete and will be field installed, namely:

1) Reactor level detector lift mechanism - parts to be shipped with pilot plant.

2) Reactor oil pump seal flush piping system - parts to be shipped with pilot plant and welded at demonstration site.

Disassembly of the LPM pilot plant in preparation for shipment to the designated demonstration site was begun on September 15 and, by September 20, ICI Corporation had finished its disassembly work tasks.

Five low-boy trailers were delivered to ICI Corporation, Texas City, on September 20. The rigger (Westheimer Rigging & Heavy Hauling Company) began work on the same day, and by September 22, all skid sections and the control house had been loaded on trailers. The crater (First Class Export Crating, Inc., Houston, Texas) began meaningful crating activities on September 21. By September 28, this work was completed. With Chem Systems personnel present on September 29, C&H Transportation Company truckers removed the five trailers from the ICI Corporation yard and the pilot plant was on its way to the Institute of Gas Technology's HYGAS Plant in Chicago, Illinois, designated by ERDA and AGA as the first demonstration plant site.

This completed all work on the construction of the LPM pilot plant, the last remaining task outstanding under contract E (49-18)-1505.

E. Process and Economic Evaluation Studies

In early 1976, conditions of the LPM/S polishing reactor design were compared with other commercial (SNG naphtha gasification plants) and proposed (RMP Process) final stage methanators. All systems lie outside the carbon forming region. The LPM/S polishing reactor designs had basically the same carbon, hydrogen and oxygen content as the other cases. Therefore, the LPM/S system should produce no more carbon than that found in other systems since the same catalysts would be employed.

During the second quarter, 1976, a detailed evaluation of the LPM/S process was completed. The study includes a discussion of several coal gasification processes and an arbitrary grouping of processes according to the H_2/CO ratio of the synthesis gas produced in the coal gasifier. For purposes of evaluation, four categories were established:

	Example Process
$H_2/CO > 3$	CO Acceptor
$2 < H_2/CO < 3$	Lurgi
$1.4 < H_2/CO < 2$	Synthane
$H_2/CO < 1.4$	Koppers-Totzek, Bi-Gas

The first group represents processes where no separate shifting is required and the LPM process can be used by itself. The second group requires some shifting, but as regards LPM/S, no separate steam injection is necessary. The third group encompasses those synthesis gas feeds where a small quantity of supplemental steam is injected into the LPM/S reactor. The fourth group involves those processes requiring substantial shifting. For this category it was determined that it is more economical to shift a portion of the gas in a conventional manner in order to achieve a gas feed to the LPM/S reactor which would then fall back into the third grouping above.

The economics of the LPM/S process were developed for two typical cases rather than attempting to study application to each and every coal gasifier. The first case was for a synthesis gas feed with a hydrogen to carbon monoxide ratio of 2/1 which also contained a substantial quantity of methane. The Lurgi and Synthane processes would be typical examples. The second case was for a synthesis gas with a hydrogen to carbon monoxide ratio of 1/1 and which contains little methane.

Process flowsheets, heat and material balances, and equipment specifications were prepared for both cases for a commercial-sized plant producing 250 MMM BTU/D of SNG. From this information, total capital investment was estimated and annual operating costs calculated.

The following table summarizes the evaluation for the two cases:

	Case 1	Case 2
Synthesis Gas Feed Composition, mol%		
H_2	54.0	49.8
CO	27.0	49.8
$CH_4 + CO_2$	19.0	0.4
Total	100.0	100.0
Total Fixed Investment (including initial catalyst and process liquid)	$29,800,000	$59,600,000
Private Financing Equivalent	$MM/YR	
Capital Related Charges @ 40%	11.92	23.84
Catalyst and Oil Replacement	2.97	7.10
Utilities	(16.57)	(33.84)
Annual Operating Cost	(1.68)	(2.90)
Utility Financing Equivalent		
Capital Related Charges @ 20%	5.96	11.92
Catalyst and Oil Replacement	2.97	7.10
Utilities	(16.57)	(33.84)
Annual Operating Cost	(7.64)	(14.82)

A simplified approach was used to estimate capital related charges. Two methods have been recommended by C.F. Braun in their "Coal Gasification Commercial Concepts Gas Cost Guidelines." In the utility financing method, the annual capital-related charges are equivalent to 20-25% of the total fixed investment including initial catalysts and chemicals. In the private investor financing method, with an after-tax DCF rate of return of 12 percent, the annual capital-related charges are equivalent to 40-50 percent of the total fixed investment. The evaluation of the LPM/S process was done for two cases; capital charges at 20 percent of total investment and capital charges at 40 percent of total investment, essentially equivalent to utility-financing and private investor financing respectively.

Overall utility charges were estimated by assuming that part of the 900 psia steam generated within the LPM/S process would be used to satisfy power requirements for the process and the remainder would be exported and utilized in the rest of the coal gasification complex. A credit of $2.00/Mlb was applied towards the net steam generated. Because of the large quantity of surplus steam generated, the credit is more than enough to off-set catalyst and oil replacement costs and all capital-related expenses regardless of which financing method is used.

For Case 1, with the $2H_2/CO$ synthesis gas feed, the LPM/S process has an annual operating credit of $1,680,000 (2.0¢/MM BTU of SNG) using a private investor financing approach and an annual credit of $7,640,000 (9.3¢/MM BTU of SNG) using a utility financing approach.

For Case 2, where all of the SNG is producted in the LPM/S reaction, since the $1H_2/CO$ synthesis gas feed contains no methane, the total investment is considerably higher than Case 1. However, a much larger quantity of steam is produced which more than offsets the higher capital investments. Using a private investor financing approach, the annual operating credit is \$2,900,000 (3.5¢/MM BTU of SNG) while for a utility financing approach the annual credit is \$14,820,000 (18.0¢/MM BTU of SNG).

A pressure optimization study was performed for the LPM/S process with a $2H_2/CO$ synthesis gas feed. It was assumed that the synthesis gas was available at 100 psia and product SNG required at 1000 psia. The study showed that optimum operating pressure for the LPM/S process encompassed a broad range of 250 to 600 psia. This result was essentially equivalent to a previous pressure optimization study performed for the LPM process alone for a synthesis gas feed, previously adjusted to the proper 3/1 H_2/CO ratio. The conclusion of this study is that if the coal gasifier operates at pressures of 300 psia or higher, the synthesis gas should go directly to the LPM/S process without intermediate compression. The SNG product should then be boosted to final pipeline pressure.

Included in the evaluation is a comparison of the LPM/S process against conventional shift and methanation processes. A hot gas recycle process, as offered by Lurgi, was chosen as an example of conventional methanation. An intermediate case, that of conventional shift followed by the LPM process, was also included in the evaluation. This comparison was done for the two different synthesis gas feeds described earlier. The following table summarizes the comparison for the $2H_2/CO$ synthesis gas feed.

Again, two methods were used to evaluate capital related charges; private investor financing at 40 percent of total fixed investment and utility financing at 20 percent of total fixed investment. With capital charges at 40 percent, the LPM/S shows an advantage of 3.7¢/MM BTU over conventional shift with LPM and 14¢/MM BTU over conventional shift with Lurgi hot gas recycle methanation. With capital charges at 20 percent, the LPM/S shows an advantage of 9.0¢/MM BTU over Lurgi HGR and 3.1¢/MM BTU over conventional shift and LPM.

	LPM/S Only	Conventional Shift To $3H_2/CO$ Plus LPM	Conventional Shift To $3H_2/CO$ Plus Lurgi HGR
Total Fixed Investment (including catalyst and oil)	\$35,860,000	\$38,360,000	\$56,680,000
Annual Costs (Private Financing), \$M/Yr			
Capital Related @ 40%)	14,340	15,340	22,670
Cat. & Oil Replacement	2,970	2,170	1,160
Utilities	(14,240)	(11,320)	(9,090)
Annual Operating Cost	3,070	6,190	14,740
Cost in ¢/MM BTU	3.7	7.4	17.7
Annual Costs (Utility Financing), \$M/Yr			
Capital Related (@ 20%)	7,170	7,670	11,340
Cat. & Oil Replacement	2,970	2,170	1,160
Utilities	(14,240)	(11,320)	(9,090)
Annual Operating Cost	(4,100)	(1,480)	3,410
Cost in ¢/MM BTU	(4.9)	(1.8)	4.1

Again, two methods were used to evaluate capital related charges; private investor financing at 40 percent of total fixed investment and utility financing at 20 percent of total fixed investment. With capital charges at 40 percent, the LPM/S shows an advantage of 3.7¢/MM BTU over conventional shift with LPM and 14¢/MM BTU over conventional shift with Lurgi hot gas recycle methanation. With capital charges at 20 percent, the LPM/S shows an advantage of 9.0¢/MM BTU over Lurgi HGR and 3.1¢/MM BTU over conventional shift and LPM.

For the $1H_2/CO$ synthesis gas feed, the advantages of LPM/S and LPM with conventional shift over the Lurgi HGR with conventional shift are even greater,. The following table summarizes the comparison for this feed gas.

	LPM/S Only	Conventional Shift To $3H_2/CO$ Plus LPM	Conventional Shift To $3H_2/CO$ Plus Lurgi HGR
Total Fixed Investment	$66,060,000	$56,180,000	$82,730,000
Annual Costs (Private Financing), $M/Yr			
Capital Related (@ 40%)	26,520	22,470	33,090
Cat. & Oil Replacement	7,100	2,580	2,110
Utilities	(31,350)	(19,470)	(13,040)
Annual Operating Cost	2,170	5,580	22,160
Cost in ¢/MM BTU	2.6	6.7	26.6
Annual Costs (Utility Financing), $M/Yr			
Capital Related (@ 20%)	13,210	11,240	16,550
Cat. & Oil Replacement	7,100	2,580	2,110
Utilities	(31,350)	(19,470)	(13,040)
Annual Operating Cost	(11,040)	(5,650)	5,620
Cost in ¢/MM BTU	(13.2)	(6.8)	6.7

As shown above, the savings are over 20¢/MM BTU for the LPM/S process compared to conventional shift with Lurgi HGR.

The results of this process evaluation demonstration the economic attractiveness of the LPM/S process in commercial sized SNG complexes. The LPM/S process can be tailored to fit a wide variety of synthesis gases and system conditions where the feed H_2/CO ratio is less than 3. In all cases, LPM/S operates at lower annual costs than separate shift and methanation processes.