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LARAMIE ENERGY TECHNOLOGY CENTER

FINAL REPORT

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RECOVERY AND HYDROPYROLYSIS OF OIL ROM UTAH'S TAR SANDS

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For the Period of December 1982 - December 1984
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HOT WATER PROCESSING OF UTAH TAR SANDS

by

Jan D. Miller

Professor

Metallurgical Engineering

SUMMARY

During the past five years, new technology has been developed for fast flotation of Tar Sand with an air-sparged hydrocyclone. Conventional flotation in a stirred, aerated tank requires retention times on the order of minutes, whereas intrinsic bubble attachment times are on the order of milliseconds. This rate limitation for conventional flotation is due to the low probability of collision events, insufficient particle inertia to penetrate the bubble film and instability of bubble/particle aggregates. The design of the air-sparged hydrocyclone was envisioned to establish a controlled, high force field by swirl flow of the slurry to increase the inertia of fine particles and to produce, by introducing air through a porous cyclindrical wall, a high density of fine bubbles with directed motion orthogonal to the particles to improve collision efficiency. The net result is a flotation rate with retention time approaching intrinsic bubble attachment times. This
corresponds to a capacity on the order of 1-2 tpd/ft of cyclone volume, at least 50 times the capacity of conventional flotation cell of comparable volume.

In this regard, research was initiated to test the effectiveness of air-sparged hydrocyclone in the flotation of bitumen from digested tar sand slurry prepared according to the conditions and procedures that had been established from previous DOE sponsored research on the hot water processing of the Utah tar sands.
SUMMARY

Hydropyrolysis is a process for conversion of residual feedstocks to high yields of distillate products without the need for heterogeneous catalysts. Under certain conditions, over 99% of a tar sand bitumen is converted to valuable gaseous and liquid products. This study involves the design, construction and operation of a semi-continuous hydropyrolysis unit capable of quantitatively assessing the effect of process variables on product yield and structure.

The major effort in this study phase concentrated on the design and fabrication of a 2 liter/hour semi-continuous unit possessing both a gas recycle and a heavy oil recycle stream. A range of the process variables to be studied was determined (450-600°C, 1200-2200 psig, 1-30 sec. residence time, etc.) and various components were designed to provide the necessary process flexibility. Major design elements were the reactor, feed system, preheaters, separators, feed and product vessels,
compressor, gas system and safety features. The design criteria and details of design are specified in this report.

Hydropyrolysis reactions are known to perform best in the vapor phase at short residence times. To better understand the vaporization process and the physical behavior of a drop of oil when injected into a hot stream of hydrogen, a computer program was developed. This program, DROPT, considered the buoyant, drag, and gravitational forces on the oil drop, the mass transfer from the droplet and the heat transfer to the drop. Also considered was the boiling point distribution of the feed. Results revealed the concept of critical boiling point above which an oil drop would never fully vaporize and below which the drop would not remain in the liquid state. The model was used to specify a critical drop radius for which droplets smaller than the critical radius would be expected to fully vaporize while those above that radius would not completely vaporize under the given conditions. This information was used to design the reactor and nozzle system.

The system has been fabricated and is currently operational. Over 24 dozen runs have been made to test the system's performance and to refine the operating procedures and the system's configuration. Good progress has been made to achieve an operational system which provides a high level of process variable control.

Process simulation, economic and thermodynamic analyses have been completed for a 10,000 bbl/day process facility. Capital costs are estimated to be $6,000/daily-bbl capacity compared to
$4,000 for coking. Higher yields of high value product compensate for the higher capital costs. Economic analyses do not reveal any factor which would clearly make hydropyrolysis uneconomical. The thermal efficiency is calculated to be 93% while the 2nd-law efficiency is 53%. Details of these analyses are specified in this report.
THE FLUIDIZED BED PYROLYSIS OF BITUMINOUS SANDS

by

Francis V. Hanson

Associate Professor

Fuels Engineering

SUMMARY C

Fluidized bed pyrolysis of bitumen-impregnated sandstones to produce hydrocarbon liquids has been investigated using the bitumen-impregnated sandstone from the P.R. Spring deposit as feed-sand to the pyrolysis reactor. The P.R. Spring tar sand deposit is the second largest deposit in Utah, and is estimated to contain 4.0 - 4.5 billion barrels of bitumen in-place. Three distinct tar sand samples, P.R. Spring Rainbow II and P.R. Spring South, were used in this investigation because it was found that the physical and chemical properties of these three bitumens were quite different. The South Rainbow II bitumens were two orders of magnitude more viscous than the Rainbow I bitumen and the asphaltene content of the South bitumen, 82.8 percent, was 60% and 38% greater than the asphaltene content of the Rainbow I and Rainbow II bitumens, respectively. Process variables investigated included pyrolysis reactor temperatures and the feed sand retention times in the pyrolysis zone of the reactor.
Experimental data were obtained in a continuously fed, fluidized bed reactor operating at a temperature range of 773 - 923 K at solids retention times that ranged from 17-30 minutes. Reactor feed sand throughput was 2.25 kilograms per hour and the particle size of the feed-sand was 300-600 microns.

The hydrocarbon liquid product yield (Cg+) generally decreased with increasing reactor temperature. Concomitant with this decline in this liquid yield, there was an increase in the light gas (C1-C4) yield. Liquid product yield increased slightly with decreasing feedsand retention time while the light gas yield decreased. Coke yield was insensitive to process operating variables above 798 K, however, it did appear to be a function of the origin of the feedsand. A maximum liquid yield of 71 weight was obtained at a pyrolysis temperature of 798 K and a sand retention time of 20 minutes with the P.R. Spring Rainbow I tar sand. The maximum liquid yield attained with the P.R. Spring South tar sand was 55 weight percent at a pyrolysis temperature of 823 K and a sand retention time of 20 minutes and the maximum liquid yield attained with P.R. Spring Rainbow II tar sand was 61 weight percent at 798 K and a 20 minute sand retention time. At a fixed pyrolysis reactor temperature and sand retention time, the C5+ liquid product yield correlated quite well with the Conradson carbon residue and with the atomic hydrogen-to-carbon ratio of the native bitumen, that is, the liquid product yield increased with decreasing Conradson carbon residue and with increasing H/C ratio.

The quality of the bitumen derived liquids was significantly
improved relative to the native bitumen, that is, the viscosity was reduced by four to five orders of magnitude for the Rainbow I bitumen. The API gravity of the liquid product was 17.6° API, an increase of almost 10 °API relative to the native bitumen and the pour point decreased from 372 K (210° F) for the native bitumen to 278 K (40° F) for the bitumen derived liquid. The volatility of the liquid product increased from 31.9 weight percent to 77.2 weight percent relative to the native bitumen. The Conradson carbon residue and asphaltene content of the product liquids were substantially reduced.
A digital control system for the thermal processing unit for tar sands has been constructed and tested. This unit consists of three sections. In the first section, tar sands is fed into a fluidized pyrolysis reactor, where bitumen is cracked into coke, oil and gaseous products at temperatures of 480°C to 500°C. In the second section, coked sand is transferred by gravity from the pyrolysis reactor to the combustion reactor (a second fluidized bed), where the coked sand combusts with air to produce the energy required in the system. A potassium heat pipe is used to transfer energy from the combustion reactor to the pyrolysis reactor. In the third section, the oil from the pyrolysis reactor is collected through a series of condensers. A digital control system coupled to the reactor consists of an HP-1000 host computer, an HP 2250 interface, and several actuator and sensors.

Solids retention times in both reactors are controlled by two solids flow control valves. A digital PID algorithm plus
Valve linearization has been used to control the solids flow valves. The best control constants, $K_c = 80$, $T^\pi = 40$, $T^\iota = 0$ for the pyrolysis reactor, and $K_c = 50$, $T_i = 5$, $T_d = 0$ for the combustion reactor, were found experimentally. From the comparison of results, the digital PID algorithm plus valve linearization has much better controllability than the conventional pneumatic PI controller.

Temperatures in the pyrolysis reactor are controlled by the amount of energy transferred into the combustion reactor from the surroundings. An SCR power controller is used to regulate power into the heating wire wrapped around the reactor wall. The best PID algorithm controller constant, $K_c = 1000$, $T_i = 10$, $T_d = 0$ were found experimentally to control the thermal cracking of bitumen in the pyrolysis reactor. Pyrolysis of bitumen can be considered as only a disturbance to the system and its influence on the control system is insignificant.
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Hot Water Processing of Utah Tar Sands

Professor: Ian D. Miller
Visiting Professor: J. Hupka
Graduate Students: A. Cortes, J. Waters

Research activities in the hot water processing program have included evaluation of the air-sparged hydrocyclone technology in hot water processing for bitumen recovery from domestic tar sands, analysis of bitumen concentrate grade in the conventional hot water processing of type III tar sands, and efforts to prepare and program a detailed data base of the results obtained from the hot water processing research program since research began in the mid-seventies.

Air Sparged Hydrocyclone

During the past five years new technology has been developed for fast flotation with an air-sparged hydrocyclone. Conventional flotation in a stirred, aerated tank requires retention times on the order of minutes, whereas intrinsic bubble attachment times are on the order of milliseconds. This rate limitation for conventional flotation is due to the low probability of collision events, insufficient particle inertia to penetrate the bubble film and instability of bubble/particle aggregates. The design of the air-spared hydrocyclone (Figure 1) was envisioned to establish a controlled, high force field by swirl flow of the slurry to increase the inertia of fine particles, and to produce by introducing air through a porous cylindrical wall, a high density of fine bubbles with directed motion orthogonal to the particles to improve collision efficiency. The net result is a flotation rate with Al
a stable froth column at the center of the cyclone transports hydrophobic particles to the overflow

hydrophilic particles are rejected with the major portion of water through the underflow

air is injected through the porous cyclone wall where fine bubbles are formed

fine particles, requiring a high force field to float, are injected tangentially as a pre-conditioned slurry in water

Figure 1 Principal Features of the Air Sparged Hydrocyclone
retention times approaching intrinsic bubble attachment times. This corresponds to a specific capacity on the order of 50-100 tpd/ft of cyclone volume, at least 50 times the specific capacity of a conventional flotation cell of comparable duty.

In this regard, research was initiated to test the effectiveness of the air sparged hydrocyclone in the flotation of bitumen from digested tar sand slurry prepared according to the conditions and procedures that had been established from previous DOE sponsored research on the hot water processing of the Utah tar sands.\textsuperscript{4-10}

(Continued on page A4)
Initial efforts were directed toward the design and construction of the system which consists of the following:

* 5 gallon stirred tank digester
* 55 gallon sump
* 1\(\frac{1}{2}\) inch Galigher vertical sump pump
* air sparged hydrocyclone

Figure 2 presents a schematic diagram of the test circuit. The sump-pump-digester system is a modular unit, thus eliminating the need for a separate system of support for the digester and pump. A three-way valve at the pump discharge allows slurry recirculation to the sump or feed to the hydrocyclone. Heat for the digester is provided in the form of steam by a Chromalux steam generator.

The digester as originally designed could not hold the digested slurry and had to be redesigned, replacing the loose fitting lid with a 1/4 inch thick stainless steel plate securely fastened by eight 7/16 inch bolts arranged on the periphery of the digestion chamber. A six inch length of two inch diameter pipe, which passes through the top plate, served as the solids feed inlet. The agitator blades were cut to fit a five gallon vessel. These changes, although defeating the purpose of easy access to the digestion chamber, ensured the containment of digester contents under the more violent agitiation conditions.

Similarly, design modifications on the air sparged hydrocyclone were required to prevent choking of the adjustable underflow pedestal with the sand stream. This difficulty was overcome by using either a tangential underflow design or fixed diameter pedestals. The cyclone consists of two concentric cylinders, the inner cylinder being porous with an effective pore diameter of
25 microns. The cyclinders are of stainless steel construction with a wall thickness of 1/16 inch. A urethane Krebs cyclone header machined to 2-3/4 inches inside diameter accepts cyclone feed.

Initial experiments were run using an Asphalt Ridge feedstock with no reagent additions to establish the baseline performance of the air-sparged hydrocyclone. The experimental procedure was as follows:

1) turn on steam generator,
2) fill sump with fresh water (approx. 40 gal.),
3) start circulation pump and agitator,
4) start steam, sparge to sump,
5) charge digester with water (5.6 liters), add reagents,
6) start digester impeller and charge tar sand into digester (16 kg),
7) start steam to digester heating jacket.

After digestion was completed, the tar sand was discharged from the digester into the hot sump, whereupon any flotation reagents were added and the slurry was allowed to condition. At this point air sparge to the cyclone was started and slurry feed to the cyclone was initiated by throwing the three-way valve on the circulation pump. Overflow and underflow samples are taken using the stream sampling device mounted on the cyclone support frame. The initial experiments were run under the following conditions.

- **Reagent Addition:** none
- **Digestion Time:** 20 minutes
- **Digestion Temperature:** 92°C
- **Dilution Temperature:** 92°C
- **Pulp Density:**
  - Digester: 75% solids
  - Sump: 5% solids
Figure 2 Schematic Drawing of the Air Sparged Hydrocyclone Circuit
The results of selected experiments are presented in Table 1.

Table 1

Experimental Results for the Air Sparged Hydrocyclone
Flotation of Bitumen from Digested Asphalt Ridge Tar Sand

<table>
<thead>
<tr>
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<th></th>
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</tr>
</thead>
<tbody>
<tr>
<td>OF 1</td>
<td>79.44</td>
<td>2.53</td>
<td>0.3654</td>
<td>61.75</td>
<td>38.25</td>
<td>0.8822</td>
<td>0.2681</td>
</tr>
<tr>
<td>UF 1</td>
<td>20.56</td>
<td>16.98</td>
<td>0.6346</td>
<td>97.06</td>
<td>2.94</td>
<td>0.1178</td>
<td>0.7319</td>
</tr>
<tr>
<td>OF 2</td>
<td>81.82</td>
<td>2.33</td>
<td>0.3619</td>
<td>45.06</td>
<td>54.94</td>
<td>0.8711</td>
<td>0.2113</td>
</tr>
<tr>
<td>UF 2</td>
<td>18.18</td>
<td>18.49</td>
<td>0.6381</td>
<td>95.39</td>
<td>4.61</td>
<td>0.1289</td>
<td>0.7887</td>
</tr>
<tr>
<td>OF 3</td>
<td>74.36</td>
<td>2.56</td>
<td>0.3295</td>
<td>99.20</td>
<td>0.80</td>
<td>0.8789</td>
<td>0.2003</td>
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<tr>
<td>UF 3</td>
<td>25.64</td>
<td>15.11</td>
<td>0.6705</td>
<td>96.56</td>
<td>3.44</td>
<td>0.1211</td>
<td>0.7997</td>
</tr>
</tbody>
</table>

* The coefficient of separation (CS) is a single parameter evaluation of separation efficiency taking into consideration both the recovery and grade of product (bitumen concentrate). In fact, the CS represents the fraction of the feed which separates ideally and can be shown to be equal to the recovery of the bitumen in the concentrate minus the recovery of the sand in the concentrate.

In addition to these tests, the effect of underflow restriction was evaluated. As the underflow was choked, the volume fraction of the slurry reporting to the underflow was reduced and the effect on separation efficiency is indicated in Table 2.

Table 2

Effect of Volume Fraction to the Underflow on Separation Performance for Asphalt Ridge Tar Sand.
Air Flow = 28 scfm

<table>
<thead>
<tr>
<th>Volume Fraction of Slurry to Underflow</th>
<th>Coefficient of Separation</th>
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<tr>
<td>0.85</td>
<td>0.067</td>
</tr>
<tr>
<td>0.54</td>
<td>0.235</td>
</tr>
<tr>
<td>0.53</td>
<td>0.491</td>
</tr>
<tr>
<td>0.43</td>
<td>0.543</td>
</tr>
<tr>
<td>0.29</td>
<td>0.426</td>
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</table>
In summary some potential for the air sparged hydrocyclone technology in the flotation of bitumen from digested tar sand has been demonstrated, although its performance is not completely satisfactory. Further research and testing will be required to reach an acceptable level of performance. Additional design modifications probably will be necessary.

**Type III Tar Sands/Bitumen Concentrate Grade**

Previous work had shown that the concentrates obtained by the hot water processing of type III tar sands, as exemplified by Sunnyside, normally contain about 25-35% organics (bitumen + diluent). On a diluent free basis, this would translate to about 18-27% bitumen. Considering that the feed grade is 9.5% bitumen, the concentration ratio obtained is relatively low.

Several experiments were performed to investigate the reasons why a low grade concentrate is obtained during the conventional flotation of type III tar sands. The first series of experiments consisted of collecting concentrates at different flotation times. The time sequence used was 2 minutes, 3 minutes and 5 minutes. The total flotation time was 10 minutes. The results given in Table 3 show that a moderately high grade concentrate can be obtained if the flotation time is cut to 2 minutes. Typically a flotation time of 10 minutes is used. The recovery associated with the 2 minutes flotation time is 68%. Succeeding concentrates contain bitumen which is significantly lower in grade.

Further tests were made to see if the way in which flotation was conducted would affect the concentrate grade. The first test involved transferring the digested tar sand into the flotation cell and gently stirring the pulp. This technique allows for the gravity separation of the bitumen
TABLE 3

Flotation Response of Sunnyside Tar Sand

Experimental Conditions:
- Digestion Time: 15 minutes
- Digestion Temperature: 65° C
- Carbonate Addition: 1.5 lb/ton
- Diluent Addition: 30%
- Flotation Temperature: 65° C
- Penetration Time: 6.0 hours

<table>
<thead>
<tr>
<th>Product</th>
<th>Property</th>
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<tr>
<td>Concentrate I (2 minutes)</td>
<td>Bitumen Content (%) = 37</td>
</tr>
<tr>
<td></td>
<td>Bitumen Recovery (%) = 68</td>
</tr>
<tr>
<td></td>
<td>Sand Recovery (%) = 11</td>
</tr>
<tr>
<td></td>
<td>Product Weight (%) = 13</td>
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<tr>
<td>Concentrate II (next 3 minutes)</td>
<td>Bitumen Content (%) = 12</td>
</tr>
<tr>
<td></td>
<td>Bitumen Recovery (%) = 17</td>
</tr>
<tr>
<td></td>
<td>Sand Recovery (%) = 11</td>
</tr>
<tr>
<td></td>
<td>Product Weight (%) = 9</td>
</tr>
<tr>
<td>Concentrate III (next 5 minutes)</td>
<td>Bitumen Content (%) = 3</td>
</tr>
<tr>
<td></td>
<td>Bitumen Recovery (%) = 4</td>
</tr>
<tr>
<td></td>
<td>Sand Recovery (%) = 10</td>
</tr>
<tr>
<td></td>
<td>Product Weight (%) = 8</td>
</tr>
<tr>
<td>Middlings (+ 28 mesh)</td>
<td>Bitumen Content (%) = 5</td>
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<tr>
<td></td>
<td>Bitumen Recovery (%) = 8</td>
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<td></td>
<td>Sand Recovery (%) = 15</td>
</tr>
<tr>
<td></td>
<td>Product Weight (%) = 11</td>
</tr>
<tr>
<td>Tailings</td>
<td>Bitumen Content (%) = 0.7</td>
</tr>
<tr>
<td></td>
<td>Bitumen Recovery (%) = 4</td>
</tr>
<tr>
<td></td>
<td>Sand Recovery (%) = 54</td>
</tr>
<tr>
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<td>Product Weight (%) = 59</td>
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from the sand. The flotation cell is used, in effect, as a settler. The bitumen concentration produced after 5 minutes of flotation gives a diluent free assay of 69%. However, the recovery was only 16%. The second test consisted of transferring the digested tar sand into the flotation cell, turning the impeller motor on but cutting off the flow of air. The bitumen concentrate produced after 5 minutes of flotation gave a diluent free assay of 60%. The corresponding recovery increased to about 30%. The third series of experiments involved regular flotation of the tar sand with an air flow rate of about 9 liters/minute. The results after 2 minutes of flotation were: a concentrate of 40% bitumen on a diluent free basis and a recovery of 53%. The results are summarized on Table 4.

The results from the preceding series of experiments show that it is possible to obtain a high grade bitumen concentrate from type III Sunnyside tar sands provided certain adjustments are made in the manner in which the flotation is conducted. The concentrates produced in this modified flotation process are comparable in grade with those obtained from type II tar sands. The intermediate concentrates produced would have to be recycled to digestion. The middlings produced would have to undergo a size reduction step before being mixed with fresh feed. The results also point out the need for a more thorough investigation of the flotation part of the hot water process. Topics which require further attention are: flotation cell design, impeller design, rotational speed, air dispersing methods, and adjustments in flotation cell operation for different tar sand types.
TABLE 4

Flotation Response of Sunnyside Tar Sand for Different Flotation Techniques.

Experimental Conditions: Digestion Time: 15 minutes
Digestion Temperature: 60° C
Carbonate Addition: 1.5 lbs/ton
Diluent Addition: 30%
Flotation Temperature: 60° C
Penetration Time: 6.0 hours

<table>
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<tr>
<th>Flotation Technique</th>
<th>Flotation Time (minutes)</th>
<th>Grade (%)</th>
<th>Recovery (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gentle Stirring</td>
<td>5</td>
<td>69</td>
<td>16</td>
</tr>
<tr>
<td>without air</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Stirring with Impeller</td>
<td>5</td>
<td>60</td>
<td>30</td>
</tr>
<tr>
<td>without air</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Regular Flotation</td>
<td>2</td>
<td>40</td>
<td>53</td>
</tr>
<tr>
<td>air flow = 9 L/minute</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Data Processing

In view of the large amount of data collected on the hot water process, an effort to organize and store this mass of information on a computer was initiated. The computer used for this work was an Apple III using the UCSD-p operating system. The language of implementation was Pascal. The data file was stored in a 5 1/4" single-sided floppy disc and entered using the following format:

- Tar Sand Name:
- Bitumen Content:
- Viscosity at 90°C:
- Viscosity at 50°C:
- Tar Sand Type:
- Sand pH:
- Diluent Type:
- Amount of Diluent Used (gal/ton):
- Modifier:
- Amount of Modifier Used (lb/ton):
- Digestion Temperature (°C):
- Flotation Temperature (°C):
- Penetration Time (Hours)
- Concentrate Grade:
- Tailings Grade:
- Recovery:
- Coefficient of Separation:
- Comments:

So far, data for about 28 entries have been collected. The data include early work by Sepulveda on Asphalt Ridge using the original hot water process, Misra's work on Sunnyside tar sands using the original hot water process, early work on the different Kentucky tar sands using the modified hot water process and work on the Utah and Canadian tar sands using the modified hot water process.

Besides a listing of the pertinent data for a particular tar sand, the program is designed to provide the following:
1) Processing conditions for a particular tar sand at a specified temperature.

2) The separation characteristics of particular tar sand at a specified temperature.

3) A listing of the different diluents used

4) A listing of tar sand types (I, II, III, IV)

5) An empirical model which, given the bitumen content, viscosity at 90°C, viscosity at 50°C, and the sand pH, would give or predict a suggested starting point for the modified hot water process providing such information as diluent, (if required), what amount, digestion temperature, flotation temperature, and modifier, (if required).

Due to memory limitations of the Apple III microcomputer, the data mass program could not be completed. The use of an IBM P.C. with 640 K of memory would allow for the completion of proposed data base. The computed program would be interactive, user-friendly and would be able to fit in an IBM P.C. w/256 K of memory.
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UPGRADE OF TAR SAND BITUMEN BY HYDROPYROLYSIS PROCESS
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INTRODUCTION

The Department of Fuels Engineering at the University of Utah has been exploring the chemistry and engineering of hydropyrolysis of tar sand for several years. It had been previously observed (1-3) that under certain reaction conditions hydrocarbon species undergo cracking reactions with no evidence for products possessing molecular weights less than the starting material. In work involving heavy feedstocks with significant (>10%) carbon residue properties, it was further observed (4-6) that conversions of over 99% to gaseous and liquid products was possible. The general conditions for those results were 450-600°C, 1200-2000 psig H₂ and 1-30 sec residence times. The prospect of developing a noncatalytic process which could convert residual material of high metals and high carbon-residue propensity to high yields of distillable liquids appears attractive.

Following initial exploratory work, a general survey of hydropyrolysis process economics was conducted. In this survey it was determined that the most critical process variable is the gas/oil throughput ratio. Other important variables are pressure and temperature, the latter relates directly to gas production which in turn relates directly to hydrogen consumption.

On the basis of these findings, the reactor was reconfigured from a coiled-tube to a vertical tube, downflow design. Provisions were made to rapidly remove nonvolatile or unreacted material from the hot reaction zone. A summary of results are given in Table 1 and are compared with coking as a process alternative. It should be noted that hydropyrolysis produces significantly less coke for a given feedstock then does coking. The coiled tube reactor yielded relatively higher gas volumes and lesser amounts of
liquid than did the vertical tube reactor. This was probably due to longer residence times and a higher hydrogen/oil ratio. The vertical tube reactor produced larger amounts of coke due to the lower f^/oil ratios and to lower gas velocities. The coke that was formed resulted entirely from coking of feedstock which had impinged on the walls of the reactor; no coke was found in the products exiting the reactor.

Table 1
Summary Results of Hydropyrolysis Versus Coking

<table>
<thead>
<tr>
<th>FEEDSTOCKS</th>
<th>SUNNYSIDE</th>
<th>ATHABASCA</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Coking</td>
<td>Hydropyrolysis</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Coiled Tube</td>
</tr>
<tr>
<td>Gas</td>
<td>8.9</td>
<td>27.0</td>
</tr>
<tr>
<td>Liquid</td>
<td>71.0</td>
<td>73.0</td>
</tr>
<tr>
<td>Coke</td>
<td>20.1</td>
<td>NIL</td>
</tr>
<tr>
<td>API Liquids</td>
<td>26.6</td>
<td>14.8</td>
</tr>
</tbody>
</table>

Note: Values are in weight percent

During this investigation it was learned that at lower process severities, higher yields of liquid were produced and that a portion of the heavy ends of these liquids may be amenable to recycle. This is shown in Figure 1 for Athabasca tar sand. It was conceivable that the process could be operated at low conversions per pass yielding little gas and coke production. The heaviest portion could be recycled to produce an overall
Figure 1

BOILING POINT DISTRIBUTION FOR BITUMEN AND HYDROPYROLYSIS PRODUCT (ATHABASCA CASE)

- Liquid Product
- To Be Recycled
- Bitumen

NOMINAL BOILING POINT, °C

0 100 200 300 400 500 600 700 800 900 1000 1100 1200
yield of distillate in excess of 90%. The apparatus used prior to the current study had no provision for recycle of heavy liquids.

The vertical tube reactor was installed in the short-residence coal liquefaction pilot plant investigated in the 1960's and 1970's (7). This equipment, while useful to test the new reactor configuration, was not dependable and did not afford the flexibility to study the effects of varying the gas/oil ratio, the most important economic variable. Due to the poor dependability of auxiliary equipment such as heaters, controllers and compressors, run turnaround times were often a month or more. It was therefore decided that a new process development unit should be designed and built and that a systematic study of the effect of process variables on product yield and structure should be undertaken. The present study was undertaken to design and construct a semi-continuous unit which would operate within a previously determined range of variables and which would be sufficiently flexible to allow a systematic study of the effect of process variables.

OBJECTIVES AND STATEMENT OF WORK

OBJECTIVE: To advance the development of the hydropyrolysis method as a viable process for conversion of heavy feedstocks.

STATEMENT OF WORK
1. Design and fabricate a semi-continuous hydropyrolysis unit, with provisions for appropriate recycle streams. Test and modify the unit as
necessary to accomplish the remainder of this statement of work. Design should take into consideration the need to evaluate process economics and thermodynamic efficiencies.

2. Select a representative tar sand bitumen agreed upon by the Department of Energy's contracting officer's technical representative and study the effect of hydropyrolysis process variables such as temperature, pressure, gas and liquid rates, etc., on product yield and quality. Develop mathematical equations describing these process relationships.

3. Using the same feedstock selected in Task-2, determine optimum process conditions to:
   a. Maximize yield products distillable below 1000° F at one atmosphere pressure while giving consideration to maximum middle distillate yields.
   b. Maximize yield of pipelineable syncrude containing not more than 1500 ppm of either nitrogen or sulfur by a combination of hydropyrolysis and hydrotreating.

4. Compare the results obtained in Task-3 with the maximum yields of comparable products obtained by delayed coking or delayed coking/hydrotreating.

5. Based on the results obtained in tasks 2-4, conduct comparative studies on other selected tar sand bitumen or heavy oil feedstocks.
6. Conduct an economic and thermodynamic analysis of the process.

**PROCESS DESIGN**

The first step taken in the design was to develop a conceptual flow diagram. This diagram is shown in Figure 2. The unit consists of both a gas recycle and heavy liquid recycle stream. It was determined early, due to funding limits as well as the current state of knowledge, that the design should be semi-continuous. A continuous operation would have required substantially greater feedstock and product collection vessels, a duel reactor system to cycle between coke burnout and reactor modes, and a substantially better process control system than is required for a semi-continuous unit. Also, the cost of operation of a continuous unit was not warranted at this stage of development.

**PROCESS VARIABLES STUDIED**

The individual sections of the unit were designed in the context of the overall range of variables to be studied. These are given in Table-2 as follows:

<table>
<thead>
<tr>
<th>Variable</th>
<th>Range</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature, °C</td>
<td>450°-600</td>
</tr>
<tr>
<td>Pressure, psig</td>
<td>&lt; 2200</td>
</tr>
<tr>
<td>Residence Time, sec</td>
<td>1-30</td>
</tr>
<tr>
<td>Gas/Oil, scf/bbl</td>
<td>5-50 X 10³</td>
</tr>
<tr>
<td>Oil Feed Rate, L/hr</td>
<td>2</td>
</tr>
<tr>
<td>Ho Concentration in Gas (Mole Percent)</td>
<td>50-95</td>
</tr>
</tbody>
</table>
The temperature range was selected based on previous work. Below 450° C hydropyrolysis reactions are too slow to be conducted in the vapor phase. Above 600° C flash hydropyrolysis occurs which is an area felt to be costly in terms of hydrogen consumption and process severities. A design pressure of less than 2200 psig was selected because it was determined earlier that coke could be completely inhibited at pressures less than this value (but probably greater than 1200 psig). High pressures would have resulted in higher equipment costs as well as required additional safety precautions.

A residence time regime of 1-30 see is somewhat arbitrary and can easily be changed by reducing the reactor volume. Optimum conditions are expected to be much closer to 1 sec or even less. For operation at low severity, high oil recycle rates are of interest in the optimization plan. The gas/oil feed ratio has been determined to be the most economically sensitive variable and was designed for study between 5-50 X 10^4 scf/bbl. As this variable decreases, better economics prevail; however, it remains to be determined what level of operability and coke production results from low gas/oil ratios.

The 2 liter/hour fresh feed rate results from consideration of workable reactor tube dimensions. Liquid hourly space velocity of 10 is sufficient to economically utilize the reactor space. This variable can be readily changed within reasonable limits and represents a target for purposes of sizing pumps, transfer lines, heat exchangers, valves and vessels. The gas composition variable is an important variable from both a chemical and economic viewpoint. The greater the requirements for high hydrogen concentrations, the greater the gas processing costs. In practice, the system could operate with virtually no hydrogen but with high light hydrocarbon concentrations and some trial experiments in this mode are anticipated in future work.
The central design problem for the system dealt with the reactor. It was previously learned that high dilution with H2 afforded low coke production (coiled tube reactor results). It had also been learned that high proportions of vapor-phase reactions enhanced the hydropyrolytic effect and afforded high conversions at low residence times (vertical tube reactor results). It was therefore concluded that an optimum reactor design might occur from an understanding of the behavior of a droplet of feedstock when injected into a flowing stream of hydrogen. The degree of atomization and the extent of vaporization became critical parameters to the hydropyrolysis reaction results.

A mathematical model was developed to better understand the partial vaporization problem. To approach the problem it was envisioned that a spray of droplets at a given temperature, size and velocity was ejected into a reactor space at a given, but differing temperature, containing H2 at a given pressure and flowing either cocurrently or countercurrent with the flow of the droplet. The temperature, patterns vs. time, droplet size, velocity and position with time are desired.

The model considers the following three factors:

- The drag and buoyant forces on an isolated moving droplet.
- Mass transfer from the droplet and the balance of heat transfer to the droplet.
- Heat effects from the vaporization due to mass transfer and the temperature change of the droplet.

The force balance equation can be written as:
The mass balance equation can be written as

$$\frac{dL}{dt} = -J_L \left( \frac{P}{r_L} \cdot \frac{r_P}{V} \right) J_L$$

And the energy balance equation can be written as

$$\frac{dT}{dt} = -\frac{3U}{C_p r_L} (T_0 - T) \cdot \frac{3A^*}{C_p r} (R_v - \frac{[P_v]}{[y]}) \cdot \frac{M_p}{RT}$$

where

- $C_d$ = drag coefficient for drop
- $C_p$ = heat capacity for reactor gas
- $D^*$ = diffusion coefficient of droplet fluid in reactor gas
- $g$ = acceleration due to gravity
- $H$ = heat transfer coefficient to droplet
- $H_v$ = heat of vaporization of droplet fluid
- $K$ = thermal conductivity of reactor gas
- $K^*$ = mass transfer coefficient to droplet
- $M$ = molecular weight of droplet fluid
- $p$ = density of droplet fluid
- $\rho$ = density of reactor gas
\[ P = \text{vapor pressure of droplet fluid} \]
\[ (P_v)_0 = \text{partial pressure of droplet fluid far from the drop} \]
\[ r = \text{droplet radius} \]
\[ R = \text{gas constant} \]
\[ t = \text{time} \]
\[ T = \text{droplet temperature} \]
\[ T_o = \text{reactor gas temperature} \]
\[ T^* = \frac{(T + T_o)}{2} \]
\[ V = \text{droplet downward velocity} \]
\[ x = \text{droplet position below nozzle} \]

A computer solution of these three simultaneous rate equations can provide drop position in the reactor, drop velocity, drop temperature, and drop size, as functions of time. The Runga-Kutta numerical integration method was applied. Transfer coefficients \( H, K \) and \( C_d \) are calculated from literature correlations. ' ' '

Fluid properties that vary with temperature were estimated from known property, temperature, pressure correlations. Hydrogen properties required are density, viscosity, heat capacity and thermal conductivity at process conditions. Oil properties required are density, heat capacity, vapor pressure, heat of vaporization, and diffusivity. These properties were first estimated for a C\(^n\) \( n \)-paraffin. For fluids boiling higher than C\(^g\) (535\(^\circ\) C), vapor pressure and \( P_v \) were estimated by extrapolating the curves from lower boiling compounds. Density, heat capacity and diffusivity were assumed to be independent of boiling point over a narrow range.

Results from the model are as follows:

Critical Boiling Feed
For a given gas/feed ratio and a given reactor temperature, there is a critical boiling point of the feed above which all of the feed can never fully vaporize. When the partial pressure of the feed at full vaporization is the same as the vapor pressure (when the feed is at the reactor temperature), the criterion of critical boiling point is met. Using these correlations, critical boiling points were determined for several cases, and are shown in Table 3. Also, results are represented graphically in Figure 3.

Critical Drop Size at Nozzle

For a given set of reactor conditions (reactor temperature and pressure, boiling point of feed, gas to feed ratio and droplet initial velocity), the computer program DROPT can be used to calculate the distance a drop will travel before it completely evaporates. For a specified reactor length, there will be a critical drop size, drop radius \((r^*)\), such that drops initially larger than \(r^*\) will not completely evaporate while those initially smaller than \(r^*\) will completely evaporate.

Several cases were considered using the reactor specifications outlined above, that included varying the reactor temperature, feed boiling point, feed temperature, and initial drop velocity. Hydrogen to feed ratio was maintained at 2. For each case, Program DROPT was used to determine the critical drop size for a reactor length of 45.75 cm. These results are summarized in Table 4. This table also includes an estimate of the actual initial drop size at the velocities considered. Figures 4 and 5 present these results in
Table 3

Critical Feed Boiling Point, $T^*$, as a Function of Gas/Feed Ratio, $R^*$, and Reactor Temperature, $T$

<table>
<thead>
<tr>
<th>$T^*(^\circ C)$</th>
<th>$T_b (^\circ C)$</th>
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</thead>
<tbody>
<tr>
<td>450</td>
<td>529</td>
</tr>
<tr>
<td>500</td>
<td>589</td>
</tr>
<tr>
<td>525</td>
<td>619</td>
</tr>
<tr>
<td>550</td>
<td>649</td>
</tr>
<tr>
<td>575</td>
<td>678</td>
</tr>
<tr>
<td>450</td>
<td>513</td>
</tr>
<tr>
<td>500</td>
<td>574</td>
</tr>
<tr>
<td>525</td>
<td>604</td>
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<td>550</td>
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<td>500</td>
<td>507</td>
</tr>
<tr>
<td>525</td>
<td>538</td>
</tr>
<tr>
<td>550</td>
<td>569</td>
</tr>
</tbody>
</table>
Figure 3

Critical Boiling Point

Reactor Temperature - T (°C)

Critical Feed Boiling Point

"b (°C)

R = 0.5
R = 1.0
R = 1.5
R = 2.0
R = Gas/Feed
Table 4

Critical Drop Radius $r^*$, as a Function of Reactor Temperature, $T_r$, Boiling Point of Feed, $T_b$, and Initial Drop Temperature, $T_o$, and Initial Drop Velocity, $V$. Feed to Gas. Ratio is Always 2. Reactor Length is 45.75 cm

<table>
<thead>
<tr>
<th>$T_r$ ($^\circ C$)</th>
<th>$T_b$ ($^\circ C$)</th>
<th>$T_o$ ($^\circ C$)</th>
<th>$V$ (cm/sec)</th>
<th>$r^*$ (cm)</th>
<th>Estimated Actual Drop Radius (cm)</th>
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<tr>
<td>525</td>
<td>525</td>
<td>450</td>
<td>10000</td>
<td>.00838</td>
<td>.0115</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>5000</td>
<td>.0104</td>
<td>.0128</td>
</tr>
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<td></td>
<td></td>
<td></td>
<td>1000</td>
<td>.0149</td>
<td>.0141</td>
</tr>
<tr>
<td>525</td>
<td>580</td>
<td>450</td>
<td>10000</td>
<td>.00575</td>
<td>.0115</td>
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<td></td>
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<td>5000</td>
<td>.00769</td>
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<td>575</td>
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<td></td>
<td>5000</td>
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<td></td>
<td>1000</td>
<td>.0158</td>
<td>.0141</td>
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<tr>
<td>575</td>
<td>650</td>
<td>500</td>
<td>10000</td>
<td>.00398</td>
<td>.0115</td>
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<td></td>
<td>5000</td>
<td>.00533</td>
<td>.0128</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>1000</td>
<td>.00814</td>
<td>.0141</td>
</tr>
</tbody>
</table>
Reactor Temperature = 525°C
Initial Drop Temperature = 450°C

Figure 4- Critical Drop Radius for Reactor Temperature 525°C.
Reactor Temperature = 575°C
Initial Drop Temperature = 500°C

Critical Drop Radius (cm)

Estimated Drop Radius

Velocity (cm/sec)

Critical Drop Radius for Reactor Temperature 575°C.

Figure 5

B17
The development of this model has allowed for reactor design using fundamental engineering principles. This program allows for predicting the fraction occurring in the vapor-phase as a function of feedstock boiling point, initial drop size and velocity, reactor temperature and pressure and vapor composition.

There are theoretically four stream flow configurations, cocurrent downflow, co-current upflow, countercurrent flow with oil flowing downward and countercurrent flow with hydrogen flowing downward. Of these only the first and last are of practical consideration. Cocurrent upflow and countercurrent flow with oil flowing downward would require uneconomically high gas velocities. Also, these configurations would be plagued with coke formation in the bottom of the reactor zone.

Of the two remaining options, it was not immediately clear which configuration provided the overall best performance, and the DROPT program was used to help with this decision. A number of cases were run using droplets of differing sizes, differing reactor diameters (affecting gas velocities) and differing temperatures. Typical results are shown in Table 5. The results show that the countercurrent case dramatically improves vaporization efficiency. This configuration also has the advantage of providing a favorable residence time distribution in the reactor. On this basis the initial design was that of countercurrent flow with gas downflow and oil upflow in a fountain type arrangement.
Table 5

Predicted Vaporization of a Droplet of Oil

**Conditions:**
- $T_{\text{Oil}} = 400^\circ C$
- $H_2/\text{Feed} = 0.5 \text{ g/g}$
- $R = 0.08 \text{ cm initial drop radius}$
- $V_0 = 2600 \text{ cm/sec initial drop velocity}$

<table>
<thead>
<tr>
<th>Reactor Configuration</th>
<th>$(R_f/R)^3$ at 525$^\circ$ C</th>
<th>$(R_f/R)^3$ at 575$^\circ$ C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cocurrent Downflow</td>
<td>0.74</td>
<td>0.68</td>
</tr>
<tr>
<td>Countercurrent Gas Downflow</td>
<td>0.51</td>
<td>0.43</td>
</tr>
</tbody>
</table>

The other major design feature of the reactor deals with the configuration of the nozzle. Previous experiments run on a 1" X 14" cocurrent downflow reactor revealed coke buildup approximately 2"-3" inches below the nozzle. The nozzle was designed for a 40° spray angle, and it was deduced that coke buildup was probably due to droplets impinging on the wall. To afford good space utilization of the reactor, a relatively narrow bore would be the most efficient. This required a jet of droplets which could be directed along the axis of the reactor. Calculations of the optimum bore diameter for these droplets revealed a bore diameter of 0.004". This size proved to be too small for convenient manufacture. However, hypodermic stock possessing internal diameters of 0.007" and 0.010" was available, and this material was used as a sleeve in a larger diameter hole drilled in the nozzle.
housing. Several different configurations have been produced and tested. A donut shaped design with a radius of 1/2 that of the reactor radius and possessing five .007" orifices equally spaced about the circumference was constructed. A three-orifice pattern of similar configuration was also constructed. A single orifice bullet shaped design of 0.010" was also constructed. The nozzle and reactor configuration needs additional study.

**PREHEATER DESIGN**

In the hydropyrolysis process envisioned, oil is heated to cracking temperature but below coking temperature (380-420°C) and may or may not be saturated with hydrogen. Hydrogen rich gas is preheated to sufficient temperature, typically 550-675°C such that when oil is sprayed into the gas stream, the desired reaction temperature is achieved. With this technique, liquid phase feed does not rise above coking temperatures unless it is in direct contact with hydrogen.

The oil preheater design was rather straight-forward in that the heat required to raise 2 liter/hour oil from ambient temperatures to 400°C was easily calculable. Any vaporization that may occur in the preheater could be handled by the overdesign factor or by heat taping the lines leading into the preheater. The oil preheater consisted of a cylindrical block of bronze-aluminum from which a V4 " circular profile groove had been cut in a helical fashion the length of the block. AV4 " O.D. heavy wall 316 S.S. tube was wound into the groove and a split stainless steel shell was clamped as a jacket around the assembly. Two calrod type heater elements were inserted into holes bored axially into the interior of the block. The entire assembly was insulated for heat loss.
The gas preheater design was appreciably more involved. Preliminary calculations of the energy required to heat hydrogen revealed that the heater would be large. Fired heaters were rejected for problems with safety and poor control in intermittent operation. Electrical block heaters such as had been used for the oil preheater were rejected due to material problems at high temperatures coupled with low heat transfer coefficients. The basic type of heater designed was a fluidized bed sand bath.

The design basis for the preheater is the heating of 15 SCF/min. \( H_0 \) at 2000° psig from 25° to 650° C. The heater tube must be designed for 700° C to allow a 50° C temperature differential between the sand and the hydrogen gas. The heat input is calculated as follows:

\[
20,000 \text{ Btu/hr} = \frac{15 \text{ SCF/min} \times 60 \text{ min/hr} \times 2 \text{ lb/lb-mol} \times 1125° \text{ F} \times 3.5 \text{ Btu/lb-°F}}{359 \text{ SCF/mol}}
\]

\[
20,000 \text{ Btu/hr} = 5.8 \text{ Kw electric power.}
\]

Allowing for 70% efficiency and 20% spare capacity, the electric power input required is about 10 Kw.

To determine the overall heat transfer coefficient, \( U \), experiments were run using a helical tube immersed in a laboratory scale fluidized-bed sand bath heater. In addition to \( U \), values for the convective heat transfer from the sand bath to the tube wall (\( H_Q \)) and from the inner wall to the hydrogen (\( H_I \)) were also determined as a function of Reynolds number (\( Re \)) of the flowing hydrogen gas and the temperature of the sand. At a sand bath temperature of 675°K, \( H_Q \) ranged from 0.9 - 3.2, \( H_I \) from 9.3-17 and \( U \) from 0.8-2.6 Btu/hr ft\(^2\) °F for \( Re \) ranging from 807-2572. At temperatures of 800° K and \( Re=8544, H_I=89, H_Q=13.2 \) and \( U=11 \).
After some discussion and calculation, it was decided to arrange the internal tubes in a checkerboard pattern alternating the axial direction between layers. The design is shown in Figure 6. Results from tests of this pattern in the laboratory scale heater are given in Table 6 and reveal $U = 5 - 14$ over a $Re=5,300 - 16,000$.

The values for overall heat transfer coefficient agree with the low end of the range of 5-50 Btu/hr ft.$^\circ$F given in Peters and Timmerhas (10) for fluidized-bed heaters.

Using published piping code data, a tubing specification of V4" by 0.065" wall 316 stainless steel was chosen for construction material. At $U=8.7$, a reasonable value from Table 6, a length of 50 ft of tubing was required. To allow for higher capacity, it was decided to build 80 ft of tubing. An input of 10,200 W heaters arranged in two tiers (see Figure 6) was also designed.

**FEED SYSTEM DESIGN**

The feed system consists of a feed vessel, mass flowmeter, filters, pumps, a surge chamber and a mixer. Due to the highly viscous nature of some feedstocks, provisions were made for heating the entire system and lines with resistance heating tapes. The feed reservoir is capable of being pressurized to 300 psig thus allowing for a more positive flow ahead of the pump. Filter sizes of 140 mm were selected. A gas (H2) filled surge chamber was installed to dampen the pulses experienced from reciprocating pumps. Both the feed pump and recycle pump were of a Milton Roy reciprocating piston design. A mixing chamber consisting of an upward flowing tube inserted within a downward flowing tube served to create an axially homogeneous stream of recycle and fresh feed to be fed to the reactor. Pressure relief valves were installed.
Figure 6
Fluidized Sand Bath Hydrogen Preheater

Top View

Cold Gas in

32 Feet 1/4" SS Tube

Hot Gas out

Porous Stainless Steel Plate

air in

3400 Watt

6800 Watt

No Scale
TABLE 6

Heat Transfer Data, Checkerboard Tube Arrangement.

<table>
<thead>
<tr>
<th>Run No.</th>
<th>Tin °C</th>
<th>Tout °C</th>
<th>T Bath °C</th>
<th>Flow Rate ft/sec</th>
<th>( H_o ) Btu/hr-ft²°F</th>
<th>( U ) ft²-°F</th>
<th>Q Btu/hr</th>
<th>Re</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>25</td>
<td>406</td>
<td>445</td>
<td>.306</td>
<td>11.3</td>
<td>9.7</td>
<td>483</td>
<td>11,200</td>
</tr>
<tr>
<td>2</td>
<td>25</td>
<td>406</td>
<td>448</td>
<td>.440</td>
<td>16.2</td>
<td>13.7</td>
<td>688</td>
<td>15,900</td>
</tr>
<tr>
<td>3</td>
<td>24.5</td>
<td>407</td>
<td>450</td>
<td>.268</td>
<td>9.5</td>
<td>8.4</td>
<td>425</td>
<td>9,800</td>
</tr>
<tr>
<td>4</td>
<td>24.7</td>
<td>420</td>
<td>450</td>
<td>.149</td>
<td>5.6</td>
<td>4.9</td>
<td>243</td>
<td>5,400</td>
</tr>
<tr>
<td>5</td>
<td>24.7</td>
<td>371</td>
<td>400</td>
<td>.208</td>
<td>7.8</td>
<td>6.8</td>
<td>300</td>
<td>7,800</td>
</tr>
<tr>
<td>6</td>
<td>24.7</td>
<td>374</td>
<td>402</td>
<td>.417</td>
<td>16.6</td>
<td>13.9</td>
<td>606</td>
<td>15,700</td>
</tr>
<tr>
<td>7</td>
<td>24.7</td>
<td>355</td>
<td>400</td>
<td>.417</td>
<td>14.8</td>
<td>12.7</td>
<td>575</td>
<td>16,000</td>
</tr>
</tbody>
</table>

Tube I.D. =0.0625 in (1/4)  O. D. =0.125 in (1/8)

length = 44"
ahead of a check valve so that if such that should coking of the tubes occurred, line rupture would be prevented and back flush of hydrogen to the atmosphere would be avoided. Provisions were made to bypass check valves when back flushing of oil from the feed system was required.

**SEPERATION AND PRODUCT VESSEL DESIGN**

The primary separation vessel was fabricated from a 3/4" I.D. S.S. tube 12" long. The inlet was positioned 4" from the top of the nozzle and entered tangentially to the walls to assist separation of phases. A water-cooled condensor was used to cool the overhead stream which entered a series of high-pressure knockdown pots. The gas stream was filtered prior to recycle. The heavy oil from the primary separator was fed to recycle or to a collection vessel, depending on the mode of operation.

**GAS HANDLING SYSTEM DESIGN**

The flow diagram for the gas handling system is shown in Figure 7. The principal features of this system are provisions to automatically draw gas from storage or dump gas to storage when pressure demands it. Provisions are also available to meter fresh make-up hydrogen to the system, automatically generating a purge or slip stream. Volumetric flow rate was readily achieved by regulating the compressor inlet pressure. The gas system functioned according to design.

**SAFETY DESIGN**

Perhaps the most important design feature of the system is that of safety. Considerations for safety were as follows:
**HP PROCESS****

GAS HANDLING SYSTEM

Figure 7
Proper specification of materials
Overpressure relief
Isolation and blowout protection
Leak detection
Power failure
Electrical hazard
Emergency preparedness.

Specification of materials was conducted using standard charts for tubing and vessels. A 100% contingency factor was designed for high pressure components. Heating of large vessels was avoided in the design, the largest vessel being the reactor tube. Overpressure relief valves were placed ahead of heaters where coke formation might restrict flow of process streams. These valves were vented to the outdoors. Check valves were installed to limit the back flushing of streams through the pressure relief valves.

Prior to each run, the system was pressure tested and leaks were traced using an electronic hydrogen gas detector. A permanent combustible gas detector was installed in the reactor vicinity to detect buildup of hydrogen in the atmosphere. The main reactor system and compressors were housed behind a steel barricade designed to provide a measure of safety in the event of catastrophic pressure release or fire. The system was so designed that power failure does not result in an increased hazard. A shunt trip was installed on the electrical panel which when depressed trips power breakers at a location remote from the process site. This "panic button" can be used also when unforeseen problems develop in the system which could become a safety problem.

All electrical wiring was run in grounded electrical conduits. The panels and reactor frame were of metallic construction and were electrically...
grounded. All instruments, variacs and switches were grounded to the panel. The compressor system was protected by sensing relay switches from overpressure, underpressure, or loss of oil pressure or cooling water.

FABRICATING THE UNIT

The fabrication stage of the project occurred over a seven month period from January to July 1983 and required approximately 150 man-weeks of effort. It had originally been thought that the previously used short-residence-time coal liquefaction unit would be upgraded for use in the present study. However, shortly after the design stage commenced, it became apparent that considerable effort could be expended to retrofit this old equipment and the results would still suffer from lack of flexibility. Also, the size of the equipment, particularly the volumes contained in the recycle loops indicated that long run times may be required to achieve steady-state operation. In consultation with the DOE Technical Project Officer, it was decided to start from scratch, salvaging useful components from the old unit, but redesigning and fabricating a new unit to meet the objectives of the work.

Design, procurement and fabrication of a grass-roots unit proved to require greater time and effort than anticipated. This was due to several factors, principally an increased awareness and desire to build in safety, particularly the electrical system where over 1000 ft of wiring was installed. Delays in procurement of some critical items including the installation of electrical power panels to the process unit location, and the inexperience of
Figure 8: Electrical Panel

Figure 9: Plumbing Panel
Figure 10: General View of Reactor Area

Figure 11: Hydrogen Compressor
Figure 12: Gas Preheater

Figure 13: Oil Preheater and Separation Vessel
personnel were also realized. In this latter regard it should be noted that 3 graduate students and 3 undergraduate students received important training in the design and construction of a major process apparatus.

Photographs of the unit are provided in Figures 8-13.

OPERATION OF UNIT AND RESULTS

An operating procedure was drafted for the unit and initial runs on kerosene were made during initial testing. After several runs with kerosene, the feedstock was changed to TS-IIC oil. This oil had been produced in an in-situ reverse combustion experiment conducted by the U.S. Department of Energy, Laramie Energy Technology Center at a northwest Asphalt Ridge, Utah site. This oil possessed a slightly lower viscosity than other available tar sand bitumens and it was felt that the ease of operation was important to the early studies on the new unit.

Properties and elemental composition of the feedstock oil are given in Table 7. Elemental composition is characteristic of Uinta Basin bitumen in that hydrogen content is relatively high and nitrogen content is higher than sulfur content. The relatively high °API gravity of 20 and low molecular weight of 410 differs somewhat from a 13 °API, 668 Molecular weight material solvent extracted from a core obtained from the same general location.

The properties of the TS-IIC oil are consistent with slightly visbroken bitumen. A total of about 30 runs have been made on the unit, for the first 16 produced few usable results. These 16 runs served as start-up runs to refine operating procedures. Modifications to the equipment, especially with regard to location of mass flowmeters and to the configuration of the heavy oil
Table 7- Properties and Elemental Composition of TS-IIC Oil

<table>
<thead>
<tr>
<th>Property</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon, Wt. %</td>
<td>86.6</td>
</tr>
<tr>
<td>Hydrogen, Wt. %</td>
<td>11.6</td>
</tr>
<tr>
<td>Sulfur, Wt. %</td>
<td>0.5</td>
</tr>
<tr>
<td>Nitrogen, Wt. %</td>
<td>0.7</td>
</tr>
<tr>
<td>Oxygen, Wt. %</td>
<td>&lt; 0.5</td>
</tr>
<tr>
<td>H/C Atomic Ratio</td>
<td>1.59</td>
</tr>
<tr>
<td>Vanadium, ppm</td>
<td>&lt; 0.5*</td>
</tr>
<tr>
<td>Nickel, ppm</td>
<td>45.0*</td>
</tr>
<tr>
<td>°API Gravity</td>
<td>16.0</td>
</tr>
<tr>
<td>Specific Gravity</td>
<td>0.959</td>
</tr>
<tr>
<td>Avg. M.W., VPO Benzene</td>
<td>410.0</td>
</tr>
<tr>
<td>Conradson Carbon Residue, %</td>
<td>6.8</td>
</tr>
<tr>
<td>Distillable Below 530°C</td>
<td>69.0</td>
</tr>
<tr>
<td>Viscosity, Kp 77°F 200 sec⁻¹</td>
<td>3.9</td>
</tr>
<tr>
<td>Heating Value, Btu/lb (Calculated)</td>
<td>18,910</td>
</tr>
</tbody>
</table>

Data taken from reference 11 (STO-5).
collection system were identified during these runs. Coking of the nozzle was occasionally experienced until procedures for keeping the nozzle cool during start-up and shut-down operations were developed. The outlet of the reactor was modified several times to better accommodate the removal of nonvolatile and unreacted material. Several trial runs were made before thermocouple placement at the reactor outlet and in separation vessels was found satisfactory. Material balances were unsatisfactory in the early runs due to less-than-optimum placement of flowmeters and to loss of light liquids in the gas stream. These problems were corrected, and by run #17 useable data were being obtained.

Table 8 specifies the process conditions for runs #17-29. All runs were done using TS-IIC as a feedstock. Run conditions were not as systematically varied as originally desired. While pressure control, feedrate control and gas rate control are easily obtained, other variables such as temperature and gas composition are not as readily controlled. Temperature is particularly difficult to control in that the final outlet temperature is a function of the H2 temperature, oil inlet temperature, guard furnace power, AH and AH. In a continuous process, adjustments could be made to reach a stable temperature, but achieving a desired, stable temperature in a short-duration semi-continuous unit requires considerable operating experience. As can be seen from Table 8, pressure has been held constant and the oil feed rate also has been held virtually constant. Higher temperatures are most easily achieved by increasing the gas flow rate, but this results in a lower residence time. For purposes of discussion of results, the data in the following tables are arranged in order of increasing temperature.
### Table 8 - Process Conditions

<table>
<thead>
<tr>
<th>Run #</th>
<th>T-Out °C</th>
<th>P, psig</th>
<th>Gas Rate, q/min</th>
<th>Oil Rate, g/min</th>
<th>Gas/Oil Ratio</th>
<th>Gas Residence Time/min</th>
</tr>
</thead>
<tbody>
<tr>
<td>17</td>
<td>497</td>
<td>1860</td>
<td>11.0</td>
<td>28.9</td>
<td>.36</td>
<td>7.2</td>
</tr>
<tr>
<td>18</td>
<td>527</td>
<td>1800</td>
<td>17.3</td>
<td>30.4</td>
<td>.57</td>
<td>4.2</td>
</tr>
<tr>
<td>19</td>
<td>513</td>
<td>1850</td>
<td>11.0</td>
<td>30.0</td>
<td>.38</td>
<td>6.6</td>
</tr>
<tr>
<td>20</td>
<td>552</td>
<td>1800</td>
<td>19.4</td>
<td>30.4</td>
<td>.64</td>
<td>3.7</td>
</tr>
<tr>
<td>21</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>failed run - improper equipment configuration</td>
</tr>
<tr>
<td>22</td>
<td>496</td>
<td>1900</td>
<td>10.0</td>
<td>30.0</td>
<td>.33</td>
<td>7.4</td>
</tr>
<tr>
<td>23</td>
<td>513</td>
<td>1800</td>
<td>12.0</td>
<td>30.0</td>
<td>.40</td>
<td>5.7</td>
</tr>
<tr>
<td>24</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>failed run - coked nozzle</td>
</tr>
<tr>
<td>25</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>failed run - improper nozzle configuration</td>
</tr>
<tr>
<td>26</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>failed run - improper nozzle configuration</td>
</tr>
<tr>
<td>27</td>
<td>495</td>
<td>1800</td>
<td>12.2</td>
<td>30.5</td>
<td>.40</td>
<td>6.0</td>
</tr>
<tr>
<td>28</td>
<td>506</td>
<td>1800</td>
<td>12.0</td>
<td>35.0</td>
<td>.34</td>
<td>5.7</td>
</tr>
<tr>
<td>29(^a,b)</td>
<td>538</td>
<td>1800</td>
<td>16.1</td>
<td>28.3</td>
<td>.57</td>
<td>4.5</td>
</tr>
</tbody>
</table>

(a) Single orifice nozzle, all previous runs 3 orifice design.

(b) Run #29 was conducted in a cocurrent downflow mode. All others were in the countercurrent "fountain" mode.
Table 9 - Product Yields

<table>
<thead>
<tr>
<th>Runs #</th>
<th>T-Out, °C</th>
<th>Gas (by Difference) Wt. %</th>
<th>Liquid Wt. %</th>
<th>Heavy Oil Wt. %</th>
<th>Coke Wt. %</th>
</tr>
</thead>
<tbody>
<tr>
<td>27</td>
<td>495</td>
<td>8.1</td>
<td>79.7</td>
<td>10.0</td>
<td>2.2</td>
</tr>
<tr>
<td>22</td>
<td>496</td>
<td>1.5</td>
<td>81.6</td>
<td>15.7</td>
<td>1.2</td>
</tr>
<tr>
<td>17</td>
<td>497</td>
<td>10.1</td>
<td>74.3</td>
<td>13.7</td>
<td>1.9</td>
</tr>
<tr>
<td>28</td>
<td>506</td>
<td>14.6</td>
<td>75.4</td>
<td>6.8</td>
<td>3.2</td>
</tr>
<tr>
<td>23</td>
<td>513</td>
<td>3.9</td>
<td>75.8</td>
<td>19.4</td>
<td>0.9</td>
</tr>
<tr>
<td>19</td>
<td>513</td>
<td>9.8</td>
<td>81.1</td>
<td>7.2</td>
<td>1.9</td>
</tr>
<tr>
<td>18</td>
<td>527</td>
<td>11.2</td>
<td>81.2</td>
<td>5.8</td>
<td>1.9</td>
</tr>
<tr>
<td>29</td>
<td>538</td>
<td>8.9</td>
<td>82.5</td>
<td>3.0</td>
<td>5.6</td>
</tr>
<tr>
<td>20</td>
<td>552</td>
<td>15.9</td>
<td>81.8</td>
<td>2.3</td>
<td>3.5</td>
</tr>
</tbody>
</table>

Product yields are shown in Table 9. In general, as temperature is increased, depth of conversion is increased. This is associated with higher coke and gas production and lower heavy oil production. The major liquid product does not appear to be highly sensitive to temperature. At a constant residence time this may not be true. The liquid product yield of 80+% is substantially below the 90+% achieved in the previous reactor at a 1-2 sec residence time. Further work will be required to optimize the nozzle and reactor configuration before a study of the effect of variables can be fully accomplished.

Liquid product properties are shown in Table 10. All properties remain fairly constant as might be expected based on the relatively constant amount of liquid production. The lower viscosity of run #29 may be due to a slightly more aromatic material resulting from cocurrent downflow system.
Table 10 - Liquid Product Properties

<table>
<thead>
<tr>
<th>Run #</th>
<th>°API</th>
<th>R.I.</th>
<th>M20° (cps)</th>
</tr>
</thead>
<tbody>
<tr>
<td>27</td>
<td>25.2</td>
<td>1.512</td>
<td>48.1</td>
</tr>
<tr>
<td>22</td>
<td>24.1</td>
<td>1.509</td>
<td>61.7</td>
</tr>
<tr>
<td>17</td>
<td>25.9</td>
<td>1.510</td>
<td>89.6</td>
</tr>
<tr>
<td>28</td>
<td>24.3</td>
<td>1.510</td>
<td>41.1</td>
</tr>
<tr>
<td>23</td>
<td>23.9</td>
<td>1.511</td>
<td>88.4</td>
</tr>
<tr>
<td>19</td>
<td>23.7</td>
<td>1.513</td>
<td>43.9</td>
</tr>
<tr>
<td>18</td>
<td>28.8</td>
<td>1.513</td>
<td>81.9</td>
</tr>
<tr>
<td>29</td>
<td>25.1</td>
<td>1.509</td>
<td>15.9</td>
</tr>
<tr>
<td>20</td>
<td>24.0</td>
<td>1.509</td>
<td>52.0</td>
</tr>
</tbody>
</table>

Simulated distillation results are given in Table 11. Results given are somewhat constant, albeit there appears to be a definite trend toward a broader boiling point distribution as temperature is increased.

Table 11 - Simulated Distillation Results

<table>
<thead>
<tr>
<th>Run #</th>
<th>Gasoline, 90-200°C</th>
<th>Kerosene, 200-275°C</th>
<th>Gasoil, 275-325°C</th>
<th>Hvy Gas Oil, 325 °C</th>
<th>Vacuum Gas Oil, 400-538°C</th>
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<tr>
<td>27</td>
<td>11.3</td>
<td>23.9</td>
<td>18.6</td>
<td>24.5</td>
<td>21.7</td>
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<tr>
<td>22</td>
<td>9.6</td>
<td>22.4</td>
<td>18.5</td>
<td>23.4</td>
<td>26.1</td>
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<td>10.3</td>
<td>21.0</td>
<td>17.4</td>
<td>22.4</td>
<td>28.9</td>
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<td>28</td>
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<td>20.7</td>
<td>17.3</td>
<td>22.6</td>
<td>28.6</td>
</tr>
<tr>
<td>23</td>
<td>8.4</td>
<td>19.1</td>
<td>16.8</td>
<td>23.2</td>
<td>32.4</td>
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<tr>
<td>19</td>
<td>10.9</td>
<td>20.3</td>
<td>16.8</td>
<td>20.7</td>
<td>31.2</td>
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<tr>
<td>18</td>
<td>17.3</td>
<td>18.3</td>
<td>15.4</td>
<td>21.1</td>
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<td>29</td>
<td>18.2</td>
<td>21.4</td>
<td>16.2</td>
<td>20.8</td>
<td>23.4</td>
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<tr>
<td>20</td>
<td>11.4</td>
<td>20.0</td>
<td>16.7</td>
<td>22.1</td>
<td>29.8</td>
</tr>
</tbody>
</table>
An important task of this project was to perform an economic and thermodynamic assessment of the hydropyrolysis process. This was done in a preliminary way with the best available data. Because process conditions have not been optimized, it was important to view the economics as less than the economic optimum possible.

An economic assessment of hydropyrolysis has been conducted for a single case. The conditions and stream compositions are based on experimental data or estimation. The conditions used do not represent an optimum, but rather it is an attempt to determine major factors affecting economic sensitivity.

A block process flow sheet is shown in Figure 14. Note that the process as drawn is once-through with a gas recycle stream. A mass balance is shown in Figure 15 for the overall input and output streams. Capital costs for a 10,000 bbl/day unit are given in Table 12. Note that the major costs are in the compressors and heat exchangers indicating that the gas recycle rate is economically the most important process variable. The total capital costs of about $6,000/daily bbl capacity compares to about $4,000/daily bbl capacity for a coker.**

Operating costs are shown in Table 13. Cost analysis shows a modest cost for both H2 consumption and utility requirements. This is consistent with the latest experimental results which reveal that H2 consumption can be quite low. The distributing and marketing costs are chart values and appear to be high for application to a tar sand processing facility.
Figure 15- Mass Balance for Hydropyrolysis Process (once-through)

10,000 bbl/day
Table 12
Capital Costs $1000's

<table>
<thead>
<tr>
<th>Major Equipment</th>
<th>429.4</th>
<th>469.7</th>
<th>3,220.4</th>
<th>1,376.7</th>
<th>3,584.0</th>
<th>201.0</th>
<th>171.8</th>
<th>506.0</th>
<th>1,539.0</th>
<th>285.0</th>
<th>786.5</th>
<th>1,063.4</th>
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<tr>
<td>Fired Heater (H-1)</td>
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<td></td>
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<tr>
<td>Fired Heater (H-2)</td>
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<tr>
<td>Heat Exchanger (H-3)</td>
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<td>Heat Exchanger (H-4)</td>
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<tr>
<td>Cold Water Exchanger (Col-I)</td>
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<tr>
<td>Flash Separator (F-1)</td>
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<tr>
<td>Flash Separator (F-2)</td>
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<tr>
<td>Pump (P-1)</td>
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<td>Compressor (C-1)</td>
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<tr>
<td>Reactor (HPRE)</td>
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<td>Tank</td>
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<td></td>
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<tr>
<td>H2 Separator</td>
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<tr>
<td><strong>Total Equipment (E)</strong></td>
<td>10,633.7</td>
<td></td>
<td></td>
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</table>

| Equipment Installation (47% E) | 4,997.9 |
| Instrumentation & Controls (18% E) | 1,914.1 |
| Piping (installed) (66% E) | 7,018.3 |
| Electrical (installed) (11% E) | 1,169.8 |
| Buildings (18% E) | 1,914.1 |
| Yard Improvements (10% E) | 1,063.0 |
| Service Facilities (70% E) | 7,443.4 |
| Land (6% E) | 638.1 |
| Engineering Supervision (33% E) | 3,509.0 |
| Construction Expenses (41% E) | 4,359.0 |
| **Total Direct + Indirect (D + I) Costs** | 44,662.1 |

| Contractors Fees (5% D + I) | 2,233.2 |
| Contingency (10% D + I) | 4,466.3 |

| Fixed Capital Investment | 51,362.6 |
| Working Capital (15%) | 9,145.0 |
| **Total Capital Investment** | 60,506.6 |
Table 13

Costs 10,000/day **Hydropyrolysis Process**

<table>
<thead>
<tr>
<th><strong>Direct Charges</strong></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Raw Materials</strong></td>
<td>142.9</td>
</tr>
<tr>
<td>(Oil @ 15% bbl)</td>
<td>149.3</td>
</tr>
<tr>
<td>H₂ @ $3.00/MMBtu</td>
<td>2.7</td>
</tr>
<tr>
<td><strong>Credits</strong></td>
<td>(9.1)</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Labor</strong></td>
<td>2.6</td>
</tr>
<tr>
<td>Operating</td>
<td>2.1</td>
</tr>
<tr>
<td>Supervisory &amp; Clerical</td>
<td>0.5</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Utilities</strong></td>
<td>6.8</td>
</tr>
<tr>
<td>Cooling Water</td>
<td>1.0</td>
</tr>
<tr>
<td>Fuel</td>
<td>2.7</td>
</tr>
<tr>
<td>Electricity</td>
<td>1.8</td>
</tr>
<tr>
<td>H₂ Separator</td>
<td>1.3</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Maintenance &amp; Repair</strong></td>
<td>15.6</td>
</tr>
<tr>
<td>Supplies</td>
<td>2.3</td>
</tr>
<tr>
<td>Laboratory Charges</td>
<td>0.4</td>
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<tr>
<td>Overhead Costs</td>
<td>10.9</td>
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<tr>
<td></td>
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<tr>
<td><strong>Total Direct Charges</strong></td>
<td>181.5</td>
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<table>
<thead>
<tr>
<th><strong>Fixed Charges</strong></th>
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<tr>
<td>Depreciation</td>
<td>15.6</td>
</tr>
<tr>
<td>Local Taxes</td>
<td>3.1</td>
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<tr>
<td>Insurance</td>
<td>1.6</td>
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</table>

<table>
<thead>
<tr>
<th><strong>General Expenses</strong></th>
<th>65.3</th>
</tr>
</thead>
<tbody>
<tr>
<td>Administrative</td>
<td>2.7</td>
</tr>
<tr>
<td>Distributing &amp; Marketing</td>
<td>40.1</td>
</tr>
<tr>
<td>R &amp; D</td>
<td>13.3</td>
</tr>
<tr>
<td>Interest</td>
<td>9.2</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>267.1</td>
</tr>
</tbody>
</table>
The venture economics for the initial year, valuing the product at $35/bbl are:

\[
\text{ROI} = \frac{\text{Income} - \text{Costs} - \text{Taxes}}{\text{Investment}}
\]

\[
= \frac{(347.7 - 267.1 - 40.3)}{60,506.6} \text{ (330 stream days/year)}
\]

\[
= 21.7\%
\]

The value of product at which the plant makes no profit is $26.7/bbl.

The main conclusion that should be drawn from this analysis is that there does not appear to be any economic factors which clearly prohibit hydroD.yrolysis from being a competitive process for heavy oil and bitumen upgrading.

SECOND-LAW THERMODYNAMIC ANALYSIS OF HYDROPYROLYSIS

The thermodynamic efficiency of hydropyrolysis was determined using the process simulation data previously determined. The availability function B was calculated for each stream where \( B = H - T Q S \) and \( T Q = 560^\circ \text{R} \). For these calculations:

\[
H = H_{\text{chemshare}} + ^\circ \text{(32 F, 1 atm)} \text{ and }
\]

\[
s^f = s_{\text{chemshare}} + ^\circ \text{(32 F, 1 atm)}.\]

The lost work in the system is calculated from the equation:

\[
W_L = W_{\text{in}} - W_{\text{out}} + W_s
\]

where:

\( W_s \) is the shaft work for pumps and compressors.

Figure 16 shows the availability function balance for the process. Calculations indicating the following:
Figure-16

THERMODYNAMIC ANALYSIS OF HYDROPYROLYSIS

Pump

\[ W_P = 696,132 \text{ Btu/hr} \]

Compressor

\[ W_C = 930,366 \text{ Btu/hr} \]

**HYDROPYROLYSIS SYSTEM**

6 Liquids

-\[ B = -14,258,959 \text{ Btu/hr} \]

9 Gas

-\[ B = -4,071,843 \text{ Btu/hr} \]

10 H₂

-\[ B = 10,042,893 \text{ Btu/hr} \]

B = -776,000 Btu/hr

B = -171,527 Btu/hr

**Feed Oil**

-33,858,364 Btu/hr

10,662,336 Btu/hr

\[ T_0 = 560^\circ \text{R} \]

**Fluid Gas (Vapor)**
The value of 53% is neither high nor low due to the nature of the calculations. For highly reversible processes the efficiency approaches 100%. This type of analyses, when applied to the unit operations within the process can often show where improvements can be made. About 7% of the heating value of the feed is required for process heat or electric drives indicating that hydropyrolysis has a high thermal efficiency.

SUMMARY

A 2 liter/hour hydropyrolysis process development unit has been designed and constructed. The unit provides for high flexibility for the study of the effect of variables. The system is capable of total or partial recycle of both non-condensible gas and residual liquids. The system is operational and preliminary data has been generated. The process does not appear to be economically flawed and it possesses reasonable thermodynamic efficiencies. The present system is available for process development studies including reaction chemistry, nozzle and reactor configuration tests, studies of kinetics and effect of process variables and application to unconventional feedstocks.
REFERENCES


For purposes of expediting the design and fabrication of the unit, responsibility for the various portions of the unit were assigned to specific individuals. The major design effort occurred over the time period of 1 April - 31 December 1983. For the record the assignments were as follows:

Don Cogswell - Overall design, panel layout, gas handling system, electrical requirements, controller of heater and piping specifications.

Hsi-Chung Tsai - Feed system, filtering, mixing and blowout protection.

Raymond Umoh - Reactor system, nozzle design, tube sizing.

Jerry Wiser - Separation vessels, primary separator temperature and pressure control.

Greg Neff - Preheater design, hydrogen preheater, oil preheater.

Chris Russell - Phase thermodynamics, DROPT model and reactor configuration.

and Tom Cable.
SECTION C

THE FLUIDIZED BED PYROLYSIS OF BITUMINOUS SANDS
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### THE aUIDIZED BED PYROLYSIS OF THE BITUMINOUS SANDS

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<td>Analysis of the Liquid Product from the Fluidized Bed Pyrolysis of the P.R. Spring South Tar Sand</td>
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<td>The Effect of Solids Residence Time on Product Distribution from</td>
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<td>Hopper, One Cover Top Section</td>
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<td>Mist Collector</td>
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The Fluidized-Bed Pyrolysis of the Bituminous Sands

Exploratory Process Variable Studies:

P.R. Spring Deposit

Francis V. Hanson and Jay C. Dorius,

ABSTRACT

The fluidized bed pyrolysis of bitumen-impregnated sandstones to produce hydrocarbon liquids has been investigated using the bitumen-impregnated sandstone from the P.R. Spring deposit as feed-sand to the pyrolysis reactor. The P.R. Spring tar sand deposit is the second largest deposit in Utah, and is estimated to contain 4.0-4.5 billion barrels of bitumen in-place. Three distinct tar sand samples, P.R. Spring Rainbow I, P.R. Spring Rainbow II and P.R. Soring South, were used in this investigation because it was found that the physical and chemical properties of the three bitumens were quite different. The locations of the sites from which the three samples were acquired are indicated in Figure 1. The South and Rainbow II bitumens were two orders of magnitude more viscous than the Rainbow I bitumen and the asphaltene content of the South bitumen, 82.8 weight percent, was 60 and 38 percent greater than the asphaltene content of the Rainbow I and Rainbow II bitumens, respectively. The process variables investigated included the pyrolysis reactor temperature and the feed sand retention time in the pyrolysis zone of the reactor. The experimental data were obtained in a continuously fed, fluidized bed reactor operating in a temperature range of 773-923 K at solids retention times that ranged from 17-30 minutes. The reactor feed sand through-put was 2.25 kilograms per hour. The particle size of the feed-sand was 300-600 microns.
Figure 1: Location of Sample Sites from the P.R. Spring Deposit
The hydrocarbon liquid product yield (C\textsubscript{g} yield) generally decreased with increasing reactor temperature. Concomitant with this decline in the liquid yield, there was an increase in the light gas (C\textsuperscript{2}-C\textsuperscript{4}) yield. The liquid product yield increased slightly with decreasing feedsand retention time while the light gas yield decreased. The coke yield was insensitive to process operating variables above 798 K, however, it did appear to be a function of the origin of the feedsand. A maximum liquid yield of 71 weight percent was obtained at a pyrolysis temperature of 798 K and a sand retention time of 20 minutes with the P.R. Spring Rainbow I tar sand. The maximum liquid yield attained with the P.R. Spring South tar sand was 55 weight percent at a pyrolysis temperature of 823 K and a sand retention time of 20 minutes and the maximum liquid yield attained with P.R. Spring Rainbow II tar sand was 61 weight percent at 798 K and a 20 minutes sand retention time. At a fixed pyrolysis reactor temperature and sand retention time the C\textsubscript{g+} liquid product yield correlated quite well with the Conradson carbon residue and with the atomic hydrogen-to-carbon ratio of the native bitumen, that is, the liquid product yield increased with decreasing Conradson carbon residue and with increasing H/C ratio.

The quality of the bitumen derived liquids was significantly improved relative to the native bitumen, that is, the viscosity was reduced by four to five orders of magnitude for the Rainbow I bitumen. The API gravity of the liquid product was 17.6° API an increase of almost 10° API relative to the native bitumen and the pour point decreased from 372 K (210°F) for the native bitumen to 278 K (40°F) for the bitumen derived liquid. The volatility of the
liquid product increased from 31.9 weight percent to 77.2 weight percent relative to the native bitumen. The Conradson carbon residue and asphaltene content of the product liquids were substantially reduced.
INTRODUCTION

A substantial portion of the petroleum processed in the United States is imported, and it is estimated that by 1990 the United States will import 50 percent of its petroleum requirements. This heavy dependence on foreign crude sources can be dangerous, as was evidenced by the 1973 OPEC oil embargo. Uncertainties in the availability of imported crude, along with diminishing domestic petroleum reserves have spurred a great deal of interest in the development of alternate sources of liquid fuels. Bitumen-impregnated sandstones, or "tar sands", represent a significant resource of hydrocarbon liquids which may eventually be utilized as refinery feedstocks and/or feedstock blending components.

There are sixteen major bituminous sand deposits known to exist in the world. These deposits contain approximately 2,100 billion barrels of bitumen in-place, which represents a hydrocarbon resource nearly equivalent to the world’s total known, recoverable petroleum reserves. About 44 percent of the world’s bituminous sand reserves are located on the North American continent.

Bituminous sands were discovered in Alberta, Canada nearly 200 years ago, however, the commercial possibilities of these tar sands were not explored until 1882. Several small-scale operations based on various processing schemes were attempted on the Athabasca tar sand deposit from 1919 to 1967. In 1967, GCOS, Ltd. began operating a commercial processing plant based on the
hot water separation technology 'originally developed by Clark'. A second hot-water plant went on stream in 1978, operated by Syncrude Canada, Ltd. (3,5,7)

In-situ steam injection methods have also been investigated for the recovery of the bitumen from Canadian tar sands. Imperial Oil, Ltd. had intended to recover a bitumen-derived liquid by in-situ steam injection from the Cold Lake deposit in Alberta by 1985. A similar process was investigated by Shell Canada, Ltd. to recover oil from the tar sands of the Peace River deposit. (9)

Above-ground thermal processing also has been proposed as a method for recovering bitumen-derived liquids from the Canadian tar sands. Peterson and Gishler (10,11,12) studied the flash distillation of Athabasca tar sands in an above-ground fluidized bed in which oil yields of up to 86 percent, based on bitumen fed to the reactor, were obtained. The pilot plant was capable of processing 36-38 kilogram per hour of tar sands. The amount of bitumen impregnated sandstone fed in each experiment ranged from 180 to 360 kilograms. Coke yields of 10-13 weight percent based on bitumen fed to the reactor were obtained and the coke was burned to supply process heat. The light gas yield was 5-6 weight percent. The viscosity of the product oil was much lower than the viscosity of the native bitumen and nearly all of the asphaltenes in the native bitumen were absent from the product oil. More recently, the possibility of utilizing the Lurgi-Ruhrgas process to upgrade bituminous sands has been investigated. (13,14) The process, known as "sand cracking," was used previously to produce C^2-C^4 olefins from gasoline or from
lignite. It was found that the process could be used to produce bitumen-derived liquids from the bituminous sands of Alberta.

In the Lurgi-Ruhrgas process, the feed is mixed with the fine-grained heat carrier (burned sand) large a high surface area. Heat transfer between the scent sand and the fresh feed sand occurs within seconds. Coked sand is withdrawn from the reactor and is pneumatically conveyed to the combustion reactor where the coke is burned. The hot clean sand is recycled to the reactor inlet to provide process heat. The Lurgi-Ruhrgas process operated at temperatures of 674-881 K. Feed rates were maintained at 363 kilograms per hour to achieve smooth operation. Liquid hydrocarbon yields of 80-82 weight percent have been reported. The sand retention time determined the eventual product distribution. That is, at longer retention times the gas and coke yields increased at the expense of liquid yield. The major problem with the process in a commercial installation would be the large sized air blowers required for the pneumatic lift pipes.

The bituminous sand reserves of the United Stated are much smaller than the Canadian reserves, and until recently development of these resources was not considered feasible. However, because domestic petroleum resources have diminished, interest in the development of domestic bituminous sands has increased. Approximately 90 percent of the documented, domestic bituminous sands are located in Utah. The Utah bituminous sands represent nearly 20 billion barrels of bitumen in place. This resource exists in four major and numerous minor deposits. Recovery of bitumen and bitumen-derived liquids from four of the major deposits (Asphalt Ridge, Sunnyside, Whiterocks,
and P.R. Spring) has been reported in the literature. The methods being employed were similar to those methods that have proven successful with the Canadian bituminous sands. A modified hot water separation process applicable to the Utah tar sands has been developed by Miller and co-workers. (18,19) The process was modified because of the high viscosity of the Utah bitumens. Nonetheless, the modified hot-water separation process was successful in recovering the bitumen from several Utah tar sands. A pilot plant, based on the modified hot-water technology and operated by Enercor, Inc., was successful in processing 100 tons/day of tar sands from the Asphalt Ridge, P.R. Spring and Whiterocks deposits. The surface tar sand extraction processing schemes were recently reviewed by Irminger and Sjogren.

In-situ recovery of Utah bituminous sands has also been explored. The Shell Oil Company attempted to recover bitumen from the Sunnyside (Utah) deposit by a steam flood method, however, the high viscosity of the bitumen and the resulting lack of its mobility resulted in disappointing yields. The Laramie Energy Technology Center (LETC) conducted in-situ combustion test in the Northwest Asphalt Ridge deposit of Utah. (23,24) The first test demonstrated the feasibility of a reverse combustion scheme to recover oil from bituminous sands. The purpose of the second test was to employ reverse combustion as the preparatory phase followed by forward combustion as the production phase. The hydrocarbon liquid recovered during the second experiment was of a high quality (less viscous, lower pour point) than the product of the first experiment. A steam injection experiment was also conducted at the same site, however, most of the steam and mobilized oil were lost in a lower permeable zone.
Above-ground thermal recovery has been investigated as a method of producing bitumen-derived liquids from the tar sands of the Asphalt Ridge, Sunnyside, Tar Sand Triangle, and Whiterocks deposits (1). Venkatesan constructed a fluidized-bed reactor capable of processing 2.25 kilograms per hour of bituminous sand. Bituminous sands from Tar Sand Triangle and Sunnyside deposits were processed in the temperature range of 675-900 K and solids residence times of 20-32 minutes. Liquid yields of 70 percent by weight of bitumen fed were obtained from the Sunnyside tar sand, whereas liquid product yields from the Tar Sand Triangle deposit were lower (50 percent by weight). Coke yields were insensitive to product variables, however, liquid and gas yield were quite dependent on both reactor temperature and solids retention time. Hydrocarbon liquid yield product decreased with increasing temperature and with increasing solid retention time. Venkatesan observed that there was no clear effect of feedsand particle size on product distribution.

Wang conducted a similar investigation using the Whiterock tar sand as feed and observed similar trends in product distribution with respect to reactor temperature and solids retention time. The maximum liquid yield from the Whiterock tar sand was 70 percent by weight of bitumen fed at a reactor temperature of 793 K and a sand retention time of 16 minutes. The product distribution was found to be insensitive to changes in the velocity of the fluidizing gas. A Box-Wilson multiple linear regression algorithm was used to correlate the liquid yield data. Liquid yields of 80 percent by weight of bitumen fed were predicted for a temperature of 798 K and a solids retention
time of 12 minutes, however, this retention time could not be achieved with the existing reactor system and the correlation has not yet been confirmed.

The objective of the present research project was to continue the investigation of fluidized-bed pyrolysis of Utah bituminous sands. A detailed investigation of the effect of process variables on the product distribution was undertaken. In addition, a more detailed analysis of the products was performed to determine the extent of liquid upgrading that occurred in the process. The third purpose of this investigation was to expand the correlation developed by Wang, which related the hydrocarbon liquid product yield and the primary process variables (reactor temperature and solids retention time). The correlation did not account for variations in properties in the native bitumens from different deposits. Consequently, it was necessary to expand the correlation to account for some property or properties of the native bitumen. Discussion in this paper will be confined to the process variable investigation and preliminary liquid product analyses.

EXPERIMENTAL EQUIPMENT AND PROCEDURES

The reactor used in this investigation was a bench scale fluidized-bed reactor with a diameter of 3.8 cm and a height of 90 cm. The reactor system, shown schematically in Figure 1, has a throughput capacity of 2.25 kilograms per hour. Nitrogen gas was used as the fluidizing and purge gas. Fluidizing gas, controlled by rotameter B, entered the bottom of the reactor through the preheater (C) and flowed upward through the reactor (D). Feed tar sand, which had been crushed to a particle size range of 297-595 microns, was fed into the top of the expansion chamber (E) at a constant rate by a calibrated screw
feeder (G). Nitrogen purge flowed through the feed system to make certain that no hydrocarbon product vapors passed into the screw feeder. The fluidized sand bed was maintained at a constant height by a differential pressure cell (U) which controlled a pneumatic slide valve (S). Any increase in the bed height caused by feed tar sand raining down through the reactor resulted in an increased pressure drop across the bed. As the differential pressure cell sensed this change in the pressure drop, the slide valve was opened, causing coked sand to drain into the receiver (I).

Product vapors were carried from the reactor by the fluidizing and purge gases. Entrained fines were removed by the cyclones (J^, Jo) and the filter (K). Product liquids were collected in the liquid receivers (Nj, ^) and by the cellulose fiber mist eliminators (Qj, Qg). Several light gas samples were collected for analysis by gas chromatography. The coke yield was determined by burning pre-weighed coked sand samples in a muffle furnace at 773 K and then measuring the weight loss of the samples. The mineral matter present in the sand matrix was known to decompose above 923 K and then the weight loss could be attributed to the carbonaceous residue on the sand.

Product liquids were analyzed by the determination of API gravity, viscosity, pour point, simulated distillation, Conradson carbon residue, gradient elution chromatography, elemental analysis and molecular weight. All of these tests were performed on the native bitumens as well, so that the extent of bitumen upgrading could be determined.
RESULTS AND DISCUSSION

The bituminous sands used in this investigation were from the P.R. Spring deposit of eastern Utah. Three distinct samples were used because it was found that the physical and chemical natures of the these three bitumens varied substantially. Properties of the three P.R. Spring tar sands are presented in Tables 1 and 2 and in Figure 2. From Table 1 it can be seen that the bitumen concentrations of the three samples were quite different. The P.R. Spring Rainbow I (PRS-RI) tar sand, which contained 14.14 percent by weight bitumen, required pretreatment with liquid nitrogen before feeding. Because of the high bitumen content, the small particles tended to agglomerate in the feed system. P.R. Spring South (PRS-S) and P.R. Spring Rainbow II (PRS-RII) did not agglomerate and therefore did not require pretreatment in order to be fed. Although the PRS-RI was more difficult to feed, the bitumen was of a higher quality than PRS-S and PRS-RII. That is, the PRS-RI bitumen was more volatile (31.90 wt.%) than PRS-S (14.29 wt.%) or PRS-RII (22.76 wt.%). In addition, the Conradson carbon residue of the PRS-RI was lower than those of the other bitumens. Although the API gravity of the PRS-RI bitumen was the lowest of the three (7.81° vs 8.84° and 11.83°) the bitumen had a lower pour point and was less viscous than the other bitumens. (See Table 2 and Figure 2).

All three bitumens were low in sulfur content, indicating that the bitumens of the P.R. Spring deposit were of fresh water origin. The PRS-S and PRS-RII were high in oxygen content, indicating that they were probably outcrop samples, whereas the relatively low oxygen content of the PRS-RI
Table 1 - Analysis of Native Bitumens from the P.R. Spring (Utah) Tar Sand Deposit

<table>
<thead>
<tr>
<th>Source</th>
<th>Rainbow I</th>
<th>South</th>
<th>Rainbow II</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bitumen Content, wt. %</td>
<td>14.14</td>
<td>6.45</td>
<td>8.53</td>
</tr>
<tr>
<td>Specific Gravity</td>
<td>1.0157</td>
<td>1.0083</td>
<td>0.9872</td>
</tr>
<tr>
<td>API Gravity, °API</td>
<td>7.8</td>
<td>8.8</td>
<td>11.8</td>
</tr>
<tr>
<td>Conradson Carbon, wt. %</td>
<td>14.0</td>
<td>24.0</td>
<td>17.4</td>
</tr>
<tr>
<td>Ash, wt. %</td>
<td>3.3</td>
<td>1.9</td>
<td>1.4</td>
</tr>
<tr>
<td>Simulated Distillation</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Volatility, wt. %</td>
<td>31.9</td>
<td>14.3</td>
<td>22.8</td>
</tr>
<tr>
<td>IBP - 400° F, wt. %</td>
<td>1.3</td>
<td>0.7</td>
<td>0.5</td>
</tr>
<tr>
<td>400 - 650° F, wt. %</td>
<td>5.1</td>
<td>1.3</td>
<td>2.2</td>
</tr>
<tr>
<td>650 - 1000° F, wt. %</td>
<td>25.6</td>
<td>12.3</td>
<td>20.1</td>
</tr>
<tr>
<td>1000° F+, wt. %</td>
<td>68.1</td>
<td>85.7</td>
<td>77.2</td>
</tr>
<tr>
<td>Elemental Analysis</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>C, wt. %</td>
<td>84.66</td>
<td>81.73</td>
<td>81.44</td>
</tr>
<tr>
<td>H, wt. %</td>
<td>11.34</td>
<td>9.29</td>
<td>10.31</td>
</tr>
<tr>
<td>N, wt. %</td>
<td>1.30</td>
<td>1.37</td>
<td>1.37</td>
</tr>
<tr>
<td>S, wt. %</td>
<td>0.53</td>
<td>0.39</td>
<td>0.44</td>
</tr>
<tr>
<td>O, wt. %</td>
<td>1.75</td>
<td>7.22</td>
<td>6.27</td>
</tr>
<tr>
<td>HC Atomic Ratio</td>
<td>1.61</td>
<td>1.36</td>
<td>1.52</td>
</tr>
<tr>
<td>Molecular Weight, g mol</td>
<td>702.</td>
<td>1561.</td>
<td>1381.</td>
</tr>
<tr>
<td>Gradient Elution Chromatography</td>
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<td></td>
</tr>
<tr>
<td>Saturates, wt. %</td>
<td>9.5</td>
<td>4.1</td>
<td>15.8</td>
</tr>
<tr>
<td>MNA/DNA Oils, wt. %</td>
<td>10.2</td>
<td>5.3</td>
<td>3.5</td>
</tr>
<tr>
<td>PNA Oils, wt. %</td>
<td>11.4</td>
<td>1.0</td>
<td>9.0</td>
</tr>
<tr>
<td>Soft Resins, wt. %</td>
<td>13.9</td>
<td>4.0</td>
<td>5.8</td>
</tr>
<tr>
<td>Hard Resins, wt. %</td>
<td>1.1</td>
<td>1.8</td>
<td>2.3</td>
</tr>
<tr>
<td>Polar Resins, wt. %</td>
<td>2.0</td>
<td>1.1</td>
<td>3.6</td>
</tr>
<tr>
<td>Asphaltenes, wt. %</td>
<td>31.3</td>
<td>55.7</td>
<td>35.9</td>
</tr>
<tr>
<td>Noneluted Asphaltenes, wt.%</td>
<td>20.6</td>
<td>27.1</td>
<td>24.1</td>
</tr>
<tr>
<td>Ni, ppm</td>
<td>110.</td>
<td>102.</td>
<td>116.</td>
</tr>
<tr>
<td>V, ppm</td>
<td>10.</td>
<td>8.</td>
<td>3.</td>
</tr>
</tbody>
</table>
Table 2 - Viscosities of the Native Biumens from the P.R. Spring (Utah) Tar Sand Deposit

<table>
<thead>
<tr>
<th>Sample Temperature (°C)</th>
<th>P.R. Spring Rainbow I</th>
<th>P.R. Spring Rainbow II</th>
<th>P.R. Spring South</th>
</tr>
</thead>
<tbody>
<tr>
<td>ion</td>
<td>373</td>
<td>8,269</td>
<td></td>
</tr>
<tr>
<td>120</td>
<td>393</td>
<td>1,816</td>
<td></td>
</tr>
<tr>
<td>140</td>
<td>413</td>
<td>561</td>
<td></td>
</tr>
<tr>
<td>160</td>
<td>433</td>
<td>226</td>
<td></td>
</tr>
<tr>
<td>180</td>
<td>453</td>
<td>161</td>
<td></td>
</tr>
<tr>
<td>200</td>
<td>473</td>
<td></td>
<td>9,000</td>
</tr>
<tr>
<td>210</td>
<td>483</td>
<td></td>
<td></td>
</tr>
<tr>
<td>220</td>
<td>493</td>
<td>2,900</td>
<td></td>
</tr>
<tr>
<td>240</td>
<td>513</td>
<td>1,372</td>
<td></td>
</tr>
<tr>
<td>260</td>
<td>533</td>
<td>795</td>
<td></td>
</tr>
<tr>
<td>280</td>
<td>553</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Figure 2 - Schematic Diagram of the Bench-Scale Fluidized-Bed Pyrolysis Reactor
sample was indicative of a deep-mined sample. Gradient elution chromatography of the native bitumens illustrated further the different natures of the bitumens. The PRS-RI contained more of the less refractory compound types (saturates, naphthenes and aromatics) than the other bitumens. PRS-RII contained fewer of the less refractory compounds than PRS-RI, but contained more of the resins and asphaltenes. The PRS-S bitumen contained much more asphaltenic material than the other bitumens, and was deficient in less refractory compound types. These results concurred with the lower volatility and higher Conradson carbon residue of the PRS-S bitumen. The poorer quality of the PRS-S bitumens was also reflected by its lower H/C ratio.

Pyrolysis of P.R. Spring Rainbow I Tar Sand

The effect of reactor temperatures on the product distribution was studied for the P.R. Spring Rainbow I (PRS-RI) tar sand. The results for a series of experiments wherein the reactor temperature was varied from 773 to 873 K while solids residence time was held constant at 20 minutes are shown in Figure 3. The liquid yield increased as the temperature increased from 773 to 798 K. The maximum liquid yield of 71 percent by weight of bitumen feed was obtained at 798 K. Above 798 K the liquid yield decreased with increasing temperature. The light gas yield increased as the liquid yield decreased. Coke yield was high (25 percent by weight) at the low temperature, but decreased steadily until at 823 K and above, the coke yield stabilized at 9 - 10 weight percent. At the higher temperatures the coke yield was insensitive to changes in the reactor temperature. The observed trends in product distribution can be explained on the basis of the following mechanism. As the tar sand particles entered the reactor and were heated, compounds such as
Figure 3 - Viscosities of Solvent Extracted P.R. Spring Tar Sand Bitumens with Respect to Temperature
paraffins, naphthenes and aromatics, were evaporated from the sand surface.

The resinous fractions were cracked on the sand surface and then evaporated. Asphaltenes decomposed at a fairly slow rate to form coke and vapor. Concurrently, secondary thermal cracking occurred in the vapor phase. The rate of vapor-phase thermal cracking increased faster with increasing temperature than did the rates of evaporation and decomposition of asphaltenes. Thus, at higher temperatures the yield was increased by thermal cracking in the vapor phase. At low temperatures the asphaltenes decomposed at a very slow rate. Therefore, the high coke yield at low temperatures actually consisted partially of coke and partially of high-molecular weight asphaltic material which was soluble in toluene.

The effect of solids residence time on product yield and distribution was also investigated. In this series of experiments the solids residence time was varied from 17 to 30 minutes, while the reactor temperature was kept constant at 798 K. The liquid yield decreased as solids residence time increased, as shown in Figure 4. The decrease was rapid as residence time increased from 17 minutes to 20 minutes, but was less pronounced between solids residence times of 20 and 30 minutes. The gas yield increased steadily as solids residence time increased. The coke yield increased slightly as solids residence time increased from 17 to 20 minutes, then remained essentially constant with further increases in solids residence time.

The liquid yield declined with increasing solids residence time because of the effect of solid-vapor contact time on the rate of thermal cracking in the vapor phase. Slugging can occur in fluidized-bed reactors with an aspect
Figure 4 - Effect of Reactor Temperature on Product Distribution
P.R. Spring Rainbow I Tar Sand
ratio greater than unity ' vapor-phase thermal cracking takes place in the rising slugs. This cracking is enhanced by longer solid-vapor contact times that occur with longer solids retention times. The increased rate of thermal cracking caused an increase in gas yield at the expense of liquid products.

The coke yield was low at low solids residence times because, at such short solid-vapor contact times, thermal cracking is minimized. Thermal cracking accounts for part of the coke yield. At short solids residence times sand particles are not in the reactor long enough for this thermal-cracking product coke to be formed on the sand.

Pyrolysis of P.R. Spring South Tar Sand

The effect of the process variables on the product yields from P.R. Spring, South (PRS-S) tar sand was investigated. In one series of experiments the reactor temperature was varied from 798 K to 923 K while the solids residence time was held constant at 27.1 min. The results for this series of experiments are shown in Figure 5. The trends in product distribution were ivery similar to those observed for P.R. Spring, Rainbow I. The liquid yield increased initially to a maximum of 54 percent by weight of bitumen fed, then declined with further increases in reactor temperature. The trend in gas yield was opposite to that of the liquid yield. The gas yield increased as liquid yield decreased, and decreased as liquid yield increased, however, the coke yield declined slowly and steadily with increasing reactor temperature rather than rapidly declining and then remaining constant as it had for the PRS-RI tar sand.
Figure 5 - Effect of Solids Residence Time on Product Distribution - P.R. Spring Rainbow I Tar Sand
Generally, the liquid yields from PRS-S were lower than those from PRS-RI, while gas and coke yields were higher. In addition, the maximum liquid yield occurred at 823 K instead of at 798 K. No pretreatment of the tar sand feed was required, however, problems were encountered because the coked sand formed a bridge across the reactor during the course of the experiment.

The changes in product yield and distribution and the bridging difficulties were related to the nature of the native bitumen from PRS-S. The PRS-S bitumen is asphaltic in nature. It also has a higher Conradson carbon residue, lower volatility and a lower H/C ratio than PRS-RI. One would expect such a feed material to give a lower yield of liquid products. In addition, the slowly declining trend in the coke yield and the occurrence of the maximum liquid yield at a higher temperature are related to the nature of the PRS-S bitumen. The PRS-S bitumen is more difficult to liberate from the sand surface because of the low volatility. Therefore at lower temperatures the liquid yield is low, and coke yield is high. As the reactor temperature increases the coke yield declines because of an increased rate of decomposition of the asphaltenes, and because there is a relatively small amount of liquid product vapors to undergo thermal cracking to form secondary coke.

The bridging of coked sand in the reactor is believed to be due to the sticky nature of coke precursors on the sand surface. Because the asphaltene content of the native bitumen is high, the amount of the coke precursors on the sand surface is also high. Consequently, the sand particles agglomerate together and eventually bridge across the reactor.
A second series of experiments was performed wherein the solids residence time was varied from 20 min. to 30 min. while the reactor temperature was held constant at 373 K. The effect of the solids residence time on product yields from PRS-S is shown in Figure 6. As with PRS-RI, the liquid yield decreased with increasing solids residence time. The gas yield increased as the liquid yield decreased and the coke yield declined slightly with increasing solids residence time. Once again the liquid yield generally was much lower for PRS-S than for PRS-RI while coke and gas yields were higher. The reason for these general trends have been discussed previously.

Pyrolysis of P.R. Spring Rainbow II Tar Sand

Finally the effect of reactor temperature on product yields for P.R. Spring Rainbow II (PRS-RII) tar sand was investigated and results of these experiments are shown in Figure 7. This series of experiments confirmed the results obtained from earlier work. The effect of reactor temperature on product distribution was similar to the effects observed for PRS-RI and PRS-S. The liquid yield increased to a maximum and then declined as reactor temperature increased. The gas yield increased as the liquid yield decreased and coke yield decreased to a constant level. In addition, the product yields were affected by the nature of the native bitumen in much the same way as the yields of the other bitumens. That is, the yield of liquid from PRS-RII was generally lower than that from PRS-RI but higher than that from PRS-S. Conversely the coke and gas yields from PRS-RII were higher than those from PRS-RI and lower than those from PRS-S. The effect of one property of the native bitumen, namely the Conradson carbon residue (CCR), on product yields
Figure 6 - Effect of Reactor Temperature on Product Distribution
P.R. Spring South Tar Sand
Figure 7 - Effect of Solids Residence Time on Product Distribution - P.R. Spring South Tar Sand
is shown in Figure 8. Product distributions are shown for four separate tar sands: Whiterock (CCR=13.0), P.R. Spring Rainbow I (CCR=14.0), P.R. Spring Rainbow II (CCR=17.4) and P.R. Spring South (CCR=24.0). As the Conradson carbon residue increased, the liquid yield decreased, the gas yield increased and the coke yield increased. Because the Conradson carbon residue is one measure of the nature of hydrocarbons, the data in Figure 8 are indicative of the effect of the nature of the native bitumen on product yields and distribution. Definite trends are also observed in product distribution with respect to the H/C ratio and the asphaltene content of the native bitumen.

QUALITY OF LIQUID PRODUCTS

Properties of the liquid products from the pyrolysis of P.R. Spring Rainbow I, P.R. Spring South, and P.R. Spring Rainbow II are presented in Tables 3, 4 and 5, respectively, along with the properties of the native bitumens. It can be seen that in all cases, the API gravity and volatility of the product liquids were increased over the native bitumens while the viscosity, Conradson carbon residue, ash content, and pour point were markedly decreased. In addition, the amount of preferred components (saturates, mono-, di- and polynuclear aromatics) in the product liquids were increased at the expense of asphaltenes. The amount of saturates in the liquid product from P.R. Spring Rainbow II was actually decreased due to thermal cracking of those compounds during pyrolysis, however, the total amount of lighter compounds in the products liquid was increased over the native bitumen.

The differences in liquid properties between the pyrolysis liquid
Figure 8 - Effect of Reactor Temperature on Product Distribution - P.R. Spring Rainbow II Tar Sand
Figure 9 - Product Distribution as a Function of the Conradson Carbon Residue of the Native Bitumen
Table 3 - Analysis of the Liquid Product from the Fluidized-Bed Pyrolysis of the P.R. Spring Rainbow I Tar Sand

<table>
<thead>
<tr>
<th>Property</th>
<th>P.R. Spring</th>
<th>Pyrolysis Liquid Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor Temperature, K</td>
<td>823</td>
<td></td>
</tr>
<tr>
<td>Solids Residence Time, min.</td>
<td>20</td>
<td></td>
</tr>
<tr>
<td>Fluidizing Gas Velocity, LPH (STP)</td>
<td>144</td>
<td></td>
</tr>
<tr>
<td>Product Yields, wt. %</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>20.1</td>
<td></td>
</tr>
<tr>
<td>Liquid</td>
<td>70.4</td>
<td></td>
</tr>
<tr>
<td>Coke</td>
<td>9.5</td>
<td></td>
</tr>
<tr>
<td>API Gravity, °API</td>
<td>7.8</td>
<td>17.3</td>
</tr>
<tr>
<td>Viscosity, cps</td>
<td>8268.8 (0 373 K)</td>
<td>-</td>
</tr>
<tr>
<td>Pour Point, K (°F)</td>
<td>372.1 (210 °F)</td>
<td>274.7 (35 °F)</td>
</tr>
<tr>
<td>Conradson Carbon Residue, wt.%</td>
<td>14.0</td>
<td>4.25</td>
</tr>
<tr>
<td>Ash, wt.%</td>
<td>3.3</td>
<td>0.01</td>
</tr>
<tr>
<td>Simulated Distillation</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Volatility, wt. %</td>
<td>31.9</td>
<td>79.0</td>
</tr>
<tr>
<td>IBP - 400° F, wt. %</td>
<td>1.3</td>
<td>2.7</td>
</tr>
<tr>
<td>400 - 650° F, wt. %</td>
<td>5.1</td>
<td>23.1</td>
</tr>
<tr>
<td>650 - 1000° F, wt. %</td>
<td>25.6</td>
<td>53.2</td>
</tr>
<tr>
<td>1000°F plus Residue, wt. %</td>
<td>68.1</td>
<td>21.0</td>
</tr>
<tr>
<td>Gradient Elution Chromatography</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Saturates, wt.%</td>
<td>9.5</td>
<td>13.9</td>
</tr>
<tr>
<td>MNA/DNA Oils, wt.%</td>
<td>10.2</td>
<td>32.9</td>
</tr>
<tr>
<td>Soft Resins, wt.%</td>
<td>13.9</td>
<td>13.7</td>
</tr>
<tr>
<td>Hard Resins, wt.%</td>
<td>1.1</td>
<td>1.6</td>
</tr>
<tr>
<td>Polar Resins, wt.%</td>
<td>2.0</td>
<td>2.3</td>
</tr>
<tr>
<td>Asphaltenes, wt.%</td>
<td>31.3</td>
<td>16.8</td>
</tr>
<tr>
<td>Non-Eluted Asphaltenes, wt.%</td>
<td>20.6</td>
<td>1.3</td>
</tr>
</tbody>
</table>

* Normalized to 100%
### Table 4 - Analysis of the Liquid Product from the Fluidized-Pyrolysis of the P.R. Spring South Tar Sand

<table>
<thead>
<tr>
<th>Property</th>
<th>Pyrolysis South</th>
<th>Liquid Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor Temperature, K</td>
<td>823</td>
<td></td>
</tr>
<tr>
<td>Solids Residence Time, min.</td>
<td>20</td>
<td></td>
</tr>
<tr>
<td>Fluidizing Gas Velocity, LPH (STP)</td>
<td>144</td>
<td></td>
</tr>
<tr>
<td>Product Yields, wt. %</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>30.6</td>
<td></td>
</tr>
<tr>
<td>Liquid</td>
<td>49.3</td>
<td></td>
</tr>
<tr>
<td>Coke</td>
<td>20.1</td>
<td></td>
</tr>
<tr>
<td>P.R. Spring</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Property</td>
<td></td>
<td></td>
</tr>
<tr>
<td>API Gravity, °API</td>
<td>8.8</td>
<td>10.2</td>
</tr>
<tr>
<td>Viscosity, cps</td>
<td>10,071. (e 373 K)</td>
<td>548.5 ( @ 298 K)</td>
</tr>
<tr>
<td>Pour Point, K (°F)</td>
<td>433. (320 '°F)</td>
<td>283. (50 '°F)</td>
</tr>
<tr>
<td>Conradson Carbon Residue, wt.%</td>
<td>24.0</td>
<td>7.2</td>
</tr>
<tr>
<td>Ash, wt.%</td>
<td>1.9</td>
<td>0.01</td>
</tr>
<tr>
<td>Simulated Distillation</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Volatility, wt. %</td>
<td>14.3</td>
<td>97.7</td>
</tr>
<tr>
<td>IBP - 400° F, wt. %</td>
<td>0.7</td>
<td>4.9</td>
</tr>
<tr>
<td>400 - 650° F, wt. %</td>
<td>1.3</td>
<td>36.4</td>
</tr>
<tr>
<td>650 - 1000° F, wt. %</td>
<td>12.3</td>
<td>56.4</td>
</tr>
<tr>
<td>1000°F plus Residue wt. %</td>
<td>85.7</td>
<td>2.3</td>
</tr>
<tr>
<td>Gradient Elution Chromatography</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Saturates, wt.%</td>
<td>4.1</td>
<td>7.4</td>
</tr>
<tr>
<td>MNA DNAs Oils, wt.%</td>
<td>5.3</td>
<td>35.6</td>
</tr>
<tr>
<td>PNA Oils, wt.%</td>
<td>1.0</td>
<td>3.9</td>
</tr>
<tr>
<td>Soft Resins, wt.%</td>
<td>4.0</td>
<td>17.8</td>
</tr>
<tr>
<td>Hard Resins, wt.%</td>
<td>1.8</td>
<td>2.4</td>
</tr>
<tr>
<td>Polar Resins, wt.%</td>
<td>1.1</td>
<td>2.7</td>
</tr>
<tr>
<td>Asphaltenes, wt.%</td>
<td>55.7</td>
<td>26.4</td>
</tr>
<tr>
<td>Non-Eluted Asphaltenes, wt.%</td>
<td>27.1</td>
<td>3.8</td>
</tr>
</tbody>
</table>

* Normalized to 100%
Table 5*—Analysis of the Liquid Product from the Fluidized-Bed Pyrolysis of the P.R. Spring Rainbow II Tar Sand

<table>
<thead>
<tr>
<th>Property</th>
<th>P.R. Spring Rainbow II</th>
<th>Liquid Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor Temperature, K</td>
<td>823</td>
<td></td>
</tr>
<tr>
<td>Solids Residence Time, min.</td>
<td>20</td>
<td></td>
</tr>
<tr>
<td>Fluidizing Gas Velocity, LPH (STP)</td>
<td>144</td>
<td></td>
</tr>
<tr>
<td>Product Yields, wt. %</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>20.6</td>
<td></td>
</tr>
<tr>
<td>Liquid</td>
<td>60.5</td>
<td></td>
</tr>
<tr>
<td>Coke</td>
<td>18.9</td>
<td></td>
</tr>
<tr>
<td>P.R. Property</td>
<td></td>
<td></td>
</tr>
<tr>
<td>API Gravity, °API</td>
<td>11.8</td>
<td>18.3</td>
</tr>
<tr>
<td>Viscosity, cps</td>
<td>9,000. ( @ 473 K)</td>
<td>158.4 ( @ 298 K)</td>
</tr>
<tr>
<td>Pour Point, K (°F)</td>
<td>433. (320 °F)</td>
<td>271.9 (30 °F)</td>
</tr>
<tr>
<td>Conradson Carbon Residue, wt.%</td>
<td>17.4</td>
<td>3.6</td>
</tr>
<tr>
<td>Ash, wt.%</td>
<td>1.4</td>
<td>0.01</td>
</tr>
<tr>
<td>Simulated Distillation</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Volatility, wt. %</td>
<td>22.8</td>
<td>82.9</td>
</tr>
<tr>
<td>IBP - 400°F, wt. %</td>
<td>0.5</td>
<td>7.6</td>
</tr>
<tr>
<td>400 - 650°F, wt. %</td>
<td>2.2</td>
<td>23.1</td>
</tr>
<tr>
<td>650 - 1000°F, wt. %</td>
<td>20.1</td>
<td>52.2</td>
</tr>
<tr>
<td>1000°F plus Residue, wt. %</td>
<td>77.2</td>
<td>17.1</td>
</tr>
<tr>
<td>Gradient Elution Chromatography</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Saturates, wt.%</td>
<td>15.8</td>
<td>13.4</td>
</tr>
<tr>
<td>MNA/DNA Oils, wt.%</td>
<td>3.5</td>
<td>37.1</td>
</tr>
<tr>
<td>RNA Oils, wt.%</td>
<td>9.0</td>
<td>5.7</td>
</tr>
<tr>
<td>Soft Resins, wt.%</td>
<td>5.8</td>
<td>17.4</td>
</tr>
<tr>
<td>Hard Resins, wt.%</td>
<td>2.3</td>
<td>3.5</td>
</tr>
<tr>
<td>Polar Resins, wt.%</td>
<td>3.6</td>
<td>2.6</td>
</tr>
<tr>
<td>Asphaltenes, wt.%</td>
<td>35.9</td>
<td>15.7</td>
</tr>
<tr>
<td>Non-Eluted Asphaltenes, wt.%</td>
<td>24.1</td>
<td>4.6</td>
</tr>
</tbody>
</table>

* Normalized to 100%
products and the native .bitumens represent a substantial improvement in liquid product quality over the native bitumen. Product liquids are amenable to transportation by pipeline because of their relatively low viscosity, pour point, and API gravity. The relatively high volatilities, along with the low asphaltene contents and Conradson carbon residues of the product liquids indicate that they could be upgraded to refinery feedstock quality by mild hydrotreating.

CONCLUSIONS

Several conclusions can be drawn from this investigation:

Fluidized bed pyrolysis is a viable method for producing a pipeline quality hydrocarbon liquid that will be compatible with modern petroleum refineries after mild hydrotreatment.

Maximum liquid yields were obtained at a solids retention time of 20 minutes and a temperature of 798 K for P.R. Spring Rainbow I and P.R. Spring Rainbow II. The maximum liquid yield for P.R. Spring South occurs at 823 K and solids retention times of 20 minutes. It is projected that the liquid yield could be increased to 80 weight percent for PRS-RI by reducing the solids retention time to 12 minutes.

Coke yield is insensitive to reactor temperature above 798 K and is relatively insensitive to changes in solids residence time. Increases in coke yield at lower reactor temperatures are probably due to unconverted high molecular weight asphaltenic material which
cannot be vaporized from the sand surface at the pyrolysis temperature.

There is a definite relationship between certain properties of the native bitumen and the product distribution. Bitumens with lower Conradson carbon residue, higher volatility, higher H/C ratio, and lower asphaltene content yield more liquid, less gas and less coke than lower quality bitumens. The API gravity does not appear to correlate with product distribution.

Substantial liquid upgrading occurs during the pyrolysis process. The product liquids are of pipeline quality, and can be upgraded to refinery quality by mild to moderate hydrotreatment.
II. LARGE DIAMETER FLUIDIZED BED PYROLYSIS UNIT

INTRODUCTION

The fluidized bed technique represents a unique method for thermal processing of tar sands to obtain a bitumen-derived hydrocarbon liquid which can serve as a hydrogen refinery feedstock. The influence of process operating variables on the product distribution and the yields have been investigated in a small scale (1 V2 inch diameter) fluidized bed reactor. These exploratory process variable studies indicated that the feed sand retention time in the pyrolysis zone significantly influenced the C₃− hydrocarbon liquid yield, that is, the C⁴ liquid yield increased with decreasing feed sand retention time. Unfortunately, the feed sand throughput for the small scale reactor is limited by the quality of fluidization at short residence times, that is, excessive slugging of the bed occurs at feed sand retention times below 16-20 minutes. Further development of this type of process requires the design and construction of a larger scale laboratory apparatus (4-6 inch diameter) than the small scale apparatus presently in use. The larger diameter reactor will permit an expansion of the range of feed sand retention times to better define the influence of this variable on the product distribution and product yields from the fluidized bed.

Successful operation of the large diameter reactor should provide scale up data for engineering design purposes. The large diameter reactor must be capable of meeting the following conditions:
• operate continuously for 48 hours or longer;
  
• accept feed sand particle size ranges from 0.032 inches (20 mesh) to 0.25 inches; and
• produce bitumen-derived hydrocarbon liquids at a rate of at least 300 g h⁻¹.

In the second operational phase the system will be required to operate in a gas recycle mode and to operate with steam and/or carbon dioxide injection into the pyrolysis zone of the reactor. A third phase will incorporate combustion of the coke on the spent sand and recirculation of a portion of the hot clean sand to the pyrolysis reactor to supply the energy required in the bitumen pyrolysis reactor.

Nitrogen will be used as the fluidization gas in the initial phase of the experimental program. The process operating variables to be investigated includes the sand retention time which will range from four minutes to 20 minutes, the pyrolysis reactor temperature which will range from 673 K to 873 K, and fluidizing gas velocity which will range from one to five times the minimum fluidization velocity. The fluidized bed will operate at atmospheric pressure. The standard operating conditions which will form the basis for comparison of the response of the process to the operating variables, the mode of operation and the source of the feed sand are as follows:

• feed sand retention time of 10 minutes
• pyrolysis reactor temperature of 723 K;
• fluidizing gas velocity of three times the minimum fluidization velocity; and;
• pyrolysis reactor sand inventory of seven kg.
A schematic of the apparatus is presented in Figure 10. The 4-5 inch diameter reactor system is essentially a large scale version of the small diameter fluidized bed pyrolysis apparatus. The following components of the system have been designed and fabricated:

- reactor;
- recycle gas blower;
- solid - vapor separation train;
  - cyclone
  - filter
  - condenser
- aerosol/mist removal system; and;
- solid feed system;
  - feed sand hopper
  - feed sand screw feeder.

Details of the design calculations are included in the following sections.
Figure 10 - Schematic of Large Diameter Fluidized Bed Pyrolysis Reactor System
1. Specification of Fluidized Bed Reactor Dimensions

It was assumed in all calculations that the product distribution and yields obtained with the small diameter reactor would also be obtained with the large diameter reactor for feed sands of the same particle size.

A number of scale-up techniques were employed in these calculations, however, a complete review of all derivations and computations is beyond the scope of this report. The following relations were used to scale the dimensions of the reactor:

\[
M_{\text{il}} = \frac{2.92 \cdot \rho}{\phi_{\text{ib}}},
\]

\[
\frac{m}{V} = 8 \cdot 1^2 \cdot \frac{\xi}{\nu(T) U_{\text{mb}}},
\]

where \( A \) is the cross sectional area of the reactor;

\( V \) is the volume of the reactor;

\( D \) is the inside diameter of the reactor;

2 indicates the large diameter reactor; and;

1 indicates the small diameter reactor.

The results of the scale up for the pyrolysis reactor are summarized in Table 6.
Table 6 - Fluidized Bed Pyrolysis Reactor Dimensions

<table>
<thead>
<tr>
<th>Reactor</th>
<th>Material of Construction</th>
<th>I.D.</th>
<th>O.D.</th>
<th>A</th>
<th>Length</th>
</tr>
</thead>
<tbody>
<tr>
<td>Large Diameter</td>
<td>10.2cm 40-S Pipe, S.S. 304</td>
<td>10.1</td>
<td>11.4</td>
<td>83.1 cm² (12.7 in²)</td>
<td>132.7 cm (52.3 in)</td>
</tr>
<tr>
<td>Small Diameter</td>
<td>3.2 cm 40-S Pipe S.S. 304</td>
<td>3.5</td>
<td>4.2</td>
<td>9.7 cm² (1.5 in²)</td>
<td>90. cm (35.4 in)</td>
</tr>
</tbody>
</table>
2. Minimum Fluidization Velocity

The recycle compressor or blower can only be sized after specification of the range of fluidizing gas velocities/flow rates has been established. Calculation of the fluidized gas velocities was based on a reactor inventory of seven kilograms of bitumen-imregnated sandstone for a range of feed sand particle sizes from 20 mesh through 0.64 cm (0.25 inch) diameter. The particle Reynolds number is given by

\[ \frac{d_p V_m P}{\mu} \]

where \( d_p \) is the equivalent of the feed seed particle, cm;

\( P \) is the fluidizing gas density at the operating temperature and pressure, g cm\(^{-3} \);

is the fluidizing gas viscosity at the operating temperature and pressure, g cm\(^{-5} \);

\( V_m \) is the minimum fluidizing gas velocity, cm\(^{-1} \);

The fluidizing gas velocity may be calculated for the various particle sizes from the following equation

\[ \frac{d_p V_m P}{\mu} = \left(33.7 \right)^{\frac{2}{3}} + 0.048 \left( \frac{P s \mu}{P g} \right)^{\frac{3}{2}} \left( \frac{V_m}{g} \right)^{-33.7} \]

where \( P_s \) is the density of the solid particle, g cm\(^{-3} \); and;

\( g \) is the gravitational constant.

The fluidized bed pyrolysis reactor operating conditions and the
appropriate fluid properties used to compute the minimum fluidization gas velocity were as follows:

\[ T = 823 \text{ K}; \]

\[ P_{\text{ave}} = 16.85 \text{ psia}; \]

\[ y = 0.00037 - \frac{3}{\text{cm sec}} \text{ at } 823 \text{ K; } 16.85 \text{ psia}; \]

\[ p = 4.75 \times 10^{-4} \text{ cm}^{-3} \text{ at } 823 \text{ K; } 16.85 \text{ psia}; \]

\[ \rho = 2.7 \text{ g cm}^{-3} \text{ at } 823 \text{ K; } 16.85 \text{ psia}; \]

The results of these calculations are presented in Table 7.

3. Pressure Drop Across Fluidized Bed and Gas Recycle Blower Power Requirements

The pressure drops across a fluidized bed is given by

\[ \Delta P = \frac{m g}{\rho A} \]

where \( m \) is the mass of the particle inventory in the fluidized bed, kg;

\( A \) is the cross sectional area of the reactor, \( \text{cm}^2 \), and;

\( g, g_c \) are gravitational constants.

If the mass of feed sand and coked sand inventory is assumed to be seven kilograms then the pressure drop would be 1.2 psi, however, at actual operating conditions the estimated pressure drop would be 1.7 psi. The total pressure drop through the system is estimated to be equivalent to three psi. Thus for a blower efficiency of 0.7 when fluidizing 20 mesh particles at the
<table>
<thead>
<tr>
<th>Feed Sand Diameter</th>
<th>U_m (cms)</th>
<th>Mass Flow Rate (kg h⁻¹)</th>
<th>Volumetric Flow Rate (1 h⁻¹STP)</th>
<th>Reₜ</th>
</tr>
</thead>
<tbody>
<tr>
<td>20 mesh</td>
<td>28.7</td>
<td>4.0</td>
<td>3,226</td>
<td>3.1</td>
</tr>
<tr>
<td>1/16 inch</td>
<td>86.5</td>
<td>12.2</td>
<td>9,724</td>
<td>17.5</td>
</tr>
<tr>
<td>1/8 inch</td>
<td>198.</td>
<td>27.8</td>
<td>22,259</td>
<td>81.</td>
</tr>
<tr>
<td>1/4 inch</td>
<td>341.</td>
<td>47.9</td>
<td>38,336</td>
<td>278.</td>
</tr>
</tbody>
</table>
minimum fluidization velocity a 27 watt blower would be required. The blower power requirements for the range of particle sizes and for the range of fluidizing gas velocities are presented in Table 8. Thus a 2000 watt gas recycle blower should have the capacity to cover the entire range of operating variables proposed for the large diameter fluidized bed reactor.

4. Particle Entrainment

The disengagement of sand particles from the vapor in the expansion chamber requires that the terminal particle velocity be greater than the fluid velocity in the expansion chambers. For a spherical particle the particle Reynolds number is given by

\[ \text{Re}_p = \frac{d_p V}{\rho_p \mu} \]

where \( V_t \) is the terminal velocity of the particle, cm s\(^{-1}\).

The terminal velocity is given by the following equation (32), depending upon the magnitude of the particle Reynolds number:

\[ V_t = \frac{g(p - p_i) d}{\rho_i g Y} \quad \text{for} \quad \text{Re}_p < 0.4 \]

\[ V_t = \frac{4^{(0.5)} (\rho_i - \rho_g) V^2}{225 \rho_U} \quad \text{for} \quad \text{Re}_p < 50 \]

At a fluidization gas rate 1.5 times the minimum fluidization flow rate for the 20 mesh particles the minimum fluidization gas velocity is 28.7 cm s\(^{-1}\) chamber pressure of 16 psi and a temperature of 673 K.

The area ratio for the expansion chamber and the reactor was assumed to
Table 8 - Recycle Gas Blower Power Requirements

<table>
<thead>
<tr>
<th>Feed Sand Equivalent Diameter</th>
<th>1.5</th>
<th>2.0</th>
<th>3.0</th>
<th>4.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>20 øtesh</td>
<td>41</td>
<td>54</td>
<td>81</td>
<td>108</td>
</tr>
<tr>
<td>1/16 inch</td>
<td>123</td>
<td>164</td>
<td>246</td>
<td>328</td>
</tr>
<tr>
<td>1/8 inch</td>
<td>281</td>
<td>375</td>
<td>562</td>
<td>749</td>
</tr>
<tr>
<td>1/4 inch</td>
<td>484</td>
<td>645</td>
<td>968</td>
<td>1291</td>
</tr>
</tbody>
</table>
be two:

A expansion chamber
A reactor tube

(7)

Thus the terminal particle velocity in this expansion chamber would be 21.5 cm s$^{-1}$ and the maximum equivalent diameter of the entrained particles would be 72.6 microns. The computed diameters for the maximum particle size entrained for the various feed sand particle sizes and presented in Table 9. These particle diameters were used to design the cyclones used to remove entrained sand finer.

5. Cyclone Design

The sand granules/particles in the fluidized bed will undergo attrition in the bed and elutriation from the bed. If the minimum sand grain size comprising the granules/particles is assumed to be five microns in diameter then the cyclones must be designed assuming, $D_{p,min} = 5 \times 10^{-6}$ cm.

The cyclone design equation was used to specify the cyclone inlet velocity and the number of cyclones required.

$$f_9 \quad Be \quad -i''\times5$$

$$V_{min} = \frac{[iN_1V_c^{p}S^{p}q]}{J}$$

where $D_{0,min}$ is the minimum diameter of particle which is completely collected, cm;

$V$ is the cyclone inlet velocity, cm s$^{-1}$;

$\nu$ is the viscosity of fluid at the inlet to the cyclone, g cm$^{-1}$ s$^{-1}$,
Table 9 - Entrained Particle Diameters

<table>
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<th>Feed Sand Equivalent Diameter</th>
<th>Maximum Particle Entrained</th>
<th>Reₚ</th>
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<tr>
<td>20 mesh</td>
<td>72.6 micron</td>
<td>0.24</td>
</tr>
<tr>
<td>1/16 inch</td>
<td>75.7 micron</td>
<td>0.75</td>
</tr>
<tr>
<td>1/8 inch</td>
<td>173. micron</td>
<td>3.93</td>
</tr>
<tr>
<td>1/4 inch</td>
<td>299. micron</td>
<td>11.66</td>
</tr>
</tbody>
</table>
N\textsubscript{c} is the number of turns made by the gas stream in cyclone;
generally, N\textsuperscript{^\textsubscript{c}} = 5, and;

B\textsubscript{c} is the width of rectangular cyclone inlet duct, cm.

If particles over five microns are to be effectively collected by the cyclones then the cyclone inlet velocity must be 489 cm s\(^{-1}\) when nitrogen is used as the fluidizing gas and the cyclone temperature and pressure are 673 K and 16 psi, respectively. If 0.64 cm particles are used in the fluidized bed then the cyclone inlet velocity would be 5321 cm s\(^{-1}\) to collect five micron sand grains entrained from the reactor. Sound design concepts indicate that four cyclones in parallel, each have a cyclone inlet velocity of 1330 cm s\(^{-1}\), will be sufficient to collect sand finer elutriated from the pyrolysis reactor.

The detailed schematic of the cyclone is presented in Figure 11 and the design for the cyclone holder is presented in Figure 12.

Entrained particles having a diameter less than five microns which pass through the cyclone system will be removed from the vapor stream with a stainless steel filter (Figure 13).

6. Condenser Heat Transfer Area

The basic heat transfer equation used to size the condenser was given by

\[
q_t = \frac{UA}{\ln \left( \frac{T_1}{T_2} \right)}
\]

where \( q_t \) is the rate of heat transfer in the condenser, Kj h\(^{-1}\),
\( U \) is the overall heat transfer coefficient, K cm\(^{-1}\)h\(^{-1}\),
\( A \) is the total area of heat transfer surface, cm\(^2\), and
\( AT \) is the logarithmic temperature difference, K.

The logarithmic mean temperature difference is given by
CYCLONE, FAN

2.54cm Tube
(1" BWG 16)
I.D. = 2.2 cm
O.D. = 2.5 cm

3.81 cm Tube
(11/2" BWG 16)
I.D. = 3.5 cm
O.D. = 3.8 cm

6.35 cm Pipe
(2 1/2" 5S)
I.D. = 6.9 cm
O.D. = 7.3 cm

Thickness about 2 mm

Figure 11 - Cyclone
Figure 12 - Cyclone Holder
FILTER

12.7 cm

2.54 cm BWG Tube
O.D. = 2.54 cm
I.D. = 2.21 cm

0.32 cm Thermal Couple Well

8.9 cm ~3'

7 cm

Thickness about 0.15 cm 35.6 cm

24 cm

1.905 cm Male NPT
Threaded

1.905 cm Female NPT

figure 13 - Filter Assembly
The calculation of the heat duty for the hydrocarbon product recovery condenser was based on the following assumptions:

- enthalpy of evaporation/condensation of the \( C_x \) hydrocarbon liquid was assumed to be 335 KJ g\(^{-1}\) (80 cal g\(^{-1}\));
- feed sand flow rate was assumed to be 42 kg h\(^{-1}\) at a retention time of 10 minutes;
- feed sand bitumen saturation was assumed to be eight weight percent;
- product distribution based on bitumen fed to the reactor was assumed to be 80 weight percent \( C_x \) liquid, 10 weight percent \( C_1-C_4 \) gas and 10 weight percent carbonaceous residue on the spent sand; and;
- fluidization gas was assumed to be nitrogen.

Thus the \( C_x \) hydrocarbon liquid produced per hour would be approximately 2.7 kilograms. Based on the small diameter reactor operating experience approximately 10 percent of the produced hydrocarbon liquid was recovered in the condenser. Thus it was anticipated that approximately 300 grams of \( C_x \) hydrocarbon liquid would condense per hour. The fluidizing gas velocity
was assumed to be 1.2 times the minimum fluidization gas velocity for 0.64 cm diameter particles and the nitrogen-hydrocarbon vapor was assumed to enter the condenser at 673 K. At these conditions the man flux of nitrogen through the condenser was 8.45 kg m\(^{-2}\) s \(^{-1}\) (mass flow rate: 47.9 kg h\(^{-1}\)). The mass of uncondensed vapor passing through the condenser was assumed to be 2400 g h\(^{-1}\) and was conservatively assumed to have a specific heat of one calorie per gram. Thus the heat duty of the condenser was calculated to be 25,000 KJ h\(^{-1}\) (6000 kcal h\(^{-1}\)).

The overall heat transfer coefficient can be expressed as

\[
U = \frac{1}{\frac{1}{h_n} + \frac{x_w}{h_o} + \frac{1}{h_t}} \tag{ii}
\]

where \(h_n\) is the convection film coefficient for flow of water through a tube, K J m\(^{-2}\) K \(^{-1}\) h\(^{-1}\);

\[x_w\] Tube wall thickness
\[K\] Thermal conductivity of tube wall, m\(^2\) K (KJ)\(^{-1}\); and;

\(h_t\) is the convection film coefficient for flow of nitrogen through a tube, K J m\(^{-2}\) K \(^{-1}\) h\(^{-1}\).

The convection film coefficient for the flow of water through a tube was estimated \(^{34}\) to be 2,900 K J m\(^{-2}\) K \(^{-1}\) h\(^{-1}\) (695 heal m\(^{-2}\) K \(^{-1}\) h\(^{-1}\)). The ratio of the tube wall thickness to the rates of the tube wall thickness to the thermal conductivity of the tube wall for thin walled steel pipe in approximately 32 x 4 \(\approx\) 4 m m\(^{-2}\) K \(^{-1}\) h\(^{-1}\) (KJ) \(^{-1}\) (0.8m\(^2\) Kh (kcal) \(^{-1}\)). The man flux of nitrogen through the condenser is calculated to be 8.45 kg m\(^{-2}\) s \(^{-1}\). Then the convection film
Therefore the overall heat transfer coefficient would be approximately 90 KJm K h^-1 (22 heat m K h^-1). The condensation of hydrocarbon vapor in the condenser will result in an increase in h^ and consequently an increase in U. Then the final value of U was increased by a factor of one-third to account for vapor condensation and U = 120 KJm k h^-1 was used to compute the required heat transfer area. The inlet and out temperature for the condenser are indicated below:

AT₄ 370K  Water, 303K < nitrogen, oil vapor, 673K
AT₅ 35K
AT₆ = 142K  Water, 303K

The heat transfer surface area was calculated to be 14,700 cm². Thus three condensers each 80 cm in length will be required to condense the hydrocarbon vapor produced in the fluidized bed reactor. A schematic of the condenser is presented in Figure 14. In addition to the condenser train a demister will be required to recover the condensible hydrocarbon vapor that exits the reactor in the form of an aerosol. A schematic of the demister is presented in Figure 15.

7. Feed Sand Hopper and Screw Feeder

The fluidized bed reactor solids inventory was assumed to be seven kilograms at the standard operating conditions. At a feed sand retention time of 10 minutes the feed sand feed rate would be 42 kilograms per hour. Under ordinary circumstances an experiment would be expected to last for 6-8 hours thus the mass of feed sand required would be 250 to 350 kilograms. If the bulk density of the feed sand is 1.4 g cm⁻³ (1400 kg m⁻³) the volume of feed sand
CON DENSER

6.35cm to 2.54cm Bell Reducer

1.0cm

1.2cm

81cm

8.89cm Pipe
(3 1/2" 5S)
0.0. = 10.2cm
1.0. = 9.7cm.

6.35 cm Pipe
(2 1/2" 5S)
O.D. = 7.3cm
I.D. = 6.9cm

3.81 cm Pipe
(1 1/2" 5S)
O.D. = 4.8cm
I.D. = 4.5cm

1.0cm

81cm

Figure 14 - Condenser Assembly
C54
MIST COLLECTOR

Figure 15 - Demister

C55
required would be 0.2 to 0.25 cubic meters. Thus the hopper was sized to hold 0.2 cubic meters of feed sand. Schematics of the hopper and hopper cover assembly are presented in Figure 16 and 17. The feed sand screw feed is presented in Figure 18.

The components of the apparatus are being fabricated and will be assembled in the near future.
Figure 16 - Feed Sand Hopper
Figure 17 - Feed Sand Hopper Cover
SCREW FEEDER
S T A I N L E E S STEEL

304

38 cm
36 cm
19cm
•
15cm—H
lem
6,1.3cm I.D. Bolt
Holes
10cm
- — 8 . 9 cm. Pipe
(3 1/2" 40S)

en
U3

>>>>

y

f—r

4.6cm

•a#

0.64 cm Female NPT

/ r j > / / / r ? i / / > r > r r > r

\

19cm

1.9 6,1.3cmI.D.
1 cm cm Bolt Holes

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H> '-Ooni

25cm

jj

15cm 19cm

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15cm6,1.3cm I.D. Bolt Holes

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8.9 cm Pipe
(3 1/2" 10S)

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7.6cm Pipe —
(3" 40S)

-\.

6,1.3cm I.D. Bolt I
Holes
15cmFigure 18 - Feed Sand Screw Feeder Assembly

•19cm-

1.3cm


Legend for Bench-Scale Fluidized-Bed Pyrolysis Reactor

A. Nitrogen Cylinder
B. Rotameter
C. Fluidizing Gas Preheater
D. Fluidized-Bed Reactor
E. Expansion Chamber
F. Feed Sand Hopper
G. Screw Feeder
H. Motor Drive
I. Spent Sand Receiving Vessel
J. Cyclones
K. Filter Housing
L. Water Condenser
M. Hydro eye 1 one
U. Liquid Product Receivers
0. Glass Jar
P. Pressure Guages
Q. Cellulose Fiber Mist Capture Tubes
R. Gas Regulators
S. Solids Valve
T. Thermocouples
U. Differential Pressure Cell
V. Valves
W. Motor Drive Controller
REFERENCES


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SECTION D

ENERGY RECOVERY IN THERMAL PROCESSING OF UTAH TAR SANDS
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### Section D

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A  Cross-section area of bed, ft

D(z)  Z transform of controller transfer function

E(t)  Error 'signal in a continuous form

E(z)  Z transform of error signal

E_n  The ith error signal in a discrete form

E_n  Error signal in a discrete form

f  Weight percent of bitumen contained in the tar sands

F_0  Feed rate of tar sands, lbs/hr

F_1  Mass flow rate- of tar sands from the pyrolysis reactor, lbs/hr

F''  Mass flow rate of tar sands from the combustion reactor, lbs/hr

2

g  = 32.174 ft/sec , acceleration of gravity

2

2
g  = 32.174 lbm*ft/lbf * (sec) , conversion factor

G  Transfer function in the control loop

G(s)  Laplace transform of the ith process transfer function

G(z)  Z transform of transfer function

H(s)  Laplace transform of hold transfer function

K  Proportional gain

K_i  The ith control valve constant
\( K_T \) Process gain of temperature control system, °C/mV

\( K_i \) Bed height at minimum fluidizing conditions, ft

\( M_i \) Mass inside the ith reactor, lbs

\( M_i(s) \) Laplace transform of mass inside the ith reactor

\( N(s) \) Laplace transform of disturbance signal

\( P(t) \) Controller output in a continuous form

\( P(z) \) Z transform of controller output

\( p_{id} \) High pressure input to the ith control valve, psi

\( p_{iU} \) Low pressure input to the ith control valve, psi

\( P_{i,n} \) The ith controller output in a discrete form

\( P^s \) Laplace transform of the ith controller output

\( p_n \) Controller output in a discrete form

\( q_c \) Total energy, generated in or transferred to, in the combustion reactor

\( R \) Correlation coefficient of linear regression

\( t \) Temperature response in the temperature control system, °C

\( T \) Sampling time, sec

\( t_d \) Derivative time, sec

\( t_i \) Reset time, sec

\( T_{IV} \) The ith control valve first-order time constant, sec

\( T_p \) Temperature in the pyrolysis reactor

\( u(t) \) Step function

\( w \) Weight of solids, lb

\( x \) Displacement of the control valve, cm

\( Y(s) \) Laplace transform of process controlled variable
\( Y(t) \) Process controlled variable in a continuous form.

\( Y(z) \) Z transform of process controlled variable

\( Y(t) \) Process controlled variable measurement in a continuous form.

\( Y_m \) Process controlled variable measurement in a discrete form

\( Y(t) \) Setpoint in a continuous form

\( Y_n \) Process controlled variable in a discrete form

\( Y(z) \) Z transform of setpoint

\( \Delta F^\text{c} \) Total flow rate change at \( t \), lbs/hr

\( \Delta P \) Pressure drop across the fluidized bed

\( \Delta P_{\text{ig}} \) Gap between high pressure input and low pressure output

\( \Delta P_{\text{c}} \) Total controller signal change at \( t \rightarrow \infty \), psi

\( \Delta Y \) Temperature change, C

Greek:

\( e_c \) Void fraction in the combustion reactor

\( e_{mf} \) Void fraction in a bed, at minimum fluidization

\( e \) Void fraction in the pyrolysis reactor

\( p \) Fluid density, lb/ft\(^3\)

\( p \) Density of solids, lb/ft

\( \theta \) Dead time of temperature control system, min

\( X \) First-order time constant in the temperature control system, min

\( T \) Retention time of solids inside the combustion reactor, min

\( T \) Retention time of solids inside the pyrolysis reactor, mm
CHAPTER I

INTRODUCTION

Over 90% of the recoverable fossil reserves in the United States are in solid forms, while about 75% of the energy consumption is geared for liquid and gaseous fuels. The gross imbalance between the physical form of the energy resources and the physical form upon which the consumption is based has resulted in the development of new technologies for the production of liquid fuels from the solid forms of the fossil reserves. But before October 1973, the abundance of imported low-cost petroleum liquids and gases had slowed down the progress. In October 1973, the OPEC declared "oil embargo" to the world and caused the energy crisis. Since that time, research has been accelerated to develop new processes for extracting oil from oil shale and tar sands and to upgrade coal. Even though the concept of OPEC has been widely thought to be a short-lived, politically motivated cartel, the utilization of these new resources could provide a long-range, independent method to solve the imbalance between energy demand and energy supply.

Compared to the crude oil reserves in the United States of 31.3 billion barrels (Auldridge, 1976), the tar...
sand deposits of Utah, recently estimated at 25 billion barrels of bitumen in place (Ritzma, 1976), represent a significant fossil reserve. Tar sands, commonly referred to as oil-impregnated rock deposits, is naturally occurring sandstone combined with bitumen, which is usually a highly viscous tar. Before bitumen is of any commercial value, it must be separated from the sand particles which it surrounds. Methods for recovering the organic matter from tar sands may be classified into two general categories: in-situ recovery techniques and methods that require the mining of the oil-impregnated rock. Factors that influence the selection of an appropriate recovery method are thickness of the overburden zone, thickness and continuity of the bitumen impregnated zone, and bitumen saturation (Wang, 1983). A detailed discussion of the methods by which the bitumen may be recovered is given by Wang (1983).

At the University of Utah, current work involves a two-stage fluidized-bed thermal processing unit for tar sands. In the first reactor, the bitumen on the sand is pyrolyzed into crude oil and gaseous product at 450-550°C. The coke by-product of the first reaction flows into the second reactor and combusts with air at 550-650°C to supply the energy required in the pyrolysis reaction. The main problem to be overcome in this type of process is the transfer of energy from the combustion reactor to the
pyrolysis reactor. Different methods have been developed to deal with the energy-transfer problem. In 1949, Gishler (1949) developed a process that used the large amount of the cycled, clean, hot sand to transfer the energy. Weeks (1977) used the hot gaseous product of combustion in the second reactor to transfer the energy into the pyrolysis reactor. This method is similar to the preheating of the fluidizing gas. Jayakar (1979) developed a new process, which used a potassium heat pipe as the medium to transfer the energy. Bezama (1983) analyzed these three different methods thermodynamically. He concluded that the use-of-heat pipes for transferring energy from the combustion reactor to the pyrolysis reactor has a clear thermodynamic advantage over the other two methods.

The reaction of the bitumen pyrolysis is endothermic and can be written as follows:

\[
\text{bitumen} \rightarrow \text{coke} + \text{oil} + \text{gaseous product}. \quad (1-1)
\]

The oil yield is strongly affected by the temperature in the pyrolysis reactor and the solids retention time in the pyrolysis reactor. The temperature in the pyrolysis reactor, when operating at a constant feed rate, is assumed to be a function of the energy transferred by the heat pipes. The solid retention time, when operating at
constant feed rate, neglecting the reacted bitumen, is. assumed to be a function of the mass inside the pyrolysis reactor. Prior to the work presented here, the solid retention time was controlled by a conventional proportional-integral (PI) controller and the temperature in the pyrolysis reactor was controlled manually. Due to the hysteresis behavior of the control valves, the amount of the mass inside the pyrolysis reactor could not be controlled well but oscillated strongly. Manual control of the temperature in the pyrolysis reactor was so difficult that the overshooting problem always existed. The work presented in this research involves the application of a digital computer to process control. Included are the dynamic response and control of the temperature in the pyrolysis reactor, the elimination of the hysteresis behavior in the control valves, the attainment of more stable control of mass, and use of a data acquisition system needed to better study the behavior of the thermally coupled two-stage fluidized-bed reactor.
CHAPTER II

OBJECTIVES

The research program discussed here was undertaken with the following objectives:

1. Review digital control theory, study different control strategies for the two-stage fluidized-bed level control, and prove that the temperature in the pyrolysis reactor is controllable by adjusting the amount of energy transferred to the combustion reactor.

2. Install and construct the hardware system for automatic data logging, data processing, and process control. The hardware system includes an HP-2250 interface, pressure transmitters, pressure transducers, thermocouples, an SCR power controller, and several different kinds of solenoid valves and relays.

3. Construct the network system for the emergency alarm to notify the operator of certain major problems in the system.

4. Develop the software systems, DCTPF, TPDSL and EMER for automatic data acquisition, filtering, control output, and emergency detection. The program listings are presented in the appendices.
The first objective is mainly discussed in Chapter III. Chapter V presents the results and conclusions and a further discussion of objective 1. Part of objective 2 is addressed in Chapter III, according to its function. The rest of objective 2 and objectives 3 and 4 are presented in Chapter IV. The last chapter is devoted to the recommendation concerning future work on the control of the thermal processing system for tar sands.
CHAPTER III

CONTROL THEORY AND STRATEGY

This chapter presents a review of control theory, including the conventional control loop and the digital control loop, and the strategy of the control of two fluidized-bed levels and the pyrolysis temperature. Much of the material was taken from Harrison (1978), Harriott (1964), Smith (1972), and Rinard (1982). In the first section, the concepts of "conventional control, the types of control systems, the application of the digital computer to process control, the mathematics of the sampled-data system, and the different control algorithms used in this research are discussed. The second section develops the model of the bed level control system for both reactors, tests four different control algorithms by model simulation and concludes that the proportional-integral-derivative (PID) control algorithm plus valve linearization is the preferred method to be used to control the experimental unit. The last section is devoted to the development of a model for a pyrolysis temperature control system using shortcut techniques, a discussion of factors that disturb the temperature in the
pyrolysis reactor, and the selection of the PID control algorithm with large proportional gain to control the temperature in the pyrolysis reactor.

**Review of Control Theory**

The fundamental objective in process control is to keep certain key process variables as close to the setpoints as possible during as much of the time as possible. Usually, the setpoints are constant; that is, they do not change with time. The variables, other than the controlled variables, can be classified into two categories. The first group is manipulated variables, which can be changed by the controller to maintain outputs at their desired values. The second group includes all the other process variables, called disturbance or load variables, which affect the output.

Two types of control systems, servo and regulatory, have been distinguished by control engineers. If the setpoints are subject to frequent changes (rarely the case) such that the system seldom, if ever, reaches steady state, the controller is said to be a servo. Its primary purpose of control is to force the process outputs to track their setpoints. If the setpoints are rarely changed, the controller is said to be a regulator. Its control purpose is to maintain the process output within the error limits in the face of disturbances.
If a step change is made in the manipulated input to a regulatory system and the process seeks a new level such as in pyrolysis temperature control, then the system of this type is said to be self-regulating. But if a step change is made in the manipulated input to a regulatory system and the process output will not seek a new level such as in the fluidized-bed level control, but will increase or decrease indefinitely, then this type of system is said to be non-self-regulating.

For control problems, it is also helpful to use a block diagram to show the functional relationships among the parts of the system. Each part is represented by a block that contains the mathematical relationship between one input and one output. The basic control loop in conventional systems is the simple feedback loop illustrated in Fig. 1. The letter G represents the mathematical function that describes the behavior of the system element. The process has a single output variable, Y(t), which is to be controlled. That variable must be measured, the value being Y(t). This value is compared to the desired value or setpoint to generate the error signal, E(t), and the control system acts on the error signal via the control algorithm, which is a proportional-integral-derivative (PID) relationship or some simplification, to produce the controller output, P(t). This controller output is imposed upon the process by an
Fig. 1. Basic conventional feedback-control loop,
actuator, which is an automatic positioning valve in many cases. The purpose of this control system is to force \( Y(t) \) to track the setpoints, \( Y(t) \), with some reasonable fidelity, even though the process is subject to one or more disturbances, called \( N(t) \).

Conventional control systems basically suffer from inflexibility. There must be almost a one-to-one correspondence between control loop functions and hardwares to perform these functions. Subsequent modifications of the control strategy require modifications of the analog hardware. To overcome this problem, the digital computer can be used. The function of the analog hardware controllers is to do some mathematical operations. They can be replaced by a digital computer such that any control strategy is programmable, and most modifications in the strategy are simply program changes.

Figure 2 illustrates the basic control loop encountered in digital control systems in which the measurement is discrete. This is the basic difference between digital control systems and conventional control systems. The basic function of the input multiplexer is to obtain values of a function at certain intervals, \( T \), an operation known as sampling. The input multiplexer is an analog to digital (A/D) converter. The output of the sensor is a continuous function of time, \( Y(t) \). The output of the input multiplexer is a number sequence, \( Y \). The computer
Fig. 2. Digital-control loop..
compares the output of the input multiplexer with a desired value or setpoint retrieved from some storage location within the computer itself. Using the appropriate control algorithm, the computer calculates a value for the manipulated variable from the error sequence, \( E \). The output of the control algorithm, \( P \), is sampled and transferred by the output multiplexer, called a D/A converter, to the input of the hold element, whose function is to construct (from the number sequence \( P \)) a continuous signal \( P(t) \) that becomes the input to the process. The form usually used is a zero-order hold, which maintains the last value of the output during the following sampling interval, i.e.,

\[
P(t) = p_n, \quad nT < t < (n+1)T.
\] (3-1)

The relationship between input and output in a block is expressed mathematically via Laplace transforms for the conventional control system. However, for discrete systems, a different transform, called the z-transform, is used to express mathematically the relationship (Smith, 1972). A block is given as an example:

\[
\begin{array}{c|c|c}
E_n & s & G_n \\
--- & --- & --- \\
E(z) & G(z) & P(z) \\
P_n & & \\
\end{array}
\]
The discrete system can be represented as

\[ P = Z E.G . = Z E .G . \quad (3-2) \]

Alternatively, this relationship can be expressed in terms of z-transforms:

\[ P(z) = G(z)E(z). \quad (3-3) \]

For a number sequence, the definition of the z-transform is given as follows:

\[
P(z) = \sum_{n=0}^{\infty} P_n z^{-n} = P(0) + P(T)z^{-1} + P(2T)z^{-2} + P(3T)z^{-3} + \ldots \quad (3-4)
\]

The use of z-transforms for discrete signals is quite analogous to the use of Laplace transforms for continuous signals. For systems with dead time, an extension of the z-transform, known as the modified z-transform, is directly applicable to the z-transform of \( G(s)e^{-8S} \) given by the modified z-transform:

\[
c[G(s)e^{-8S}] = C_mG(s)] = G(z/m) , \quad (3-5)
\]

where \( m = 1 - \frac{8}{T} \). In taking the modified z-transform, the dead time is implied to exist, and therefore one
writes $c[G(s)]$ instead of $f[G(s)e^{-}]$. As an example, consider taking the modified, z-transform of $e^{-a}$:

$$
\sum_{n=0}^{\infty} e^{-a(nT-t)}u(nT - 9)z^{-n}
$$

$$
e^{-a(T-t)}z^{-1} + e^{-a(2T-t)}z^{-2} + e^{-a(3T-t)}z^{-3} +
$$

$$
e^{-a(T-t)}z^{-1} + e^{-aT}e^{- mat^2}z^{-2} + e^{-aT}e^{-2aT}z^{-3} +
$$

$$
e^{-aT}z^{-1} + e^{-aT}z^{-1} + e^{-aT}z^{-1} + e^{-aT}z^{-1} + .
$$

A detailed table of z-transforms and modified z-transforms is given by Tou (1959).

Consider the typical block diagram of a sampled-data system as shown in Fig. 3 (Smith, 1972). First, the sampled error $E(z)$ is given by

$$
E(z) = Y_g(z) - Y(z). \quad (3-7)
$$

The sampled output $Y(z)$ is given by

$$
Y(z) = dN(s)G(s) + D(z)\{H(s)G(s)\}E(z)
$$

$$
= NG(z) + HG(z)D(z)E(z). \quad (3-8)
$$
Fig. 3. Typical sampled-data block diagram,
Substituting Equation (3-7) into Equation (3-8), gives

\[
Y(z) = \frac{HG(z)D(z)Y_s(z) \cdot NG(z)}{1 + HG(z)D(z) + 1 + HG(z)D(z)} \quad (3-9)
\]

Four different control algorithms were used in this research. The first one is the Deadbeat algorithm (Kuo, 1963). A deadbeat or minimal response is one that satisfies the criteria that (1) the settling time must be finite, (2) the rise time should be a minimum, and (3) the steady-state error should be zero (Ash, 1981). One specific case that satisfies the above criteria is that the response to a step change in the setpoint should have zero error at all sampling instants after the first. Specifically,

\[
Y(z) = \frac{1}{I - z}\text{ (a step input)}
\]

\[
Y(z) = z^\frac{1}{\sqrt{2}} + z^2
\]

\[
z^{-1}
\]

\[
1 - z^{-1}
\]

Therefore, \( y^{2*} = z^{-1} \). \quad (3-10)

Substituting this into Equation (3-9) and assuming that the disturbance \( N(s) \) is zero gives

\[
D(z) = -\frac{1}{(1 - z^{-1})} \quad (3-11)
\]
The second control algorithm was developed by Dahlin (1968), who suggested that the closed-loop response should behave like a continuous first-order lag with dead time. For a step change in setpoint, the response would be

\[ Y(s) = \frac{-6s}{\delta + 1} \quad (3-12) \]

where \( \delta \) is the closed loop dead time and \( \tau \) the closed loop time constant. In discrete form, the equation is

\[ Y(z) = \frac{-T/X. -N-1}{(1 - 2.\ell) (1 - e^{-\tau z^{-1}})} \quad (3-13) \]

where \( N \) is the whole number of control intervals in 0.

Suppose \( Y(z) \) is a unit step, then

\[ Y(z) = \frac{v}{(1 - e^{-\tau z^{-1}}) z} \quad (3-14) \]

Substituting this into Equation (3-9) and assuming that the disturbance \( N(s) \) is zero gives

\[ \omega_M = \frac{-T/X. -N-1}{1 - e^{-\tau z^{-1}} (1 - e^{-\tau z^{-1}}) z} \quad (3-15) \]

Note that the closed-loop time constant, \( \tau \), appears explicitly in the controller equation and can be used as a tuning parameter.
The third control algorithm is called Kalman's approach (Kalman, 1954). Suppose the system response to a step input would reach the final value in two sampling times and remain at the final value thereafter. The expression for $Y(z)$ is

$$Y(z) = c_1 z^{-1} + z^{-2} + z^{-3} + \cdots \quad (3-16)$$

In order to accomplish this, the manipulated variable will assume two intermediate values and then assume its final value thereafter. The expression for $P(z)$ is

$$P(z) = P_0 + P^x z^{-1} + P^z z^{-2} + P_f z^{-3} + \cdots \quad (3-17)$$

Suppose $Y(z)$ is a unit step, then

$$Y(z) = (1 - z^{-1})(c_1 z^{-1} + z^{-2} + z^{-3} + \cdots)$$

$$= c_1 z^{-1} + (1 - c_1)z^{-2}$$

$$S_1 z^{-x} + S_2 z^{-2} = S(z), \quad (3-18)$$

$$Y(z) = (1 - z^{-1})(P_0 + P_x z^{-1} + P_f z^{-2} + P_f z^{-3} + \cdots)$$

$$= P_0 + (P_x - P_0) z^{-1} + (P_f - P^z) z^{-2}$$

$$= q_0 + q_x z^{-1} + q_2 z^{-2} = Q(z). \quad (3-19)$$
Recall that HG(z) is the ratio of Y(z) to P(z):

\[ \text{HG}(z) = \frac{Y(z)}{P(z)} \]

The following relationships among the coefficients should be noted:

\[ S_1 + S_2 = x \]  \hspace{1cm} (3-21)

\[ q_0 + q + q_2 = 2r^- \]  \hspace{1cm} (3-22)

These relationships do not generally hold, but simply dividing by the sum of the numerator coefficients will insure that both hold. Substituting Equation (3-20) into Equation (3-9) and assuming that the disturbance N(s) is zero gives

\[ D(z) = \frac{Y(z)/Y_s(z)}{\text{HG}(z) \times 1 - Y(z)/Y_s(z)} \]

\[ = \frac{Q(z)S(z) - Q(z)}{S(z)1 - S(z)1 - S(z)} \]

The last algorithm used here is the most common algorithm in use today; the discrete equivalent of the proportional-plus-integral-plus-derivative (PID) controller or its simplified form, the PI algorithm. Using the trapezoidal rule, the PID algorithm can be expressed as follows:
The retention time of material in the fluidized-bed depends on the level of the bed when operating at constant feed rate. Usually, the level is controlled by a standpipe. If the level of solids in the reactor is higher than the standpipe, then the solids will fall down through the standpipe to keep the level constant. In this case, the reactor is open at all times to the surroundings through the standpipe. This could not be permitted for the reactor units used in this experiment. For the pyrolysis reactor, air would flow through the standpipe into...
the reactor and combust with the bitumen on the tar sands. For the combustion reactor, the level is fixed at some value. If a different retention time needs to be tested, then the unit would need to be dismantled, causing a great deal of mechanical problems.

To maintain an unchanged solids inventory in the fluidized-bed, the feed rate, \( F_n \), is fixed at 4 lbs/hr, which is the design specification for the existing laboratory unit, and the outlet flow rate is automatically controlled by adjusting the displacement of the solids flow control valve, which is controlled by the 3 psig-15 psig pneumatic signal from the pressure transducer or the conventional PI controller.

Response measurement

- Levenspiel (1969) suggested that the inventory of solids in a fluidized-bed can easily be found by pressure drop measurement across the bed. The onset of fluidization occurs when

\[
\Delta P_{\text{t}} = W = (A_t L_{mf}) (1 - \epsilon_{mf}) (p_s - p_g) J L ^{1/n}
\]

By rearranging, we find for minimum fluidizing conditions that,

\[
\frac{\Delta P}{L_{mf}} = \frac{(1 - \epsilon_{mf})(p_s - p_g)}{J L} f_{\text{crit}} \quad (3-27)
\]
At constant gas velocity, greater than the minimum fluidization velocity and less than the terminal velocity of particles, Equation (3-27) becomes

\[
\frac{A P}{r} = (1 - e_f) \left( p_s - p_q \right) g = \text{constant.} \quad (3-28)
\]

Multiplying Equation (3-28) by \(A_{fe}L_f\) and neglecting the \(p'\) (compared with the \(p\)) gives

\[
A P A_{fe} = W = (A_t L_f) (1 - e_f) p_s t^\prime = M^\prime. \quad (3-29)
\]

By rearranging, we find that the relationship between the pressure drop \(A p\) and mass \(M\) is linear:

\[
A P (\text{psi}) = (\text{constant}) M (\text{kg}). \quad \ldots (3-30)
\]

The constants for the reactors in the laboratory are 0.66 for the pyrolysis reactor and 0.76 for the combustion reactor as calibrated by Bezama (1983).

Development of the transfer function \(D(z)\)

The block diagram of the control system for the two-stage fluidized-bed is presented in Fig. 4. Taking the mass balance in the pyrolysis reactor gives

\[
dM, \quad "\text{at}" = F_0 - F_i - \text{reaction terms.} \quad (3-31)
\]
Fig. 4. Levels-control block diagram.
The reaction terms are the amount of the bitumen cracked into oil vapors and gases per unit time. Because of the small amount of the bitumen contained in the tar sands (<8 wt %), the reaction terms can be neglected. By taking the Laplace transform and assuming the initial condition $M_{1}(0) = 0$, Equation (3-31) becomes

$$M_{1}(s) = \frac{F(s)}{V} - \frac{F(s)}{V}.$$  \hfill (3-32)

At fixed flow rate $F_{-}$, we find the following relationship between $M_{..}(s)$ and $F_{..}(s)$:

$$5ITO--I.$$  \hfill (3-33)

For the actuator (control valve) dynamics, a first-order equation is suggested by Murrill (1967):

$$P_{1}(s) = \frac{\frac{1}{T_{lv}s} + 1}{\frac{1}{T_{lv}s} + 1}.$$  \hfill (3-34)

Combining Equation (3-33) with Equation (3-34) gives

$$G_{I}(s) = \frac{M(s)}{\frac{1}{T_{lv}s} + 1} = K \frac{M(s)}{\frac{1}{T_{lv}s} + 1}.$$  \hfill (3-35)

For the combustion reactor, the same relationship between $M_{n}(s)$ and $P_{2}(s)$ can be obtained by assuming that the level
of the pyrolysis reactor is stable, i.e., $F(s)$ is constant:

$$M(s) = -K.$$ 

Recall that the only unknown term in Equations (3-11), (3-15), and (3-23) is the pulse transfer function $HG(z)$. Smith (1972) related the pulse transfer function to the continuous elements of the loop:

$$HG(z) = \{H(s)G(s)\}.$$  \hspace{1cm} (3-37)

The transfer function of the zero-order hold is

$$H(s) = \frac{1}{sT} \exp(-sT).$$  \hspace{1cm} (3-38)

Substituting Equations (3-35), (3-36), and (3-38) into Equation (3-37) gives

$$HG_i(z) = c(H(s)G_i(s)) = dt \frac{-sT}{s(T + 1)} \left( \frac{-K}{s(T + 1)} \right)$$

$$= d \frac{-K}{s^2(T + 1)} + d \frac{K \exp(-sT)}{s^2(T + 1)}$$

$$= d \frac{K^{\pm \nu}}{s^2(T + 1)} + z^{-1}d \frac{-\nu}{s(T + 1)}$$

$$= d \frac{K^{\pm \nu}}{s^2(T + 1)} + z^{-1}d \frac{-\nu}{s(T + 1)}$$
Using the z-transform table given in Tou (1959) to obtain expressions for

\[
5 \to \frac{1}{s (s + \gamma)} \quad \text{gives}
\]

\[
HG_i (z) = \frac{-K_{iv} (1 - z^{-1})}{s (s + \gamma)} T (1 - e^{-i/T}) z \quad (i = 1, 2)
\]

The sampling time T for the mass control is set at 5 seconds as suggested by Smith (L972). The time constants T. of the actuators are 1.92 seconds for the pyrolysis solids flow control valve and 0.464 second for the combustion solids flow control valve; the values are measured from the experimental unit in the laboratory. Substituting the values into Equation (3-39) gives

\[
\begin{align*}
&\frac{3.222}{s} - \frac{1.408}{s^2} + 0.464 z^{-2} \\
&HG_i (z) = -K \left(\frac{1}{s} \frac{1}{(1 - z^{-1})(1 - 0.074 z^{-1})}\right)
\end{align*}
\]
For the Deadbeat algorithm, substituting Equations (3-40) and (3-41) into Equation (3-11) gives

\[
P_{1,n} = -0.074E_{1,n-1} + 1.408E_{1,n} + 3.222K_{1v} + \ldots
\]

\[
P_{2,n} = -2.09E_{2,n-1} + 0.464K_{2v} + \ldots
\]

In difference-equation form, they become

\[
P_{1,n} = (E_{1,n} - 0.074E_{1,n-1} + 1.408E_{1,n} + 3.222K_{1v})
\]

\[
P_{2,n} = -2.09E_{2,n-1} + 0.464K_{2v}
\]

For Dahlin's method, substituting Equations (3-40) and (3-41) into Equation (3-15) and assuming \(N = 0\) (no dead time), \(X = 1.92\) and \(X'' = 0.464\) (the time constants of the actuators) gives

\[
P_{1,n} = \frac{-0.926(1 - 0.074Z^{-1})}{K_{1v}(3.222 + 1.408z)}
\]

\[
P_{2,n} = \frac{-0.999(1 - 2.09E - 5Z^{-1})}{K_{2v}(4.536 + 0.464z^{x})}
\]

In difference-equation form, they become
\[ P_{1,n} = (E_{1,n} - 0.074E_{1,n-1} + 1.521K_{1\nu} P_{1,n-1}) / -3.479K_{1\nu} \cdot \] 

(3-44)

\[ P_{2,n} = (E_{2,n} - 0.09 E_{2,n-1} + 0.464K_{2\nu} P_{2,n-1} + 0.536K_{2\nu} \cdot \] 

(3-45)

For Kalman's approach, normalizing the numerator in \( HG(z) \) gives

\[ HG(z) = \frac{0.696z^{-1} + 0.304z^{-2}}{(0.216 - 0.232z^{-1} + 0.016z^{-2})} \cdot \] 

(3-46)

Substituting the above two equations into Equation (3-23) gives

\[ D(z) = \frac{1}{E_{x}(z)} \left( \begin{array}{c} 0.216 - 0.232z^{-1} + 0.016z^{-2} \\ 0.907z^{-1} + 0.093z^{-2} \end{array} \right) \cdot \] 

(3-47)

In difference equation form, they become

\[ P_{n-1}^{1}(0.216E_{1,n} - 0.232E_{1,n-1} + 0.016E_{1,n-2}) \cdot \] 

(3-48)
For the PID controller, the transfer function $D(s)$ is not a function of $HG(s)$. The unknown terms in Equation (3-26) are $K_i$, $T$, and $T_i$. In difference-equation form, Equation (3-26) becomes

$$P_n - P_{n-1} = K_i \left[ E_n - E_{n-1} + \frac{1}{2T_i} (E_n + E_{n-1}) \right]$$

$$+ \frac{1}{4} (E_n - 2E_{n-1} + E_{n-2})$$

In tuning the PID control algorithm, the integral of the absolute value of the error, (IAE) is used as the integral criteria:

$$IAE = \int_0^\infty |e(t)| dt - \sum_{n=0}^{\infty} |e(nT)| T_n$$

Calibration of the final control element (control valve)

The purpose of this calibration is to find the value of the control valve constant $K_i$. Recall the following relationship between the control output signal and the flow rate through the control valve:

$$AF = K_{iv}$$
The calibration procedure of the solids flow control valves is described below. For the pyrolysis (combustion) reactor:

1. Open fully the combustion (pyrolysis) solids flow control valve.

2. Close fully the pyrolysis (combustion) solids flow control valve.

3. Feed the clean sands into the pyrolysis (combustion) reactor to reach the operating level.

4. Open the pyrolysis solids flow control valve by the pressure transducer to some position, which is controlled by the output of the HP-25510A 4-channel voltage/current function card in the HP-2250. Software, called RDOWR, which was developed by Bezama (1983), is used to command the pressure transducer to open the solids flow control valve.

5. Adjust the feed rate into the pyrolysis (combustion) reactor, until the variation of the level of the fluidized-bed is within some limit. This procedure is to ensure that the flow rate through the control valve is calibrated at a constant level.

6. Obtain the hourly flow rate of solids by collecting the sand for an hour and weighing it.

7. Repeat Step 6 twice and take the average of two flow rates to get the hourly average flow rate.
8. Repeat Steps 4 through 7 for different displacements until the valve is totally open.

9. Decrease the output of the pressure transducer, to some extent, "to close the valve gradually.

10. Repeat Steps 4 through 7 for different displacements until the valve is totally closed.

The results of the calibration for the two control valves are given in Figs. 5 and 6. The K values are

24.264 lbs/hr psi for the pyrolysis reactor and -32.044 lbs/hr psi for the combustion reactor. It can be seen that hysteresis behavior strongly exists in the control valves. The hysteresis will cause slow response and system instability. A method, called gap-make-up, is developed in the following paragraphs to eliminate the hysteresis behavior.

Figures 7 and 8 show the plots of displacement vs. input pressure (to control valve). The data from 1.55 cm to 1.2 cm, for the combustion solids flow control valve, and from 0 cm to 0.4 cm, for the pyrolysis solids flow control valve, are fitted by the following linear equations:

\[ P_{\text{c}} (\text{psi}) = 5.875 x_{\text{cm}} + 8.060 \]  
(3-52)

\[ P_{\text{lu}} (\text{psi}) = 4 - 2.65 x_{\text{cm}} + 6.893 \]  
(open \( \rightarrow \) close) \( R = 0.990 \)
Fig. 5. The calibration result of the pyrolysis solids flow control valve.
Fig. 6. The calibration result of the combustion solids flow control valve.
Fig. 7. Hysteresis behavior of combustion solids flow control valve.
Fig. 8. Hysteresis behavior of pyrolysis solids flow control valve.
\[ P_{2u} \text{ (psi)} = 2.684 \times x_2 \text{ (cm)} + 4.294 \]
(closed \rightarrow open) \quad (R = 0.988), \quad (3-54)

\[ P_{2d} \text{ (psi)} = 3.773 \times x_2 \text{ (cm)} + 4.052 \]
(open \rightarrow close) \quad (R = 0.993). \quad (3-55)

The data, fitted above, are within the operating range.

The flow rate, \( F_{1u} \), through the pyrolysis control valve for the 0.4 cm displacement is approximately 50 lbs/hr, which is well above the nominal feed flow rate of 4 lbs/hr; for the 0.0 cm displacement, it is 1.3 lbs/hr. The flow rate \( F_{2d} \) through the combustion control valve for the 1.55 cm displacement is 0 lb/hr and for the 1.2 cm displacement is approximately 26.5 lbs/hr, which is well above the feed flow rate too.

Combining Equations (3-52) and (3-53) and solving the gap \( \Delta P_{1g} = P_{1d} - P_{1a} \) (psi) at the same displacement \( x \) gives the following equations in terms of \( P_1 \) or \( P_{1u} \):

\[ \Delta P_1 \text{ (psi)} = 0.274P_{1d} \text{ (psi)} - 1.042 \] \quad (3-56)

\[ \Delta P_1 \text{ (psi)} = -378P_{1u} \text{ (psi)} - 1.435 \] \quad (3-57)

Combining Equations (3-54) and (3-55) and solving the gap \( \Delta P_{2d} = P_{2u} - P_{2d} \) (psi) at the same displacement \( x \) gives the following equations in terms of \( P_{2u} \) or \( P_{2d} \):

\[ \Delta P_2 \text{ (psi)} = 0.406P_{2u} \text{ (psi)} - 1.985 \] \quad (3-58)
\[
\text{AP}_2 \ (\text{psi}) = 0.289 \ P_{2d} \ (\text{psi}) - 1.412. \quad (3-59)
\]

After the gaps have been determined, they can be added to or subtracted from the output to the solids flow control valve, depending on which curve, the down or the up, the output is on and the direction of the change of the output. For instance, the output to the pyrolysis solids flow control valve is \(P_e\), and according to the next error signal, the output should be decreased to close the valve. Then after the control output has been determined by the controller, the gap is subtracted to eliminate the hysteresis. The success of using this method is discussed in Chapter V.

---

**The Control Strategy of the Pyrolysis Temperature**

The thermal processing system for tar sands is a thermally coupled two-stage fluidized-bed reactor, which uses a heat pipe to transfer energy from the combustion stage to the pyrolysis stage. All of the energy required in the pyrolysis reactor is transferred directly from the combustion reactor; no additional energy is allowed to heat up the pyrolysis bed. In other words, the temperature in the pyrolysis reactor is controlled by the amount of the energy going into the combustion reactor.

Investigating the energy model developed by Bezama (1983) gives
The variables e, x, e, x, and F are classified as disturbance variables, which are supposed to be constant as long as the levels in the fluidized beds have been controlled. The weight percent of bitumen contained in the tar sands, f, is not a variable but a constant which depends on the kind of tar sand being used. The gas property, g, depends on the kind of gas used to fluidize the pyrolysis bed. The gas may be nitrogen, steam, oil vapor, or gaseous products. The gas property will become invariable, if the condition in the pyrolysis reactor becomes stable. The total energy, Q, generated in or transferred to the combustion reactor is the manipulated variable, which may come from the combustion of the coked sand with air or the energy supplied by the surroundings. As discussed by Bezama (1983), the heat of the combustion for coked sand with air may not be enough to maintain the process. In this case, the amount of the energy supplied by the surroundings is considered to be the manipulated variable, whose value is adjusted to control the temperature in the pyrolysis reactor.

Because of the complexity of the temperature-control system, which contains a heat pipe, the transfer
function for the process, $G$, cannot be obtained by applying either the general material or energy-balance equations to the system. A simple process model is obtained, using a shortcut technique (Smith, 1972), to study the relationship between the system response and the output of the controller. In order to obtain the system response curve, the control loop is opened and the controller is controlled manually. The block diagram for this method is presented in Fig. 9. The following procedures are used to obtain the response curve.

1. Set the output of the controller at some particular value.

2. Wait until the system reaches the steady state.

3. At Step 2, time is assumed to be zero.

4. Add a small amount of change to the output of the controller.

5. Record the system response until it reaches another steady state.

The pyrolysis temperature response curve is presented in Fig. 10. A first-order-lag-plus-dead-time model, $(Ke^{-9s})/(As-1)$, is used to simulate the response curve. The analytic solution of the step response of this model is

$$Y(t) = (AY)(1 - e^{-(t-6)/X}) + Y_0, \ t > e.$$  \hspace{1cm} (3-S1)
Fig. 9. Block diagram for a step input change to the manipulated variable.
Fig. 10. The pyrolysis temperature response to a step input change in manipulated variable.
Calculating $Y(t)$ at two times, "say 9 + (*/3) and 6 + X, gives

$$
Y(S + |-) = 0.284(AY) = 0.284(446 - .412) + 412
= 421.66 \; (^\circ C),
$$

$$
Y(e + \sqrt{}) = 0.632(AY) = 0.632(446 - 412) + 412
= 433.49 \; (^\circ C).
$$

Finding the temperature in the response curve gives

$$
\frac{rc_{421.94}}{515} = \frac{498}{17} = 6 + \frac{f}{r} \tag{3-62}
$$

$$
\frac{rc_{433.49}}{548} - \frac{498}{-48} = 0 + x \; , \tag{3-63}
$$

Solving the two equations simultaneously gives

$$
9 = 1.5 \; \text{(min)},
$$

$$
X = -46.5 \; \text{(min)}.
$$

The $K$ value is the ratio of the change of the response to the change of output of the controller.

$$
K = \frac{*I \; 3i - 17/!c,}{\times T \; AP \; 2400 \; 1200 W^*}
$$

For the temperature range, 400 C to 450 C, the energy flow from the combustion reactor to the pyrolysis reactor is limited by the fluidized-bed heat-transfer coefficients. From the model, it can be seen that the system has a
dead-time and a very large time constant. Compared with the time constant, the dead-time is not so important and can be neglected.

The drawback of this model is that it assumes no disturbance exists in the unit. In the real experiment, the disturbance variables, considered before, always exist during the run. It is impractical to substitute this simple response model into Equations (3-11), (3-15) or (3-23) to obtain the transfer function \( D(z) \). Therefore, the model developed is only used to realize that the system has a very large time constant. The fourth control algorithm discussed before, PiD, needs no process transfer function, \( G(s) \), and is considered the best algorithm to control the temperature in the pyrolysis reactor.

An inferential control method (Rinard, 1983) was also considered. It uses a secondary-process-variable, the temperature in the combustion reactor, as a process variable to be controlled. As discussed before, the energy for the pyrolysis reactor is transferred directly from the combustion reactor through the heat pipe. It is believed that some relationship exists between these two temperatures. If the combustion temperature is controlled around the setpoint, the pyrolysis temperature will be around some temperature, which can be estimated by the relationship. The drawback of this method is the same as that of the simple model, that is, no disturbance can
exist during the run. If the disturbance happens during the run, the relationship between these two temperatures will no longer hold. In this case, the pyrolysis temperature will be out of control, even if the combustion temperature is constant.
CHAPTER IV

EXPERIMENTAL

The purpose of this chapter is to present a description of the experimental apparatus, including sensors, control actuators (including calibrations), interface, host computer, data acquisition and processing system, and emergency alarm system.

The detailed design of the experimental apparatus is given by Weeks (1977) and Jayakar (1979). Therefore, only a short description of the experimental unit is presented here. More emphasis is spent on the discussion of the control actuators, sensors, and network system between the HP-2250 interface and the equipment. Several relays are used to accommodate control either by conventional controllers or by digital control.

Data acquisition and processing are completed through several pieces of software. Two tasks are developed to be run inside the HP-2250 interface using the MCL/50 programming language. These tasks can sample temperatures and pressures in the system at each second, store these data for up to 3 minutes, filter the stored data, calibrate the function cards, check any emergency in the process, and decide on the control output.
Software, used to read the data kept in the interface, store the useful data in the disk of the HP-1000 computer, and display a table describing the process, is scheduled to run every minute. The data kept in the disk is ready for plotting using an HP-7225A plotter by executing software discussed by Bezama (1983).

Because of the sticky property of tar sands, an alarm system was constructed to check the temperature and pressure limit.

**Experimental Apparatus**

As shown in Fig. 11, the experimental unit consists of a two-stage fluidized-bed reactor thermally coupled by a potassium-heat, pipe, a tar sand feeding system, and a product recovery system.

Raw tar sands are fed into the hopper and screw feeder. By the rotating torque of a motor, the screw feeder turns and carries the sand into the fluidized pyrolysis reactor. The falling sand contacts the hot gas going upward and cracks into the oil vapor, gas, and coke. Oil vapor and gas pass through a cyclone and filter, which remove the dust. A cooling system and a small amount of nitrogen purge flow exists in the reducing adapter to prevent the oil vapor and gas from going upward to the feeding system. The oil vapor is condensed and collected through a series of condensers. The exhaust gas, after
Fig. 11. Process flow diagram of tar sands reactor.
passing through the electrostatic precipitator, is sent to the vent without further recovery. The coked sand produced in this reactor is transported to the combustion reactor by gravity through a standpipe, which can be opened by a solids flow control valve. Then, the coked sand reacts with the air in the combustion reactor producing energy which is transferred to the pyrolysis reactor by means of a 3/4-inch heat pipe containing potassium as the working fluid. Spent sand is discharged through another solids flow control valve into a receiver at the bottom of the reactor.

During all the experiments run during this research, the feed rate of tar sands is set at 4 lbs/hr. A higher feed will cause plugging in the reducing adapter, which is on the top of the pyrolysis entrainment section. The flow rate of air for the combustion reactor is 0.68 lb/hr, giving a superficial velocity of 0.38 ft/sec at an operating temperature of 550 C. It is suggested that this amount of air is required for processing 4 lbs/hr of tar sand with 8 wt % bitumen and a coke yield of 18 wt % of bitumen by Bezama (1983). This amount of air provides a superficial flow rate of 35 lbs/hr-ft , which is enough to fluidize the combustion reactor at above ambient temperature.

Several nichrome heating wires are used in the experimental unit. The heating wires used in the recovery
system are not controlled, but set a certain amount of output to maintain high temperatures around 410 C in the entrainment section of the pyrolysis reactor, the cyclone, and the filter. If the oil vapor condenses in those sections, the recovery system will get plugged and the system will have to be shut down and dismantled.

Instrumentation and Control

The following paragraphs introduce the instrumentation, which are classified into five groups according to their functions.

Feeding system control

_The motor that drives the screw feeder to carry the sand into the pyrolysis reactor is controlled by a motor controller-, which performs the primary control functions required to control the speed of, or current supplied to, the DC motor. The calibration of this motor controller is carried out as follows:

1. Put ten pounds of clean sand into the hopper,

2. Set the dial on the speed-control knob at a particular value at the operating condition.

3. Turn on the motor and open fully the control valve.

4. Obtain the hourly flow rate of solids by collecting the sand for an hour and weighing it._
5. Repeat Steps 3 and 4 twice and take the average of two flow rates to get the hourly average flow rate.

6. Set different values on the speed control knob and repeat Steps 3 to 5 to get a different flow rate.

The calibration for the sand feed rate is presented in Fig. 12. When the pressure and temperature are out of limits, the first step is to stop the feed automatically. A triple-pole, single-throw, normally closed relay is installed at the line connecting the power supply and motor controller to execute this function. In case of emergency, the interface will detect the problem, energize this relay to stop the feed and set an alarm to warn the operator. The circuit diagram of the control system of the motor is presented in Fig. 13.

**Temperature control**

As shown in Fig. 14, sixteen K-type, ungrounded, magnesium-insulated, stainless-steel thermocouples, which measure the temperatures at all pertinent locations of the equipment, are sampled by the HP-2250. The sample frequency can be programmed to be as high as 10 Hz for all channels. The most important thermocouple in this system is Tr, the temperature in the pyrolysis reactor. Also, it is the controlled variable for temperature control in this research. Thermocouple T10, representing the temperature in the combustion reactor, is not controlled but kept
Fig. 12. The calibration result of the motor controller.
The circuit diagram of the motor control system.

Fig. 13. The circuit diagram of the motor control system.
Fig. 14. The instrumentation diagram for temperature control.
between 550 C and 700 C to make sure that combustion occurs in this reactor.

A nichrome heating wire is used on the entrainment section of the combustion reactor to make up the heat losses from the heat pipe and thus generate a near-adiabatic zone in the middle part of the heat pipe. A powerstat variable transformer is used to control the amount of the power into this section.

As discussed before, the energy required to maintain the temperature in the pyrolysis reactor is transferred directly from the heat pipe. No heating wire is used in this area. The energy for the combustion comes from two sources. One is the heat of combustion of the air with coked sand. The other uses the heating wire to supply external energy. This external energy supply is used to control the temperature in the pyrolysis reactor. The amount of this external energy going into the combustion reactor is controlled manually by either a powerstat variable transformer or automatically by a single-phase, phase-angle-fired power controller, depending on the status of the double-pole, double-throw relay, which is installed at the line connecting the 120-volt power supply and the heating element, and can be energized by a relay in the HP-2250. After this relay is energized, it still can be de-energized by a master switch on the control.
The circuit diagram of this power control system is given in Fig. 15.

The output of the single-phase, phase-angle-fired power controller is controlled between 0% and 95% of supply voltage in proportion to the 2 V to 10 V DC input signal from the HP-2250. We used a 30-ampere current rating and 120 volts voltage rating. In the experimental unit, two heating elements are wrapped around the combustion reactor. One is on the outer layer to make up the heat losses to the surroundings. It is controlled manually by a powerstat variable transformer. The other, on the inner layer, is used to control the amount of the external energy going into the combustion reactor by the method discussed above.

**Level control**

As discussed in Chapter III, at a constant gas flow rate, the pressure drop across the bed in the fluidized reactor is proportional to the amount of particles in the bed. The mass inside the reactor determines the retention time of the reactor. As shown in Fig. 16, two coupled, fluidized-bed reactors are used. Therefore, four pressure taps are required to measure the pressure signals below and above the bed in each reactor. Those pressure signals are connected to four pressure gauges, two dp cells, and two pressure transmitters, PT1 and PT2, which measure the
The Eighth Channel In HP25514A

Fig. 15. The circuit diagram of the temperature control system.
Fig. 16. The instrumentation diagram for level control.

PTn: Pressure Transmitter
PTRn: Pressure Transducer
maximum pressures in the two reactors, i.e., the pressures at the bottom of the reactors. The output signals of the pressure transmitters, which are measured on the second and sixth channels of the HP-25501A 16-channel high-speed analog input function card in the HP-2250, are between 2 V and 10 V in proportion to the 0 psig to 10 psig input signals. The sampling frequency can be programmed to be as high as 200 Hz for one channel. The calibrations for these two pressure transmitters were made by the manufacturer. The results of the calibration are given in Figs. 17 and 18. By linear regression, the following two equations are obtained for the different reactors:

\[ P_1 = 1.237 \times V_1 \text{ (volt)} - 2.492 \text{ (psig)} \]
\[ (0 < P_1 < 1.225), \quad R = 1.00, \quad (4-1) \]

\[ P_2 = 1.250 \times V_2 \text{ (volt)} - 2.501 \text{ (psig)} \]
\[ (0 < P_2 < 2.155), \quad R = 1.00. \quad (4-2) \]

Two dp cells are used to transfer the pressure signals from the reactors into 3 to 15 psig signals, which are fed to the Foxboro Model 40 proportional-integral pneumatic controllers and two pressure transmitters, PT3 and PT4. Because of bubbles produced in the fluidized bed, the pressure drop across the bed fluctuates strongly. Therefore, an analog filter was installed at the exit of each dp cell to eliminate the high-frequency noise. The output
Fig. 17. The calibration result of the pressure transmitter PT1.
Fig. 18. The calibration result of the pressure transmitter PT2,
signals of two pressure transmitters used here are between 2 V and 4 V in proportion to the 0 psig to 20 psig input signals. Those output signals are measured at the fourth and eighth channels of the HP-25501A function card. The calibrations of these two pressure transmitters are made as follows:

1. Set the setpoints of PI controllers at the desired values.
2. Turn on:
   a) control air
   b) process air and adjust flow rate to the proper value.
3. Feed the clean sand into the reactor.
4. Sample the output signals from the four pressure transmitters simultaneously by the HP-2250.
5. Transfer the voltages obtained from PT1 and PT2, which measure the maximum pressures in the reactors, into the gauge pressures by Equations (4-1) and (4-2).
6. Because the system is open to the vent, the pressure at the top of each reactor equals the atmospheric pressure. In this case, the numerical values of the pressure drops across the beds, equal that of the maximum pressure, which is obtained at Step 5.

The results of the calibration are given in Fig. 19. By linear regression, the following two equations are obtained for the different reactors:
Fig. 19. The calibration result of the pressure transmitters PT3 and PT4
\[ \text{AP}_3 = 0.373 \times V_3 \text{ (volt)} - 1.205 \text{ (psi)} \]

\[ 0.041 < \text{AP}_2 < 1.22', \quad R = 0.999, \quad (4-3) \]

\[ \text{AP}_4 = 0.401 \times V_4 \text{ (volt)} - 1.234 \text{ (psi)}. \]

\[ 0.069 < \text{AP} < 2.028, \quad R = 0.997. \quad (4-4) \]

The pressure drop across the bed is controlled either by a Foxboro Model 40 proportional-integral pneumatic controller or by pressure transducers, depending on the status of the 3-way solenoid valve installed at the line between the solids flow control valve and the controllers. When this valve is energized, the control air is delivered from the pressure transducers, which provide the 3 psig to 15 psig pneumatic signal in proportion to the 2 V to 10 V input signal from the HP-2250. If the solenoid valve is de-energized, the solids flow control valve is controlled by the conventional PI pneumatic controller. There are two ways to energize this solenoid valve. One is to use the switches on the control panel manually and the other is to use the relay in the HP-2250. After the 3-way solenoid valve is energized using a digital relay in the HP-2250, it still can be de-energized by the master switches on the front panel. The detailed diagram of this control system is presented in Fig. 20. The purpose of the master switch is to have a backup system in case of the digital control failure.
Fig. 20. The circuit diagram of the bed-level control system.

From PI Controller

2) The Fifth Or
Sixth Channel
In HP25514A

Power Supply

Switch

\( S_n \)

3-Way Solenoid Valve

3-15 psig Control Valve!

Pressure Transducer

Solids Flow 1, m/min

Regulator

Air Supply

25 psig

80 psig

0–10 V

From The First
Or Second Channel
In HP25510A
The calibration of the pressure transducers were made by the manufacturer. By linear regression, the following two equations were obtained:

\[ P_1 = 1.335 \times V \text{ (volt)} + 1.31 \text{ (psig)} \]
\[ (0 < V < 10), \quad R = 0.999, \quad (4-5) \]

\[-P = 1.180 \times V \text{ (volt)} + 1.46 \text{ (psig)} \]
\[ (0 < V < 10), \quad R = 0.999. \quad (4-6) \]

Interface

In order to function properly, the host computer needs a bridge to receive data from the process and transmit other data to the process. The bridge or computer/process-interface in this research is the HP-2250 Measurement and Control Processor. Fig. 21 shows such an automation system. Here, a standard interface bus permits the host computer to communicate with the HP-2250. External system sensors translate external events to input signals, while actuators translate output signals to external events.

The HP-2250 is much more than simply an interface to analog and digital signals. The HP-2250 contains the HP-2104 processor unit and the HP-2251 measurement and control unit. The HP-2104 processor unit consists of the following three process cards that supply the bus...
Fig. 21. An automation network system for data acquisition and control.
interface, measurement and control processing, and memory capability:

1. Processor card: It, like the processor in a computer, converts machine code to operations. "An important function of this card is to handle communications between the software and the function cards. At power-on, this processor starts reading from the ROM portion of the memory card. When the power-on sequence is complete, the processor has the information it needs to manage the operation of the HP-2250.

2. Memory card: It holds 16 k two-byte words of read-only memory (ROM) and 16 k words of random-access memory (RAM). The ROM integrated circuits are permanent memory components, which contain power-on instructions, interruption and-status instructions, and the MCL compiler. The RAM storage area is temporary memory, which is lost when the power is off unless there is a battery backup card in the HP-2250. The existing unit in the laboratory of the Chemical Engineering Department of the University of Utah has no battery backup card. The RAM storage area belongs to the tasks programmed in the HP-2250. The processor reserves about 2 k words for storing its own management data. The RAM command is used to read the amount of memory remaining for compiling tasks and storing the resulting data.
3. Measurement and control interface card: It contains the HP-IB interface card and the measurement and control interface (MCI) card. The HP-IB interface card is connected to an HP-IB bus cable and carries control signals and data to and from the host computer. The MCI card, which links the 2250 processor to the BIF cards located below the first function card slot in the 2251 unit, contains a time-of-day clock and internal times for various measurement pacing modes.

The HP-2251 measurement and control unit has card frames designed to hold the backplane interface (BIF) card and up to eight I/O (function) cards.

1. Backplane interface card: The function of this card is to interface the function card backplane to the measurement and control interface (MCI) card.

2. Function cards: Up to eight function cards can be installed in one HP-2251 measurement and control unit. The function cards match the type of sensor or actuator in the external process. Each card occupies one slot in an HP-2251 unit. The existing HP-2250 unit has only the following four cards:

   a) HP-25501A 16-channel high-speed analog input: It provides the basic analog-to-digital conversion capability for the HP-2250 measurement and control process system. Its combination of high speed (50 kHz) analog-to-digital converter, sample and hold amplifier, and
programmable gain amplifier (gains of 1, 2, 4, .8, or autorange) provides the ability to process a wide variety of analog input signals. It has 16 different input channels with a full-scale range of 1.25, 2.5, '5 and 10 volts, positive or negative. This card is used in this research to read the output signals of the four pressure transmitters.

b) HP-25503A 32-channel low-level multiplexer: "The LLMUX adds low-level voltage input expansion capability to the HP-25501A 16-channel high-speed analog input card. Low-level inputs, down to 1.56 microvolts, can be accepted at a channel-to-channel scan rate of 20 kHz.' This card is primarily used in interfacing to microvolt and millivolt inputs, such as the input from thermocouple devices. The HF-25594A thermocouple reference connector (TRC) is used to provide a reference voltage and inputs up to 15 various thermocouple devices. This card provides the reading capability of the sixteen K-type thermocouples in this research.

c) HP-25510A 4-channel voltage/current analog output: It provides digital-to-analog output capability for the HP-2250, with 12-bit resolution and voltage or current output. Each channel is isolated one from the other and from earth ground. The output provides either a unipolar/bipolar voltage source or a current source,
depending on the positions of the unipolar/bipolar and voltage/current switches on the card. The voltage range is 10 volts, positive or negative; the current range is 20 milliamperes. The maximum analog output rate is 31 kHz. The purpose of this card is to send 0 to 10 volt signals to the two pressure transducers, which control the solids flow valve, and the single-phase, phase-angle-fired power controller, which controls the temperature in the pyrolysis reactor in this experiment.

d) HP-25514A 16-point relay output: This function card has 16 points of single-pole, double-throw (form c) relay contacts, which are individually isolated. The relays on the card are controlled separately and the card supplies relay coil power. The 16 channels can be programmed as independent points or as a 16-point field. The relays are rated for switching open-circuit voltage of up to 250 VAC at 1.5 A, 125 VAC at 3.0 A, and 30 VDC at 2.0 A. Operating times are less than 15 milliseconds. The existing unit in the laboratory has 125 VAC, 2.0 A fuses installed to prevent overload damage to this function card.

The HP-2251 unit is a real-time programmable interface, which contains programs in ROM that allow it to accept streams of commands and data from the host computer, compile the commands, and then execute the commands
in the real-time environment of the external system. The commands received from the host computer are written in MCL/50 language. In the Hewlett-Packard programmer manual, a full description of the MCL/50 language can be found. Table 1 shows the commands used to develop the software for data acquisition and reduction, decision-making, and execution of control algorithms in this research.

The tasks written in MCL/50, which is scheduled to run in the HP-2250 interface, are loaded into the interface by the host computer in the input/output statement, like the "write" statement in Fortran. After storing the tasks in the RAM, the interface is free from the host computer and can execute the streams of commands according to their priority. A software run in the host computer may be needed to read the data kept in the interface to prevent memory overflows in the RAM of the interface. The main defect of the HP-2250 is that only integer operations can be done. It limits the capability of engineering calculations and causes significant round-off errors. The defect can be overcome by running the arithmetic operations in the host computer to decide the output of the function cards, then sending the tasks to the HP-2250 for execution. In this case, the host computer will be tied up and slow down all other programs currently running.
TABLE 1

MCL/50 COMMANDS USED IN THIS RESEARCH

<table>
<thead>
<tr>
<th>Command</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>AAVERAGE</td>
<td>Average analog input data</td>
</tr>
<tr>
<td>AI</td>
<td>Read analog input</td>
</tr>
<tr>
<td>CLB</td>
<td>Calibrate analog input function cards</td>
</tr>
<tr>
<td>CTIMER</td>
<td>Clear time elapsed</td>
</tr>
<tr>
<td>DIMENSION</td>
<td>Dimension buffers and variables</td>
</tr>
<tr>
<td>DO</td>
<td>Write sequential digital outputs</td>
</tr>
<tr>
<td>GOSUB</td>
<td>Run a task as a subroutine</td>
</tr>
<tr>
<td>GOTO</td>
<td>Unconditional transfer of program control</td>
</tr>
<tr>
<td>IN</td>
<td>Store input data</td>
</tr>
<tr>
<td>KTEMP</td>
<td>K-type thermocouple readings</td>
</tr>
<tr>
<td>LABEL</td>
<td>Program control label</td>
</tr>
<tr>
<td>NTASKS</td>
<td>Maximum number of tasks</td>
</tr>
<tr>
<td>OUT &quot;-'</td>
<td>Output stored data</td>
</tr>
<tr>
<td>PAUSE</td>
<td>Suspend background task</td>
</tr>
<tr>
<td>PTIMER</td>
<td>Pause for task timer</td>
</tr>
<tr>
<td>RANGE</td>
<td>Set analog range</td>
</tr>
<tr>
<td>RCLOCK</td>
<td>Read clock</td>
</tr>
<tr>
<td>REF</td>
<td>Read thermocouple reference</td>
</tr>
<tr>
<td>REPEAT</td>
<td>Repeat . . . n times</td>
</tr>
<tr>
<td>RESET</td>
<td>Reset firmware and memory</td>
</tr>
<tr>
<td>REWIND</td>
<td>Rewind buffer</td>
</tr>
<tr>
<td>SCLOCK</td>
<td>Set clock</td>
</tr>
<tr>
<td>SKIP</td>
<td>Skip words in buffer</td>
</tr>
<tr>
<td>START</td>
<td>•Start background task</td>
</tr>
<tr>
<td>STOP</td>
<td>Stop resident task</td>
</tr>
<tr>
<td>TASK</td>
<td>Define resident task</td>
</tr>
<tr>
<td>VO</td>
<td>Output single word integer voltage</td>
</tr>
<tr>
<td>WNOW</td>
<td>Wait now .</td>
</tr>
<tr>
<td>WTIMER</td>
<td>Wait until time elapsed</td>
</tr>
</tbody>
</table>
Host computer

The host computer used as controller for the HP-2250 interface unit is a 16-bit HP-1000 minicomputer, which is a real-time, multi-user, multiprogramming, time-sharing system that is interrupt driven. Under the HP-RTE-IVB operating system, the computer can execute instructions at the rate of one million per second. For floating-point processing, the speed can go up to 200,000 "operations per second. Scientific and engineering computations, including floating-point arithmetic, matrix operations, polynomial evaluation, and trigonometric and logarithmic operations, make the HP-1000 computer suitable for manipulation of large data arrays. Together with the HP-2250 interface, the system is especially suited for measurement, control, and instrument automation. System resources are allocated according to job priorities. The tasks with highest priority get immediate attention; lower priority jobs run as computer time becomes available. As discussed above, if the host computer determines the control action for the control system, the tasks will have the highest priority to execute immediately and occupy all the computer time.

The HP-1000 computer manages the HP-7906 MR hard disk system, HP-2645A system console, HP-7225A plotter, HP-12792A RS-232C 8-channel multiplexer and HP-2250
interface to construct a network system for data acquisition, storage, display and printout.

Data Acquisition and Processing—
(Software Development)

Acquisition and processing of data by digital techniques can be divided into four areas, which generate different types of difficulties. The first area is the acquisition of on-line data, which is generated at the rate of 0.3 megabyte per hour. The second area is noise rejection, i.e., data filtering and compression. The third area is the method of finding and storing useful data, including data transfer between two real-time software packages. The last area is the processing of final data. The four areas are fully discussed in the following paragraphs:

Acquisition of on-line data

An MCL/50 task, which is scheduled to run at each second, samples the on-line data, including 16 thermocouples, 4 pressure transmitters, 2 real-time digital inputs, 1 analog input, and the real-time clock. The data sampled from the same channel are stored in one buffer in the HP-2250 for up to 3 minutes. In other words, the size of one buffer is declared to be 180 elements. The reasons for storing the data for 3 minutes in the interface are to prevent the RAM from overflowing and to have a long
history of data available for the host computer. The sampling time of these 24 channels takes about 120 milliseconds. The rest of the time in one second is used to run the control task. The two real-time digital inputs provide the real-time branching capabilities for the control system. This branching capability allows four real-time options. Two are used to display the information obtained in two different formats, one decides when the host computer stores the data into disks, and the last option stops the program at the end of the experiment. Program DCTPF initializes the HP-2250 and loads two tasks into the interface; one is the data sampling task discussed above and the other, which is the control task, will be discussed later.

Data filtering and compression

Signals emanating from the sensor systems have noise superimposed upon the useful information from which meaningful control action can be determined. It is necessary to incorporate noise-rejection schemes into the control loops. A two-stage analog/digital filtering system is used in the pressure measurement. For the temperature measurement, which involves less noise, only digital filtering is used.

The source of noise for the temperature measurement mainly comes from stray electrical pickup. Four
signal conditioning modules have been installed in the HP-25503 multiplexer for low-pass RC (resistance-capacitance) filtering. The source of noise for the pressure measurement comes from uncontrollable disturbances, which are generated by bubbles in the fluidized bed, and the stray electrical pickup. The analog filter installed for pressure measurement is a small tank, the retention time of which is around 0.1 second, which is high enough to eliminate the high-frequency noise generated by the bubbles and low enough to prevent any delay in the pressure measurement.

Suppose that readings, say \(X_i\), are taken every \(T\) units of time. From these readings, two algorithms are used as the digital "filter to calculate the filtered or smoothed values \(Y\) (Smith, 1972). The first algorithm is simply the arithmetic average:

\[
Y = \frac{1}{N} \sum_{i=0}^{N-1} Z \quad (4-7)
\]

Since \(N\) is the number of readings for \(Y\) required to calculate one value of \(Y\), the control action is taken only at intervals equal to \(NT = T\) units of time. This algorithm is easily programmed in the HP-2250 because there is an MCL/50 command "AAVERAGE" to average the analog input.
The second algorithm is the numerical equivalent.
of the first-order lag. For continuous signals, the
differential equation describing the first-order lag is

\[ \frac{dY(t)}{dt} + aY(t) = Y_m(t). \]

Expressing the equation by finite differences,

\[ x(nT^n) + y_{m,n} = y_{n-1,m,n} \]

or \[ y_n = a y_{m,n} + (1 - a) y_{n-1} < \]

where \( a \) is between 0 and 1. The smaller the value \( a \) is,
the more strongly the first-order filter will behave.
If \( a \) is too small, the measurement of the sensors is
delayed and causes the system to be controlled with more
difficulty. The implementation, of this algorithm has
problems in calculating the \( a \) value in the HP-2250, which
has no floating-point arithmetic operations and has
integer operations in the range only between -32768 and
32767. This problem is solved by the following procedure:

1. Divide the values \( Y_{n-1,m,n} \) and \( Y_{n,m,n} \) by a small
   factor, CI, which is between 1 and 50.
2. Multiply the first-order constant \( a \) by 100.
3. Substitute the \( Y_{n-1,m,n} \), \( Y_{n,m,n} \) and \( a \) into Equation
   \[ (4-8) \]
   and obtain the value \( Y_n \).
4. Divide the value $Y'$ by a factor $C_2$, which is equal to $100/C_1$.

The readings of the measurements for pressures, which are between 2000 mV to 8000 mV, and temperatures, which are between 300 mV to 7000 mV, are so large that round-off errors in the above procedure are negligible. The filter performance on the readings of the measurement "for the pressure drop across the combustion fluidized bed is presented in Fig. 22. The filtered data has been separated by 0.5 psi in order to see the difference. It can be easily seen that the first-order filter can attain the better result. In this experiment, all the data obtained have been filtered by the first-order filter using a constant of -0.1.

**Method of finding and storing useful data**

Because of the RAM size in the interface, the data kept in the interface needs to be read out by the host computer and updated. There are two ways to transfer the data from the HP-2250 to the HP-1000. The first is to use the "RELEASE" command in MCL/50 language; the other is to use the secondary address "6" using "EXEC" calls. Before the discussion of these methods, it is necessary to introduce the buffers in the HP-2250, which store the data sampled by the function cards.
Fig. 22. Comparison of the different filtered data and raw data.
Before using the buffers in the HP-2250, one uses a "DIMENSION" command to declare the number and size of MCL buffers available to the tasks in the system. The buffer space is limited only by the amount of memory available after allowing space for all tasks and subroutines to be stored, for the largest main task, and for the mainresult buffer. In MCL, buffers have specific names, starting with B. The buffer names include a numeric suffix in addition to the letter, like B4. To use buffer elements in arithmetic and logical operations, one must add an indexing parameter to the buffer name in order to specify the buffer element. Fig. 23 shows how the index number references buffers with respect to pointers. The buffer pointer points to the last word read into the buffer by the "IN" command or to the last word written out of the buffer by the "OUT" command. Commands "DIMENSION," "REWIND," and "RELEASE" reset the pointer to zero; otherwise, the pointer rests at the last word input by means of the "IN" command or output by means of the "OUT" command. Buffer access through the read and write secondary addresses, ports, and arithmetic and logical operations is not controlled by the pointer and does not affect the position of the pointer. Wherever the pointer is, that location is the zero reference for the buffer index parameter; in other words, Bk(0) is always the location of the
Fig. 23. Buffer pointers and index numbers in HP-2250.

Before Moving A Word In Or Out:

<table>
<thead>
<tr>
<th>Pointer</th>
<th>Index Numbers</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Bb(1)</td>
</tr>
<tr>
<td>Word 1</td>
<td>Bb(1)</td>
</tr>
<tr>
<td>Word 2</td>
<td>Bb(2)</td>
</tr>
<tr>
<td>Word 3</td>
<td>Bb(3)</td>
</tr>
<tr>
<td>Word k</td>
<td>Bb(^)</td>
</tr>
<tr>
<td>Word 5</td>
<td>Bb(5)</td>
</tr>
</tbody>
</table>

After Dim Or Rewind:

<table>
<thead>
<tr>
<th>Pointer</th>
<th>Index Numbers</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Bb(1)</td>
</tr>
<tr>
<td>Word 1</td>
<td>Bb(-3)</td>
</tr>
<tr>
<td>Word 2</td>
<td>Bb(-2)</td>
</tr>
<tr>
<td>Word 3</td>
<td>Bb(-1)</td>
</tr>
<tr>
<td>Word ^</td>
<td>Bb(0)</td>
</tr>
<tr>
<td>Word 5</td>
<td>Bb(1)</td>
</tr>
</tbody>
</table>

After Moving Word 4 In Or Out:

<table>
<thead>
<tr>
<th>Pointer</th>
<th>Index Numbers</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Bb(1)</td>
</tr>
<tr>
<td>Word 1</td>
<td>Bb(-3)</td>
</tr>
<tr>
<td>Word 2</td>
<td>Bb(-2)</td>
</tr>
<tr>
<td>Word 3</td>
<td>Bb(-1)</td>
</tr>
<tr>
<td>Word ^</td>
<td>Bb(0)</td>
</tr>
<tr>
<td>Word 5</td>
<td>Bb(1)</td>
</tr>
</tbody>
</table>

Pointer Increments One Count Before Moving A Word In Or Out.
The "RELEASE" command is the first and simpler way to transfer the stored data in the buffers to the computer. This command makes buffer data from MCL buffers available at a port. After a task executes the "RELEASE" command, the computer can obtain the data by reading from the port, addresses. Totally, there are four ports, A, B, C, and D, which can be accessed. On execution of the "RELEASE" command, the HP-2250 seals the buffer against execution of an "IN" command designating the same buffer, or a "DIMENSION," or an "OUT" command. Any of these commands will suspend execution of the task containing the command until the transfer of data is complete. When the "RELEASE" command finishes, the buffer is rewound. As discussed before, "the HP-1000 is a multi-user, multi-programming system. It executes programs according to job priorities. If the stored data is transferred from the HP-2250 to the HP-1000 by the "RELEASE" command, the program used to read out the stored data in the buffers must have the highest priority to be executed in order to guarantee that the data transferring process can be finished and does not seal the buffer against storing the incoming data. In this case, the HP-1000 will be tied to the experiment all the time. This is not the situation allowed in this computer system.
The second and more complicated way is to use the secondary address "6" using "EXEC" calls. To read from the secondary address 6, one must first write two words to 6, then read the number of data items specified in the second word. The difficulty with this data transferring process is generated from finding the pointer in the buffers. Totally, there are 24 buffers, each buffer containing 180 elements, to be transferred each time. The data in each buffer are updated after each second. It takes 120 milliseconds to finish sampling 24 channels. In order to have same pointer positions in all the buffers, the data transferring process should not proceed during the sampling time. In other words, the data transferring process has only 830 milliseconds to be completed. A flag is designed to send a message, indicating that the sampling task was just finished, to the program being run in the HP-1000. After receiving this message, the program will start transfer of the process data, which takes only 650 milliseconds.

The data transferred to the HP-1000 by the method discussed above has the same pointer position in every buffer. The time buffer is used to find the pointer position by finding the element that contains the largest integer. After the pointer position is found, the useful data are stored from the previous largest integer.
(previous time) to the largest integer (the time when the HP-1000 starts reading the buffers in HP-2250).

There is no need for this method to assign the highest priority to the program and tie up the HP-1000 with the experimental run all of the time. If the program, which reads out data from the HP-2250, has been delayed by another higher-priority program, the HP-2250 still is able to sample the data, store the data in the buffers, filter the data, and determine the control output of the reactor. The only loss is the data sampled during the delayed time. From the above discussion, it is concluded that the second method is better than the first one. In this study, the second method was used for the data transfer.

Processing of final data

After the data have been filtered, they are compressed and stored in the HP-7906 MR hard disk system. A real-time table shown in Fig. 24, which describes the process, is scheduled to be displayed on a CRT located at the laboratory each minute. The operator can display the same table, by running the program TARPC (listed in Appendix C), at different CRTs as long as they are connected to the HP-1000.

Management of the stored data in the disk is done by a series of programs developed by Bezama (1983), which
allow the display of the data in different resources. Fig. 25 shows schematically the processing of the data and the software used to manipulate the data off-line. The program SELEC is used to check a specific data file, i.e., FILEnn, locate the interesting data, plot the information at the CRT, and record the information into a different file (DATAF0), which is ready to be processed by the program XDATA.

The program XDATA acts on the file DATAF0 and generates the parameters required for the proper operation of the HP-7225A plotter. These parameters and the information to be plotted are then stored in a new file named DATAF1. The program PLOTY takes all the information from the file DATAF1 and sends the required information to the plotter. Finally, the program TITLE is used to add alphanumeric information to the plot such as title, subtitle, axis, and comments.

Emergency Alarm System

Because of the stickiness of some tar sand feeds, the lines and the valves in the system are very prone to plugging. Since the thermal-processing system for tar-sands is a long-time running system, it is necessary to have an alarming system to warn the operator to take notice of the emergency. Two alarming systems have been developed for the experimental unit. One, designed for
Fig. 25. Off-line data management flow chart,
the operators in the laboratory, includes the alarm, the indicating lights, and the pressure gauges on the control panel. The alarm will notify the operator to watch the control panel, which will show where the emergency occurs. The second is designed for operators at other locations, as long as there is an alarm and a CRT connected to the HP-1000. After the alarm has warned the operators, they can check the emergency by running the program EMER (listed in Appendix B). This program has the secondary address to the HP-2250 buffers, which designates the emergency flag. It can detect the status of the emergency flag, then show the emergency on the screen of the CRT accessed by the operators. The operator then can decide what action to take.

The following emergency conditions will send the alarm signal to the operator:

1. The temperatures in the system are out of limit, 700°C for combustion and 600°C for the pyrolysis. This might occur if too much energy is added to the system.

2. The temperature in the filter or the cyclone is below 410°C. If so, the oil vapor will condense and cause the lines in the recovery system to plug. This problem is caused by the breakage of the heating wires used in the recovery system.
3. The pressure at the exit of the pyrolysis reactor is out of limit. This happens when the exit or recovery system gets plugged.

4. The pressure at the exit of the combustion reactor is out of limit. It is least likely to happen, but may occur, if the lines of the exit are too small.

5. The pressure drop across the bed in either the combustion reactor or the pyrolysis reactor is above the controlled value. Plugging of the solids flow control valve will cause this problem. It is most likely to happen when the solids flow control valve for the pyrolysis reactor is plugged.

6. The exit of the screw feeder is plugged. This may occur when the product gases enter the water-cooled neck of the screw feeder and condense on the walls and on the feed material. When the system is running with tar sands and one opens the lid of the hopper, then the pressure in the hopper will be smaller than that in the recovery system. In this case, the product gases will enter the water-cooled neck of the screw feeder.
CHAPTER V

RESULTS AND DISCUSSION

As discussed in chapter III, the temperature in the pyrolysis reactor is strongly influenced by the amount of tar sands in the reactor. On the other hand, the mass of solids in each reactor is independent of the temperature in the reactor, if the amount of the reactants (bitumen in the pyrolysis reactor or coke in the combustion reactor) is negligible, compared to the clean sand. For this reason, mass control in the fluidized-bed reactor was studied first. After the fluidized-bed level has been controlled, i.e. the mass in the reactor is approximately steady, we can assume no disturbance on the control of temperature in the pyrolysis reactor to simplify the control strategy. The following presents the result of fluidized-bed level control in the first section and the result of temperature control in the second section. The best control algorithm and parameters which were found in the first section were used, when temperature control was studied.

Fluidize'd-Bed Level Control

A model developed by Bezama (1983) was used to test different control algorithms, including the Deadbeat
algorithm, Dahlin algorithm, Kalman's approach,, and the •
PID algorithm. The accuracy-of the model simulation was
discussed and proved by Bezama. From the results of the
different control algorithms, the PID algorithm was chosen
to control the fluidized-bed level in the experimental
unit. First, -the results of the model simulation are
presented and discussed with reasons for using the PID
algorithm to control the experimental unit. The results
and discussion of PID control in the experimental unit are
then presented.

System response in mass control
with model simulation

As discussed in Chapter III, hysteresis behavior
strongly exists in the solids flow control valve. To see
the influence of' hysteresis of the control valve on
control of the fluidized-bed level, Figs. 26 and 27 are
presented. The same controller constants were used except
that hysteresis was eliminated by a Gap-Make-Up method
developed in chapter III. The results in Fig. 27 have a
smaller-amplitude and faster-frequency response than the
results in Fig. 26. This is explained by the fact that
the linearization of the valve has increased the speed of
valve response. Therefore, the linearization of the
control valve is added to the different control algorithms
studied here to eliminate the hysteresis.
SIMULATION OF BEDLEVEL CONTROL
PRESSURE DROP VS TIME

<table>
<thead>
<tr>
<th>Time (MIN)</th>
<th>0.00</th>
<th>0.30</th>
<th>0.60</th>
<th>0.90</th>
<th>1.20</th>
<th>1.50</th>
<th>1.80</th>
<th>2.10</th>
<th>2.40</th>
<th>2.70</th>
<th>3.00</th>
</tr>
</thead>
<tbody>
<tr>
<td>COMBUSTION REACTOR</td>
<td>SETPOINT - 1.55</td>
<td>KC - 30, TI - 10</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>PYROLYSIS REACTOR</td>
<td>SETPOINT - 0.7</td>
<td>KC - 10, TI - 100</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Fig. 26. Simulation of pressure drop responses using the digital PID control algorithm (valve with hysteresis).
Simulation of Bed Level Control
Pressure Drop vs Time

Fig. 27. Simulation of pressure drop responses using the digital PID control algorithm plus valve linearization.
Figures 28 to 30 show the system response to a step change in mass control simulation, when using three different digital control algorithms. In deriving the controller $D(z)$, one assumes that the response should have zero error at the second sampling time for the Deadbeat algorithm and the third sampling time for Kalman's algorithm. From the response, it is apparent that the zero error is impossible to obtain at the second or third sampling time for a sampling time of 5 seconds. To overcome this problem, the controller $D(z)$ is used only when the response approaches the setpoint so closely that at the next sampling time or the sampling time after the next, the error can be assumed approximately to be zero. Before this occurs, the solids flow control valves are closed or opened, depending on whether the error signals are positive or negative. Due to the modeling errors, the response does not exactly equal the setpoint of the sampling instants. The major modeling error came from the calibration of the control valve. The control valve constants, $K$, used in the controller $D(z)$ are assumed to be constant for the whole range of the displacement; but in Figs. 5 and 6 the calibration curves are not straight lines. The reason for making the assumption that the $K$'s are constant is to simplify the simulation. In order to calculate more accurate $K$ values at each point in the calibration curves, a high-order equation with
SIMULATION OF BED LEVEL CONTROL.
PRESSURE DROP VS TIME

1.80

1.60

1.40

1.20

1.00

0.80

0.60

0.40

0.20

COMBUSTION REACTOR SETPOINT - 1.55

0.00  .30  .60  .90  1.20  1.30  1.80  2.10  2.40  2.70  3.00

TIME <MIN>

PYROLYSIS REACTOR SETPOINT - .7

0.00

TIME <MIN>

0.00 .30 .60 .90 1.20 1.30 1.80 2.10 2.40 2.70 3.00

Fig. 28. Simulation of pressure drop responses using the Deadbeat control algorithm plus valve linearization.
Fig. 29. Simulation of pressure drop responses using the Dahlin control algorithm plus valve linearization.
SIMULATION OF* BED LEVEL CONTROL
PRESSURE DROP VS TIME

1. 80
1. 80
1. 40
1. 20

COMBUSTION REACTOR SETPOINT - 1.53

A H (L)
0.

0.

. 60

. 40.

PYROLYSIS REACTOR SETPOINT - .7

m Q D
. 20.

. 20.

0.00 30 . 60 . 00 1.20 1.50 1.80 2.10 2.40 2.70 3.00

TIME CMIN> •M-U3-M---I--1+

Fig. 30. Simulation of pressure drop responses using the Kalman control algorithm plus valve linearization.
complicated functions must be used for data fitting. But, it is impossible to carry this out to control the experimental unit using the HP-2250 interface. The purpose of studying different control algorithms in mass control with model simulation is to choose the most practical control algorithm to control the experimental unit. Therefore, the complexity of the controller $D(z)$ must be limited to the capacity of HP-2250.

Several instantaneous decreases in pressure drop response of the pyrolysis-reactor were found when the system is controlled either by a Deadbeat algorithm or by a Dahlin algorithm. They can be explained by the fact that the solids flow-rate increases suddenly in the pyrolysis control valve, when the control signal to the valve is higher than 10.5 psi (close-to-open) or 8.5 psi (open-to-close). Referring to Equation (3-31), a big drop happens in $M_{\infty}$, when $F_0$ is constant and $F_1$ suddenly becomes very large.

To tune a digital PI control algorithm, a trial-and-error method was used by finding the minimum value of the integral of the absolute value of the error with a model simulation. Investigating the PI difference form, Equation (3-50), it can be easily realized that the output of the controller is more strongly influenced by the proportional gain than by the reset time. Therefore, a search for an optimum value of proportional gain was made.
first with an arbitrary constant reset time. After the proportional gain was found, the best reset time was found at that optimum value of proportional gain. Table 2 shows some results of tuning the digital PI control algorithm at the following conditions: (1) the setpoint and initial pressure drop in the combustion reactor are 1.55 and 1.5 psi; (2) the setpoint and initial pressure drop in the pyrolysis reactor are 0.7 and 0.65 psi; and (3) the value of IAE is obtained by a trapezoid rule from $t = 0$, to $t = 1000$ seconds. Program MCIPI, listed in Appendix A, is the software used. The result of the digital PI control algorithm with the best controller constants is presented in "Fig. 31.

Comparing the results from the different control algorithms, it can be seen that the PI control algorithms can achieve a more stable response. Besides, the PI control algorithm has the following advantages over the other control algorithms:

1. It does not need an accurate model to simulate the process. Only an approximate, simple model is needed to tune the algorithm.

2. The process still can be controlled, even if disturbances are occurring. For instance, when the feed rate is not constant, but subject to some small changes, which often happens in the experimental unit, then the
TABLE 2

SOME OF THE TUNING RESULTS FOR THE DIGITAL PI CONTROL ALGORITHM IN MASS CONTROL SIMULATION

<table>
<thead>
<tr>
<th>Pyrolysis Reactor</th>
<th>Combustion Reactor</th>
</tr>
</thead>
<tbody>
<tr>
<td>$i$</td>
<td>$K_c$</td>
</tr>
<tr>
<td>100</td>
<td>90</td>
</tr>
<tr>
<td>100</td>
<td>150</td>
</tr>
<tr>
<td>100</td>
<td>250</td>
</tr>
<tr>
<td>TOO</td>
<td>255</td>
</tr>
<tr>
<td>100</td>
<td>200</td>
</tr>
<tr>
<td>100</td>
<td>300</td>
</tr>
<tr>
<td>150</td>
<td>255</td>
</tr>
<tr>
<td>95</td>
<td>255</td>
</tr>
<tr>
<td>90</td>
<td>255</td>
</tr>
</tbody>
</table>
Fig. 31. Simulation of pressure drop responses using the digital PID control algorithm plus valve linearization with the best controller constants found analytically.
model is not accurate any more. In this case, only the PI control algorithm still has control ability.

3. It is easy to program the HP-2250 interface. The error signals, used in the other control algorithms, are always less than unity and the final control output equations always contain many real numbers. The capability of the HP-2250, which can only do integer operations, will cause a significant round-off error when it tries to calculate the final output value to the actuator. Because of these advantages, the PID control algorithm was chosen to control the experimental unit.

The system response in mass control for the experimental unit

Before going into the tuning of the PID control algorithm, a study of the control valve was completed. The design feed rate of tar sands is around 4 lbs/hr for the experimental unit and the solids flow rates are 0 to 27 lbs/hr for the combustion control valve and 1.3 to 262 lbs/hr for the pyrolysis control valve. Comparing those numbers indicates that the control valves were not well selected. Eckman (1958) suggested that the maximum flow rate which the control valve is to provide should be approximately equal to two times the normal maximum flow in the process. If the control valve is oversized, the valve must operate at low lift and the minimum controllable flow is too large. On the other hand, if the
control valve is undersized, the maximum flow desired for operation of a process may not be provided. To solve the oversize problem of these two control valves, maximum output pressures are set to only partially open the control valves. This is easy to do by programming the controller equations in the HP-2250 interface. Because the relationship between the displacement and the control pressure in the valve does not hold constant, which may happen when solid particles go inside the control valve and cause friction with the sliding stenr, the limits on the output pressure are programmed in a way that they can be changed in real time. This means that the limits still can be changed, even if the control program is running in the HP-2250 to control the process. To obtain better control responses, the maximum flow rates through the control valves should be limited to the values that are close to twice the feed rate. However, too small openings of the control valves will cause plugging of the valves by particles fed into the system. It was found by experiment that a maximum output pressure to the pyrolysis control valve of 8.8 psig and a minimum output pressure to the combustion control valve of 7.9 psig seemed, to give better control ability without plugging of the control valve at room temperature. Because of the different thermal stress in the control valve at higher temperature, the maximum or minimum output pressure to the control valve was adjusted.
to 9.85 and 7.8 psig when the temperature in the combustion reactor is higher than 450 C and the temperature in the pyrolysis reactor is higher than 300 C.

Due to strong fluctuations in the pressure drop across the bed at higher temperatures, which is caused by the existence of bubbles inside the reactor and may result in a meaningless control action, the fluidized-bed level control was studied at room temperature to obtain a better response. Figures 32 to 35 present the pressure-drop response to a step change in setpoints when using the PID control algorithm to control the experimental unit at different controller constants for the pyrolysis reactor. The cyclic response of pressure drop in Fig. 32 results from the slow action of the control valves. Due to small proportional gains, the control valves opened slowly when the bed level was increasing and closed slowly when the bed level was decreasing, thus never having a chance to reach steady state. When increasing the proportional gain for the pyrolysis control valve gradually from 10 to 80, it can be seen that a faster frequency, smaller amplitude response was obtained. The result presented in Fig. 35 for the pyrolysis reactor can be considered as the best control. It does not exhibit overshooting and unstability problems. The reason that the response from the pyrolysis reactor can not become a straight line is because of fluctuations existing in the system. Besides, the system
Fig. 32. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization, File 23.
Fig. 33, Experimental pressure drop responses using the digital PID control algorithm plus valve linearization. File 52.
Fig. 34. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization, File 53.
EXPERIMENTAL DATA FROM FILE T FILE31
PRESSURE DROP VS TIME

Fig. 35. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization with the best controller constants found experimentally.
is subject to a disturbance coming from the feed rate $F_n$, which is not truly constant because of the feeding system.

To see the effect of the derivative time in the PIE) control algorithm, Figs. 36 and 37 are presented. The same proportional gain and reset time for the pyrolysis reactor are used except that the derivative time of the controller is 1 in Fig. 37 and 0 in Fig. 36. It can be seen that adding the derivative time to the controller can speed-up the action of the control valve and obtain a better result.

Figures 38 to 40 present the pressure drop response to a step change in setpoints at the different control constants for the combustion reactor when the bed level in the pyrolysis reactor is steady. Comparing those results with that in Fig. 35 shows that the controller used in Fig. 35 has the better controller constants but if compared with the controlled result in the pyrolysis reactor, it is not so good. It can be explained as follows:

1. The bed level control system in the combustion reactor is subject to more fluctuations in response than in the pyrolysis reactor.

2. The disturbance coming from the feed rate $F_n$ in the combustion reactor is more serious than that in the pyrolysis reactor.

3. The minimum output pressure of 7.9 psig, which is set to have better control, is still too low. The
Fig. 36. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization. File 51.
Fig. 37. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization, File 54.
Fig. 38. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization, File 55.
Fig. 39. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization, File 25.
Fig. 40. Experimental pressure drop responses using the digital PID control algorithm plus valve linearization, File 56.
solids flow rate corresponding to that input, pressure in
the control valve is 15 lbs/hr, almost four times the feed
rate of tar sands. Increasing the minimum output pressure
to the control valve will cause the opening of the valve
to be too small and cause plugging by the particles.

In Fig. 41, two step changes in the setpoints have
been made to the bed-level control system when using the
digital PID control algorithm. The response to these two
step changes is so fast that the system becomes stable in
just a few minutes. In order to see the improvement of
fluidized-bed level control by a digital PID control
algorithm, the results of pneumatic control of levels
using conventional PI controllers with a proportional band
width of 3% and an integration constant of 0.5 minute,
which are suggested to be the controller constants by
Bezama (1983), in both reactors is presented in Fig. 42.
Comparison between Fig. 41 and Fig. 42 indicates that
digital control can achieve much better responses.

Control of Temperature in
the Pyrolysis Reactor

The control of temperature in the pyrolysis
reactor was studied with clean sand first to find adequate
controller constants. Then, these constants were applied
to control the unit when feeding the tar sands into the
pyrolysis reactor. The different effects on the unit with
clean sand and with tar sands can be considered as
Fig. 41. Experimental pressure drop responses for two setpoint changes using the digital PID control algorithm plus valve linearization.
Fig. 42. Experimental pressure drop responses using the pneumatic PI controller (valve with hysteresis).
follows: (1) an additional amount of energy is needed to pyrolyze the tar sands; (2) due to oil vapor and gaseous products, produced from pyrolyzing the bitumen, the heat transfer coefficient between the heat pipe and solids in the pyrolysis reactor will change; and (3) an additional amount of energy will be produced in the combustion reactor by the combustion of coke with air. The first and second effects can be considered together as the load changes to the temperature control system when the temperature is controlled stably with only clean sand inside the reactor. The third effect, which is delayed by a time proportional to the solids retention time in the pyrolysis reactor, is considered as another load change to the system.

Before turning on the electrical power to the combustion reactor, clean sand is fed into both reactors and the steady-state levels are reached by the controller in the first section. Then, after electrical power is applied to the combustion reactor, a time delay generated by the heat pipe is observed in Fig. 43 before the energy can be transferred to the pyrolysis reactor. Therefore, tar sand is fed to the system only if the temperature of the pyrolysis reactor is above 440°C. The time-delay behavior of the heat pipe at low temperatures seems of no importance when studying the control of operating temperature in the pyrolysis reactor. Therefore, only a PID
Fig. 43, Reactor temperature history.
control algorithm without any dead-time compensation techniques is used here. Figures 44 to 48 show the responses of $T_1$, which is inside the pyrolysis reactor but above the fluidized bed, $T_2$, which is inside the fluidized bed of the pyrolysis reactor, and $T_0$, which is inside the fluidized bed of combustion reactor, to a step change in setpoint of controlled variable, $TV$. The oscillating response of $TV$ in Fig. 46, when using the small proportional gain, can be explained by the big process gain and large time constant existing in the system. From the results, it seems that increasing the proportional gain to 250 can result in a small amplitude, but still oscillating response. When the proportional gain reaches 1000, the response of $TV$ is very, stable and almost equals the setpoint, even though the total energy entering the combustion reactor is not constant, which is evidenced by the small oscillation behavior response of $T_n$ in the combustion reactor. This phenomenon indicates that a small amount of energy change in the combustion reactor will not have the influence on the temperature in the pyrolysis reactor. This fact is more strongly proved by the results presented in Fig. 49, in which a small proportional gain is used, but the controlled variable is $T_n$, instead of $T'$. The amplitude of oscillation in the response of $T_0$ is around 15 C, but the response of $T_1$ seems to have reached steady state. Comparison between Figs. 50 and 51,
Fig. 44. Temperature responses to a setpoint change of $T_3$ using the digital PID control algorithm, File 34
**EXPERIMENTAL DATA FROM FILE T FILE45 J**

**TEMPERATURE VS TIME**

**Fig. 45.** Temperature responses to a setpoint change of $T$ using the digital PID control algorithm, File 4
Fig. 46. Temperature responses to a setpoint change of $T_5$ using the digital PID control algorithm, File 43.
EXPRESSMENTAL DATA FROM FILE: FILE 4B
TEMPERATURE VS TIME

SETPOINT: T3 > -450
KC = 500, TI = 7
TD = 0

Fig. 47. Temperature responses to a setpoint change of T using the digital PID control algorithm, File 4
Fig. 48. Temperature responses to a setpoint change of $T_r$ using the digital PID control algorithm with the best controller constants found experimentally.
Experimental data from PTEE: TEMPERATURE vs TIME

Fig. 49. Temperature responses to a setpoint change of $T_{10}$ using the digital PID control algorithm.
Fig. 50. Temperature responses to a setpoint change of T using the digital PID control algorithm. File 4
Fig. 5.1. Temperature responses to a setpoint change of Tr using the digital PID control algorithm, File 50.
in which the same proportional gain and integration time are used, except that in Fig. 50 the derivative time is zero and in Fig. 51 the derivative time is one, indicates that the derivative term in the PID algorithm has the ability to improve the control response.

The lower reading of \( T > \), compared with that of TV, is due to no solids there to increase the heat transfer coefficient between the heat pipe and gas. Figure 52 shows the importance of solids on the heat transfer between the heat pipes and reactor. At a time equal to 160 minutes, the setpoint of \( T_r \) is changed to 480' C. After the controller received this signal, the controller started to increase the power into the combustion reactor. At about 240 minutes, the feeder was disconnected and the bed level in the pyrolysis reactor started to decrease. Due to lack of solids to improve the heat transfer, the temperature in the pyrolysis reactor decreased even though the temperature in the combustion reactor was increasing. At a time of 290 minutes, the feeder was connected again and the solids inside the reactor increased. The temperature in the pyrolysis reactor increased rapidly due to the energy accumulated in the heat pipe.

When the system is running with clean sand, air is used to fluidize the pyrolysis reactor. Air is changed to nitrogen when operating with tar sand. Many problems were encountered during a run with Sunnyside tar sands. Two of
Fig. 52. Temperature responses to a setpoint change of T using the digital PID control algorithm, File 44.
the most serious problems were bridging of sticky feed material in the hopper and plugging of the 1/4-inch solids port which carried sand from the pyrolysis bed to the combustion bed. The first problem is due to the nature of the bitumen feed when the particles are sized to -24 to +150 mesh, and can be solved by feeding -16 to +24 mesh tax sand particles into the hopper. The second problem is due to the small size of the discharge solids port and can be eliminated by using larger sizes of the pyrolysis reactor standpipe and solids flow control valve.

Figure 53 presents the results of temperature control when running with Sunnyside tar sands. Tar sand was fed into the hopper at 470 minutes. Due to the time delay in the feeding system, the pyrolysis reactor started to pyrolyze tar sands at 480 minutes. As expected, the pyrolysis of tar sands is considered as a load (disturbance) change to the control system and temperature inside the pyrolysis reactor decreases. A sudden drop of temperature can be explained as follows: (1) an additional amount of energy is needed for the endothermic pyrolysis reaction; (2) a smaller heat transfer coefficient results due to the presence of gaseous products; and (3) an expansion of the bed is caused by the generation of gaseous products with a consequent rapid cooling of the fluidized bed when it was in contact with cooler sections of the reactor (Bezama, 1983). It can be seen in Fig. 53 that
Fig. 53. Temperature responses of two reactors using the digital PID control algorithm when running with tar sands, File 44.
the temperature of the pyrolysis reactor reached steady state very quickly again after being subject to a load change.

In order to observe the controllability of temperature in the pyrolysis reactor under different disturbances, Figs. 54 and 55 are presented. In Fig. 54, Sunnyside tar sand was fed to the system at 140 minutes. At 142 minutes, a disturbance to the system was made by turning off the fluidized gas to the pyrolysis reactor. Due to a better heat transfer coefficient when the voidage of bed became very small, the temperature increased rapidly, stayed at 490°C for a few minutes, and dropped again to reach steady state. The solids flow control valve for the pyrolysis reactor plugged at about 210 minutes. The compressed nitrogen was used to purge the line to unplug the valve and caused a sudden drop of bed level. Due to a decreased amount of sand inside the pyrolysis reactor, the temperature increased rapidly, but returned quickly to the setpoint. The solids flow control valve for the pyrolysis reactor plugged at 240 minutes and again at 262 minutes. The same procedures were used to unplug the valve and caused the same problem as discussed above.

In Fig. 55, due to poor controller constants, the temperature in the pyrolysis reactor was not very stable, even when the system was running with clean sand only.
Fig. 54. Temperature responses of two reactors using the digital PID control algorithm when running with tar sands, File 58.
Fig. 55'. Temperature responses of two reactors using the digital PID control algorithm when running with tar sands, File 47.
Sunnyside tar sand was fed into the reactor at 220 minutes. When the tar sand started to pyrolyze, the temperature dropped for reasons discussed before. At 292 minutes, the solids flow control valve plugged. The same procedure as that of the previous run was used to unplug the valve and caused the sudden rise in the temperature of the pyrolysis reactor. At 346 minutes, the feeding system plugged and the bed started to decrease. Because of the absence of solids to improve the heat transfer between the heat pipe and inside of the reactor, the temperature in the pyrolysis reactor decreased even though the temperature in the combustion reactor increased at the same time.

From the above results and discussion, it can be pointed out that no matter what kind of disturbance happened to the system, the temperature in the pyrolysis reactor can always be controlled to get close to setpoints. As long as the pyrolysis reactor has solids inside to improve the heat transfer coefficient, the temperature can be controlled to be stable with good control constants.
CONCLUSIONS AND RECOMMENDATIONS

The network system for data acquisition and digital control strategies for fluidized-bed levels and temperature in the pyrolysis reactor, when recovering oil from tar sands by a thermal processing method with a potassium heat pipe for transferring energy from the combustion reactor to the pyrolysis reactor, have been studied experimentally. The results obtained indicate that better controllability can be obtained with digital control. When running the experimental apparatus, one of the most serious problems is plugging of the pyrolysis solids flow control-valve. It can be solved by increasing the standpipe diameter-to-particle diameter ratio. It has been observed that the influence of bitumen pyrolysis on the temperature control can be considered as a step disturbance to the system and it does not change the system dynamic behavior appreciably.

As a result of this study, the following conclusions were reached:

1. The oscillating response of bed levels in both reactors can be minimized by eliminating the hysteresis behavior of the solids flow control valves by using a
digital PID control algorithm plus valve linearization. It is believed that the control of bed level can be greatly improved when using a larger-scale unit, a higher solids feed flow rate, and a suitable size of control valve.

2. The total energy required in the pyrolysis reactor can be transferred by a potassium heat pipe from the combustion reactor. The amount of energy generated in or transferred to the combustion reactor can be considered as the manipulated variable to control the temperature in the pyrolysis-reactor. The effect of the inlet temperature of the pyrolysis reactor fluidizing gas on the control is insignificant because of its small heat content, compared with the heat transferred from the heat pipe. Most of the energy required in the pyrolysis reactor is used to maintain the fluidized solids at some temperature.

3. The solids retention time in the pyrolysis reactor, the feed rate of tar sands, the volumetric flow rate of pyrolysis bed fluidizing gas, the weight percent of bitumen in tar sands, and the amount of energy dissipated to the surroundings are important disturbances to the pyrolysis temperature-control system. Due to the uncontrollable disturbances, the development of an accurate model is difficult. Therefore, a PID algorithm, which does not need an accurate model, is considered to be the
best algorithm, even though the system is subject to every kind of disturbance.

4. The large time constant in the pyrolysis temperature-control system will become small when a pilot-plant scale unit is used and more energy is available in the combustion reactor. Then the system will be more stable and easier to control.

Further research on the thermal processing system for tar sands is suggested as follows:

1. Use cheaper, energy resources, such as coal, to supply the energy in the combustion reactor. The amount of coal in the combustion reactor will be considered as a manipulated variable-to control, the temperature in the pyrolysis reactor if enough air is available for combustion of coal in the fluidized bed. Because the energy is supplied inside the reactor instead of from outside the reactor wall, when using heating wires, the time constant in the control system will be smaller and the temperature response in the pyrolysis reactor will be faster.

2. Try a richer bitumen content in the tar sands to reach a self-sufficient energy supply. In this case, a heat exchanger immersed in the pyrolysis fluidized bed will be needed to take extra energy out of the reactor. Before feeding tar sands into the reactor, the manipulated variable for the pyrolysis temperature control system is the amount of energy supplied by the surroundings. After
feeding tar sands and coke, produced in the pyrolysis reactor, combust with air to supply the energy to the pyrolysis reactor. The amount of coolant in the heat exchanger, or the amount of the energy supplied by the surroundings, or both together can be used as the manipulated variable, to control the temperature in the pyrolysis reactor.
APPENDIX A

MASS CONTROL SIMULATION PROGRAM

Most parts of this program are adopted from program DYNAM, developed by Bezama (1983). The main difference between these two programs is that the program listed here is for digital control and program DYNAM is designed for continuous control.

The control algorithm used in this program is a digital PID control algorithm. For the different control algorithms, the only part that must be changed is the controller equations; the other parts of the program are the same. Therefore, programs DCIDB which uses the Deadbeat algorithm, DCIDA which uses Dahlin algorithm, and DC1KA which uses Kalman's approach, are not listed here.
PROGRAM MCIPI

COMMON/NASS1/CK(2),CI(2),CMAX(2),VM(7,3),VC(6,3),VX(2,2)
COMMON/MASS2/CFMA(2),CFMI(2),SOL(2),SOHC2,SLH(2)
COMMON/BUF1/XBU000),YB(1000),IB(20),IDCB(272),NAME(3),YC<1000>
COMMON/BUF2/CH(2)
DIMENSION CKK(2),CII(2)

INDI=0
RIAEP=0
RIAECP=0

INPUT THE INITIAL, TOTAL, AND SAMPLING TIME

WRITE(1,101)
101 FORMAT(' ENTER INITIAL, TOTAL, AND SAMPLING TIME (SEC)')
READ(1,*)TINI,TTOT,TDEL

INPUT THE SETPOINTS IN BOTH REACTORS

WRITE(1,102)
102 FORMAT(' ENTER TAR SAND FLOW <LB/HR) AND PRESSURE DROP FOR'
', EACH REACTOR (PSI)')
TSFL=TSFL*.4536/3600.0
NITER=IFIX(TTOT/TDEL+0.5)
IF(NITER.GT.1000) NITER=1000

INPUT THE INITIAL BED LEVEL

WRITE(1,106)
106 FORMAT(' ENTER INITIAL PRESSURE DROP IN EACH REACTOR (PSI) ')
READ(1,*)DP1,DP2

INPUT THE CONTROLLER CONSTANTS

WRITE(1,103)
103 FORMAT(' ENTER PROPORTIONAL GAIN AND RESET TIME (S) FOR',
', PYROLYSIS')
READ(1,*)CKK(1),CII(1)
CK(1)=CKK(1)*.66*2685.*.1335E-2
CI(1)=CII(1)

WRITE(1,104)
104 FORMAT(' ENTER PROPORTIONAL GAIN AND RESET TIME <S) FOR',
', COMBUSTION')
READ(1,*)CKK(2),CII(2)
CK(2)=CKK(2)*.76*2495.4*.118E-2
CI(2)=CII(2)
0 059 URITE<1,105>
0 060 105 FOR HAT < ' ENTER LU FOP. RESULTS <+LU=PRINT SOME, -LU=PRINT ALL,' 
0 061 ' 0=ENTER DATA AGAIN')
0 062 READC1,*) LU
0 063 IF(LU.EQ.0) CO TO 10
0 064' IPRN=0
0 065 IFCLU.LT.O) IPRN=1
0 066 IF<LU.LT.O) LU=-LU
0 067 C
0 068 IB(1)=3
0 069: IB<2)=NITER
0 070 IB(3)=NITER
0 071 IB<4)=2
0 072 IB(9)=7
0 073 CALL OPENF<IDCB,IERR,NAME,0,-50,-9)
0 074 CALL URITF<IDCB,IERR,IB,20>
0 075 DO 26 J=1,3
0 076 VM(l,J)=DP1/.66
0 077 VH<2,J)=DP2/.76
0 078 VM(5,J)=XLE1/.66
0 079 VM(6,J)=XLE2/.76
0 080 VH(7,J)=TSFL
0 081 DO 25 1=1,2
0 082 VC(I,J)=VH(I+4,J)-VM(I,J)
0 083 VH(2+I,J)=0.0
0 084 25 CONTINUE
0 085 VC<3,J>=4. 
0 086 VC(5,J)=7.5
0 087 VC(4,J)=13.3
0 088 VC(6,J)=7.5
0 089 26 - - CONTINUE
0 090 TO=TINI
0 091 WRITE<LU,110>TSFL,XLE1,XLE2,CKK,CII
0 092 FORHATC FLOW,PRE.DROP,CKS,CIS=',7F9.4)
0 093 DO 50 11=1,NITER
0 094 JK1=11
0 095 TO=TO+TDEL
0 096 CALL HSDYN(TDEL,IPRN,LU,INDI>
0 097 TOO=TO/60.0
0 098 XB(ll)=TOO
0 099 YBU1)=VM(1,25*0.66
1 100 YCC1>=YMCE2,2)«0.76
1 101 *RITE<LU,14T00,YB(Il),VC(Il),<VH(J,2>,J=3,4),VC(3,2),VC<4,2>
1 102 141 FORHATC 'F5.1,'PRD='2F7.4,' F='2F8.5,' *='' PV''2F5.1)
1 103 RIAEP=RIAFP+<ABS(VC(1,1))+ABS(VC1,2))>«TDEL/2,.#66
1 104 RIAEC=RIAEC+<ABS(VC(2,1))+ABS(VC(2,2)))*TDEL/2 .* .76
1 105 DO 48 J=1,4
1 106 VM(J,1)=VM(J,2>
1 107 VC(J,1)=VC(J,2)
1 108 48 CONTINUE
1 109 VC(5,1)=VC(5,2)
1 110 VC(6,1)=VC(6,2>
1 111 VX(1,1)=VX(1,2)
1 112 VX(2,1)=MX(2,2)
1 113 50 CONTINUE
1 114 NLEN=2*NITER
1 115 CALL URITF<IDCB,IERR,XB,NLEN>
1 116 CALL URITF,IDCB,IERR,YB,NLEN>
1 117 CALL URITF<IDCB,IERR,XB,NLEN)
1 118 CALL URITF<IDCB,IERR,YC,NLEN)

D144
0119 WRITE(1,120)
0120 120 FORMAT('ENTER PLOTTER RANGE FOR Y-AXIS (TWO NUMBERS)')
0121 READU,*)Y&(1),YB<2>
0122 CALL URITF(IDC(I),IERR,XB,4)
0123 CALL WRITF(IDC(I),IERR,YB,4)
0124 90 WRITE(1,121)CKK(1)
0125 • 121 FORHATC '/' CONTROL CONSTANTS ARE (CK,CI,CT)'/ REACTOR 1: '
0126 *3F10.3/' REACTOR 2: ',3F10.3/)
0127 )WRITE(1,123)RIAEP,RIAEC
0128 123 FORMAT('IAE (REACTOR 1) = ',E12.5/
0129 * ' IAE (REACTOR 2) = ',E12.5/) /
0130 URITEO, 127 JKL, NITER
0131 122 FORHATC * OF POINTS='14,10X,' TRUE * OF POINTS='14)
0132 CALL CLOSEUDCB,IERR>
0133 STOP
0134 END
0135 . C
0136 C
0137 C
0138 C
0139 C

SUBROUTINE HSDYN(TDEL,IPRN,LU,INDI)
0140 _C
0141 c *«***K»*«****»*««**««***SK************«****tt«*******XX*tt*****K*
0142 c » *
0143 C * MASS BALANCE DYNAMIC SIMULATOR SUBROUTINE. *
0144 C « *
0145 C >>>*«***«****»#**************»«*«x********»*************
0146 C
0147 C THERMAL RECOVERY OF OIL-FROM TAR SAND SYSTEM.
0148 C
0149 COMMON/MASS1/CK(2),CI<2),CT<2>,CMAXC2),VK(7,3),VC(6,3),VX(2,2)
0150 - COMMON/MASS2/CFHA(2),CFMIC2),SQL(2),SOH(2),SLL(2),SLH<2)
0151 C
0152 C DEFINE INTEGRATION TIMF STEP
0153 C
0154 DTT=TDEL \ 
0155 201 IF(DTT.LE.CT<1)*.05) GO TO 202
0156 " DTT=DTT/2.
0157 GO TO 201
0158 C
0159 C INITIALLIZE DEPENDENT VARIABLES
0160 C
0161 202 DO 210 1=2,3
0162 DO 205 J=1,6
0163 VC(J,I)=VC(J,1)
0164 205 CONTINUE
0165 DO 210 J=1,4
0166 VM(J,I)=VH(J,1)
0167 210 CONTINUE
0168 C
0169 C INITIALLIZE INDEPENDENT VARIABLES
0170 C
0171 DO 215 1=2,1,-1
0172 DO 215 J=5,7
0173 VH(J,I+1)=VH(J,1)
0174 215 CONTINUE
0175 C
0176 C INITIALLIZE INTEGRATION
0177 C
0178 NSTEP=IFIX(TDEL/DTT)
INTEGRATE MASS BALANCE USING TRAPEZOIDAL RULE

DO 242 J=1,NSTEP
   DO 245 L=1,2

SYSTEM DELAY RESPONSE:

AL=0.5*DTT/CT(L)
VC(L+4,3)=VC(L+4,2)*(1.-AL)+VC(L+2,2)+VC(L+2,3)*AL/(1.+AL)
IF<VC(L+4,3).LT.0) VC(L+4,3)=0.
CONTINUE

VALVE RESPONSE EQUATIONS (INCLUDING HYSTERESIS)

FLPH=<.54841*VC(5,3)-5.6458)*CFHA(L)
+"PL= (.74650*VC(5,3)-6.3450)*CFHA(L)
PREH=(VM<3,2)/CFMAd 5+5.6458)/.54841
PREH=(VM<3,2)/CFMAd 5+5.6458)/.54841
PHAL=(VM<3,2)/CFMAd 5+5.6458)/.54841
PHAH=(VM<3,2)/CFMAd 5+5.6458)/.54841

IF<VM(L+2,2) .GE. 7.05E-03) PHAH=PREH
IF<VM(L+2,2) .GE. 7.05E-03) PHAH=PREH

PM(L+2,3)=-PM(L+2,2)
IF<VM(L+4,3).LT.PMAL ,AND. VM(L+4,3) .GE. 8.81) VM(L+2,3)=FLPL
IF<VM(L+4,3).GT.PMAH .AND. VM(L+4,3) .GE. 1D.7) VH(L+2,3)=FLPH
IF<VM(L+4,3).GT.CFMH) VM(L+2,3)=CFMH)
IF<VM(L+4,3).LT.CFMH) VH(L+2,3)=CFMH)
CONTINUE

CONTROLLER ACTION

IF<VM(NR.NF.1) GO TO 251
Y1=VM(1,3)*.66
Y2=VM(2,3)*.76
VRITE(LU,310)NIT,Y1 , Y2 , <VM(JK ,3) , JK=3,4) , <VC(JK,3) , JK = 1,6) , VM<1,2)
VM(1,2)=VMC1,3)
VM(2,2)=VMC2,3)
VM(3,2)=VMC3,3)
VM(4,2)=VMC4,3)
DO 240 J=1,6
VC(J,2)=VC(J,3)
CONTINUE

MFASURED ERROR:

DO 249 L=1,2
VC<(L+4,3)=VM(L+4,3)-VM(L,3)
CONTINUE

CONTROLLER ACTION
```plaintext
AX1 = 0.5 * TDEL > (VCCL, 1) + VC(L, 2) / CI(L) .
VC(L+2, 2) = VC(L+2, 1) + CK(L) * <VC(L, 2) - VC(L, 1) + AX1>
VXCL, 2) = VC(L+2, 2) - VC(L+2, 1)
IF(UNDIS.EQ.0) GO TO 252
CSIN = VX(L, 2) * VX(L, 1)
IF (CSIN.GT.0) GO TO 252
IF(VCCL+2, 2).GT.10.43 .OR. VC(L+2, 2).LT.5.72) GO TO 252
ELIMINATE THF HYSTERESIS OF VALVES
IF(VXCL, 2).GT.0 .AND. L.EQ.1) VCCL+2, 2) = 1.274 * VC(L+2, 2) - 1.042
IF(VX(L, 2).LT.0 .AND. L.EQ.1) VC(L+2, 2) = 0.625 * VC(L+2, 2) + 1.435
IF(VX(L, 2).GT.0 .AND. L.EQ.2) VC(L+2, 2) = 0.7114 * VC(L+2, 2) + 1.412
IF(VXCL, 2).LT.0 .AND. L.EQ.2) VC(L+2, 2) = 1.406 * VC(L+2, 2) - 1.985
GO TO 252
IF(VXCL+2, 2).LT.10.43) VC(L+2, 2) = 20.
CONTINUE
INDI = 1
AFTER INTEGRATION PROCEDURE
DO 294 1 = 5, 7
DO 294 J = 1, 2
VM(I, J) = VM(I, J + 1)
294 CONTINUE
RETURN
END

C BLOCK DATA - KS3L1
COHMON/MASS1/CK(2), CI(b), CT(2), CMAX(2), VKC7, 3), VC(6, 3), VX(2, 2)
COHMON/MASS2/CFMA(2), CFHIC(2), SOL(2), SOH(2), SLL(2), SLH(2)
COHMON/BUF1/XB(1000), YB(1000), IB(20), IDCB(272), NAHF(3), YC(1000)
DATA NAHF,.2HDA, .2HTA, .2HF0/

C CONTROL PARAMETERS:
DATA CK/200., 100. /
DATA CI/30., .30. /
DATA CT/1.92, .464 /
DATA CHAX/2.72, 2.45 /
DATA CFttA/3.96E-2, 3.44E-3 /
DATA CFHI/1.6E-4, 0.0 /
DATA SOL/- .67668, 10.588 /
DATA SOH/- .5776, 8.835 /
DATA SLL/0.096362, -1.2626 /
DATA SLH/0.070087, - .90X3 /

C VECTOR INITIALIZATION
DATA VM/21#.0.0 /
DATA VC/18#.0.0 /
END

END*
```

APPENDIX B

EMERGENCY INDICATOR PROGRAM

Program EMER is used to detect the emergency flag in the process and show the emergency conditions on the CRT terminal.
This program is used to read the emergency in the process and display the results on the screen.

```fortran
* PROGRAM EMFR
DIMENSION IEM(20).
IEM(1) = 28
IEM(2) = 11
CALL SECU(22, 6, IEM, 2)
CALL SECRC22, 6, IEM, U>
DO 55 1 = 1, 11
IF(IEM(1).EQ.1) GO TO 65
CONTINUE
GO TO 98
WRITE(1, 5)
FORMAT(1X///////////)
IF(IEM(2).EQ.1) WRITE(1,10).
DO 30 I = 1, 11
IF(IEM(I).EQ.1) WRITE(1,20).
IF(IEM(I).EQ.1) WRITE(1,30).
IF(IEM(I).EQ.1) WRITE(1,40).
IF(IEM(I).EQ.1) WRITE(I,40).
IF(IEM(I).EQ.1) WRITE(1,50).
IF(IEM(I).EQ.1) WRITE(1,60).
IF(IEM(I).EQ.1) WRITE(I,60).
IF(IEM(I).EQ.1) WRITE(1,70).
IF(IEM(I).EQ.1) WRITE(I,70).
IF(IEM(I).EQ.1) WRITE(1,80).
IF(IEM(I).EQ.1) WRITE(I,80).
IF(IEM(I).EQ.1) WRITE(1,90).
IF(IEM(I).EQ.1) WRITE(I,90).
IF(IEM(I).EQ.1) WRITE(1,100).
IF(IEM(I).EQ.1) WRITE(I,100).
IF(IEM(I).EQ.1) WRITE(1,110).
IF(IEM(I).EQ.1) WRITE(I,110).
IF(IEM(I).EQ.1) WRITE(1,120).
IF(IEM(I).EQ.1) WRITE(I,120).
STOP
END
```

APPENDIX C

TABLE DISPLAY PROGRAM

Program TARPC can access the buffers in the HP-2250, which stores the information of the process and displays a table describing the process.
PROGRAM TARPC, HSU-2Q MAR. 83

* THIS PROGRAM IS USED TO REAP DATA FROM HP-2250, AND DISPLAY A TABLE DESCRIBING THE PROCESS.

* PROGRAM TAP.PC

DIMENSION IV(10), I1<195, 4>, IT(780), I2(39, 25), D<8>, THP(23)

A DIMENSION LFl(5), LF2(25), TIM(iy5), IND(19S, 2), TIKN(59), IBU(20>

DIMENSION II1ME(5)

COMMON ITX(23, 12), TMP1(23), IBAD(20)

SET KNOWN DATA

DATA D/.12497E-2, -.25094E+1, .40074E-3, -.12342E+1, .12368E-2, 
,.24922E+1, .37244£-3, -.12046E+1/

DO 24 1 = 1, 20
24 IBAD(I) = 0

SET KNOWN VARIABLES

J0 = 195

J1 = 780

NL1 = 0

INC = 0

JIII = D

-RLT IS THE LAST SECOND TO TAKE THE DATA FROM HP-2250 EACH TIME.

RLT = 0

FORMAT ENTER FLOW RATE (LB/Hk) '

READO, *) THPK22)

TMP(22) = TMPU22)

yRITE(1, 83)

FORMAT ENTER FIRST ORDER LAG FITER CONSTANTS ' /

' COMB. BED, PYRO. BED, PYRO. TEMP. '/)

READU, *) A, B, C

THE FOLLOWING STATEMENTS ARE USEI> TO READ THE BUFFERS IN HP-2250.

THE DATA CONTAIN PRESSURE AND TEMPERATURE.

CHECK IF THE HP-2250 HAS FINISHED SAMPLING DATA ?

IDET = 0

IV(1) = 7

IV(2) = 1

CALL SECW(27, 8, IV, 2)

CALL SECW(22, 8, IV, 1>

IF(IVO) .NT.IDET) GO TO 40

IDET = 1

IF<IV(1).NE.IDET) GO TO 40
THE HP-2250 HAS FINISHED SAMPLING DATA.
REAP DATA FROM HP-2250.

DO 10 I=1,25
   IF (ICQ.EQ.1).OR.(CI.EQ.10)) GO TO 10
   IF (ICI.EQ.18).OR.(I.EQ.20) GO TO 10
   IF (ICI.EQ.22) GO TO 11
   IV(1)=I
   TVC2(1)=I

DO 18 I=1,2
   IVC1(I)=I*5
   IVC2(I)=J0
   CALL SECU(22,6,IV,2)
   CALL SECRC22,6,I1(1,I),J0>

CONTINUE

DO 18 I=1,2
   IVC1(I)=1*5
   IVC2(I)=J0
   CALL SECU(22,6,IV,2)
   CALL SECRC22,6,I1(1,I),J0>

CONTINUE

DO 50 N=1,J0
   RNT=3.6E+3*ITC4*N-3)+ITC4*N-2)+1.E-3*IT(4*N-1)+1.E-6*IT(4*N)
   TIM(N)=RNT/60.
   IF (ABS(RNT-RLT).LT.0.1) IS=N
   IF (RPT.GT.RNT) IM=N-1
   P.PT=RNT

CONTINUE

IFLAG=1
IF (IS.EQ.0) IFLAG=0
IF (IM.EQ.0) IM=J0
IM=IH+2
IF (IM.GT.J0) IM=IM-J0

IF (IM.GT.J0) IM=IM-J0

IM=IH+2
JX=J0
DO 80 M=1,2
LX=1
IF (LX.LT.IM) LX=IM-IS+1
IF (LX.GT.IM) LX=IH-IS+1+JX
L?=LX
K=IS-1
DO 70 N=1,LX
K=K+1
IF(K.GT.JX) K = L -
IND(N,M)=K
CONTINUE
IF(M.GT.L) GO TO 80
JX=J2
L1=LX
IH=IFIX(IM/5.)
IS=IFIX(IS/5.)
TFDS.LE.O) IS=J2
IF((IS-IM).EQ.1 .OR. (IS-IM),EQ.(J2-1)) IS=IM+2
IF(CIS-IM).EQ.O) IS=IM+2
IF(IS.GT.J2) JS=IS-J2
CONTINUE
C FILTER THE BIG NOISE OUT
C
HIT=100
DO 17 N=1,20
IMN.GT.16) MIT=150
L11=L1-1
DO 15 I=1,L11
IF(ABS(I1(IND(I,1),N)-I1(IND(I+1,1),N)).GT.MIT) I1(IND(I+1,1),N)=
I1(IND(I,1),N)
CONTINUE
L22=L2-1
DO 45 I=1,L22
IF(ABS(I2(IND(I,2),N)=I2(IND(I+1,2),N)).GT.MIT) I2(IND(I+1,2),N)=
* I2(IND(I,2),N)
CONTINUE
CONTINUE
C 'FILTER THE DATA BY FIRST ORDER" LAG FILTER
C
DO 44 N=1,4
IF(IFLAG.EQ.0) GO TO 42
I1(INDU,1),N)=LF1(N)
DO 42 N.LE.2) CC=C
DO 42 N.LE.2) CC=A
DO 42 N.LE.2) CC=3
DO 45 M=2,L1
J=IND(M,1)
JM1=IND(H-1,1)
I1(J,N)=IFIX(CC*I1(J,N)+(1-CC)*I1(JM1,N))
CONTINUE
LF1(N)=IMIND(L1,1),N)
DO 44 N=1,23
IF(N.EQ.21) GO TO 48
IF(IFLAG.EQ.0) GO TO 46
I2(IND(1,2),N)=LF2(N)
IF(N.LE.16) CC=C
(IF(N.EQ.17) CC=A
IF(N.EQ.18) CC=B
DO 49 M=2,L2
J=IND(H,2)
C REARRANGE THE DATA BUFFER TO BE DISPLAYED
0185 C
0187 DO 38 H=1,L2
0188 J=IND(<H,2)>*5
0189 I2(IND(<M,2),5)=I1(J,1)
0190 I2(IND(H,2),10)> = IKJ,2>
0191 I2(IND<H,2),18)=I1<J,3>
0192 I2CIND<H,3>,20)= IKJ,4>
0193 TIIHCIND<M,2>)=TIM(J)
0194 38 CONTINUE
0195 66 CONTINUE
0196 C
0197* C STORE THE LAST DATA INTO THE BUFFERS TO BE DISPLAYED BY A TABLE
0198 C
0199 DO 31 1=1,16
0200 THP1(I+1)=i2<IND<L2,2),I>/10.
0201 31 CONTINUE
0202 K=0
0203 DO 41 1=1,7,2
0204 K=K+1
0205 TWP1(K+17)=D<I) ii2<IND<L2,2),K+16)+D(I+1)
0206 IIXCK+17,INC>=1MIX<rHP1CK+17>*100>
0207 41 CONTINUE
0208 TMPK23)=TIMN<IND<L?),2n
0209 C
0210 C CALL THE TABLE DISPLAYING PROGRAM
0211 C
0212 CALL DISP(1,0)
0213 END
0214 END*
SUBROUTINE DISPCLX, JDIG

* SCHEMATIC DISPLAY OF DATA *

COHHON ITXC23,12>, THPC23), IBADC20>
IHRS=IFIXC"HPC23>/60. >
IHIN=IFIXCTHPC23>-IHKS/60
NLI=IFIX(ThP (1>)
IF<JDIG.EQ.1) GO TO 160

DISPLAY SCHEMATICALLY DATA

IF<JDIG.EQ.1) GO TO 160

KEEP TRACK OF DATA DISPLAYED

GO TO 199

DISPLAY TABLE

GO TO 199
FORHATS

FCRMATC1')

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Program DCTPF will load two MCL/50 tasks into HP-2250 interface. After loading the tasks, DCTPF is scheduled to run every 15 seconds to check the status of the tasks running in the interface.
&DCTP1 T=0U004 IS ON CK01012 USING 00037 BLKS R=0000

0001  FTPN4,L
0002  C
0003  C PROGRAM DCTPF,HSU-29 DEC. 82
0004  C
0005  C **"***""***"***""***""***"**""***""***""***"**""***""***"**""***""***"**""***"**"**"***"**""***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**" ***
0006  C
0007  C « THIS PROGRAM IS USED TO CONTROL THE TAR SANDS PROCESS *
0008  C * BY THE DIGITAL COMPUTER. IT WILL LOAD TWO TASKS INTO *
0009  C « HP--2250. TASK U ) IS USED TO READ THE TEMPERATURES *
0010  C * AND PRESSURES. TASK(2) IS USED TO CHECK EMERGENCY *
0011  C « AND CONTROL LEVELS AND TEMPERATURES OF REACTORS. *
0012  C « THE FIRST ORDER LAG FILTER IS USED IN THIS PROGRAM. *
0013  C 0 013 C « *
0014  C »**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"**"**"***"***
0015  C
0016  C PROGRAM DCTPF
0017  DIMENSION IV(30),IBU<1000),NAKE1(3),NAHE2(3)
0018  DATA NAME1/2HTP,2HDS,2HL /
0019  DATA NAME2/2HDC,2HTP,2HF /
0020  CALL LGBUF<IBU,1000)
0021  C CLEAR THE HP-IB AND OBTAIN CONTCL OVER THE HP-2250
0022  C
0023  CALL CLEAR<22,1)
0024  C INPUT SET-POINTS FOR THE PROCESS.
0025  C
0026  C
0027  C
0028  C " " HRITE<1,105)
0029  105 FORHAT<- INPUT-DESIZED PRESSURE DROP IN COMB. BED IN P.S.I.')
0030  READU,«) DP1
0031  WRITEU,107)
0032  107 FORMAT< INPUT DESISED PRESSURE DROP IN PYROL. BED IN P.S.I. ')
0033  DP2
0034  WRITE(1,109)
0035  109 FORMATC INPUT MAXIMUM TEMP.. IN COMB. BED IN C.DEG. ')
0036  T10
0037  WRITEU,1U)
0038  111 FORMATS INPUT DESISED TEMP. IN PYROL. BED IN C.DEG.'
0039  T5
0040  C C INPUT FIRST ORDER LAG CONSTANTS ALPHA
0041  C
0042  C
0043  C WRITEU,113)
0044  113 FORMAT<" INPUT FIRST ORDER LAG CONSTS FOR PRESSURES AND TEMP.'/
0045  » ' .1< CCB,CPB,CT <.9 ,'/
0046  » ' FIRST FOR COM. BED,SECOND FOR PYRO. BED,THIRD FOR TEMP.')
0047  V30,V31,V32
0048  C C INPUT CONTROLLER CONSTANTS
0049  C
0050  C
0051  C WRITE(1,115)
0052  115 FORMAT<" INPUT PROPORTIONAL GAIN FOR CONTROLLERS.'/
0053  » ' COMB. BED, PYRO. BED, PYRO. TEMP. '/
0054  * ' ALL THE VALUES MUST BE INTEGERS ')
0055  IKC1,IKC2,IKC3
0056  IFUKC2.GT .0) IKC2=-IKC2
0057  WRITEU,17>
0058  117 FORMAT<" INPUT RESET TIME FOR CONTROLLERS.'/

D158
0059  ' COMB. BED, PYRO. BED, PYRO. TEHP. '
0060  ' ALL THE VALUES MUST BE INTEGERS '
0061  ITR1, ITR2, ITR3
0062  WRITE(1,118)
0063  118 FORMAT(//' INPUT DERIVATIVE TIME FOR CONTROLLER'S.'/
0064  ' COMB. BED, PYRO. BED, PYRO. TEMP. '/
0065  ' ALL THE VALUES MUST BE INTEGERS '
0066  IDT1, IDT2, IDT3
0067  C
0068  C  RESET THE HP-2250
0069  C
0070  WRITE(22,200)
0071  200 FORMAT(//' HP 2250 HAS BEEN RESET. '
0072  CALL CHECK(IFL,2,0)
0073  IF(IFL.EQ.1) GO TO 499
0074  WRITE(1,55)
0075  55 FORMAT(' HP 2250 HAS BEEN RESET. '/)
0076  C
0077  C  DECLARE DIMENSION, NTASKS AND RANGE IN HP-2250
0078  C
0079  WRITE(22,201)
0080  201 FORMAT(' DIMC90, 41, 55, 55, 55, 55, 55, 55, 200, 55, 55, 55, 200, '
0081  ' NTASKS(2), RAN(1,2) 1 0000, RAN<1,4)10000, RAN<1,6)10000, '
0082  ' RAN(1,8), RAN<2,15)100, 100', ' RAN(1,8), RAN<2,15)100, '
0083  ' RAN(2,17)100, '
0084  ' RAN(2,28,3)500, 500, 500, '
0085  ' RAN(2,28,3)500, 500, 500, '
0086  C
0087  C  INITIALIZE CONTROLLING VARIABLES
0088  C
0089  V19=0, B39U) =100, B39<2) = 100, V20=8500, V21=3500, '
0090  V22=10000, V23=0, V53=S500, V54=5600, V55=5000, V56=7500, '
0091  V57=3500, '
0092  C
0093  CLB(1), CLB(2), V1=0, REPEAT(11), V1=V1+1, '
0094  B28(V1)=0, NxtXT I'
0095  CALL CHECK(IFL,2,0)
0096  IFdFL.EQ.1) GO. TO 499
0097  WRITE(1,56)
0098  56 FORMAT(//' MAIN TASK HAS BEEN SUCCESSFULLY COMPILED. '
0099  C
0100  C  LOAD THE LIMITS INTO HP-2250
0101  C
0102  IV(1)=27
0103  IV(2)=11
0104  IV<3)=IFIX((DP1+1.5)*2495.4+.30799E4) -
0105  IV(4)=IFIX((DP2+5.5)*2685+.32344E4)
0106  IV(7)=IFIX((DP1+2.0+.25094E1)/.12497E-2)
0107  IV(8)=IFIX((DP2+2.0+.24922E1)/.12368E-2)
0108  IV<5)=IFIX((T5*10)+300
0109  IV(6)=IFIX((T1+0*10)+300
0110  IV<9)=IV<6>
0111  IV<10)=IV<6>
0112  IV<11)=IV<6>
0113  IV<12)=IV<11>
0114  IV<13)=IV<11>
0115  CALL SECW(22,5, IV,13)
0116  C
0117  C  LOAD THE SET-POINTS INTO HP-2250
0118  C
0117  IV(1)=33
0120  IV(2)=4
0121  IV(3)=IFIX(V3*24.54+.3079*E+4)
0122  IV(4)=IFIX(DP2*2685+.3234*E+4)
0123  IMC5=IFIX(T5*10)
0124  IV(6)=IFIX(T10*10)
0125  CALL SECUC22,5,IV,6>
0126  C
0127  C LOAD THE CONTROLLER CONSTANTS INTO HP-2250
0128  C
0129  C
0130  IVC2>=3
0131  IVC(3)=IKC1
0132  IVC(4)=IKC2
0133  IVC(5)=IKC3
0134  CALL SECWC22,5,IV,5>
0135  IVU>=37
0136  IM(2)>=3
0137  IMC3=ITR1
0138  IMC(4)=ITR2
0139  IMC(5)=ITR3
0140  CALL SECWC22,5,IV,5>
0141  IV<1)=40
0142  IVC2>=3
0143  IVC(3)=IDT1
0144  IVC(4)=IDT2
0145  IVC(5)=IDT3
0146  CALL SECUC(22,5,IV,5>
0147  C
0148  C LOAD THE FIRST ORDER LAG CONSTS INTO HP-2250
0149  C
0150  IV<1)=30
0151  IVC2>=3
0152  IVC(3)=IFIX(V30*10)
0153  IVC(4)=IFIX(V31*10)
0154  * IV(5)=IFIX(V32*10)
0155  * CALL SECUC(22,7,IV,5>
0156  C
0157  C LOAD TASK(1) INTO HP-2250
0158  C
0159  C WRITE<22,202)
0160  202 FORHAT<" TASK<1>,V1=0,V2=0,V3=0,Y4=0,V5=0,U6=0,U7=0,V11=0,'/
0161  C
0162  C SET CLOCK IN HP-2250 AND BEGIN SAMPLING DATA.
0163  C
0164  * V6=0,V9=0,SCL=0,0,00,REPEAT<0),CTI,V7=0,'
0165  C
0166  C SAMPLE CONTROLLED VARIABLES AT EACH 1 SECOND.
0167  C
0168  * IN<18>,AI<1,4>,IN<20>,AI<1,8>,IN<23>,REF<2,16),'/
0169  * IN<18>,KTEMP<1,32),IN<18>,KTEMP<2,10),'/
0170  * V1=V1+1,IF VK5 THEN WNOU<0,0,80> ENDIF,'
0171  C
0172  C SAMPLE OTHER VARIABLES AT EACH 5 SECONDS. I
0173  C
0174  * IF V1<5 THEN GOTO(1) ENDIF,V1=0,IN<21),AI(2,30),'/
0175  * IN<17),AI(1,2),IN<17),AI(1,6),IN<23>,REF<2,32),'/
0176  * IN<18),KTEMP<2,1),IN<21),KTEMP<2,2),IN<23>,KTEMP<2,3),'/
0177  * IN<18),KTEMP<2,4),IN<26),KTEMP<2,6),INCB7),KTEMP<2,7),'/
0178  * IN<28),KTEMP<2,8),IN<28>,KTEMP<2,9),'/
REUND THE DATA BUFFERS TO PREVENT OVERFLOW OF MEMORY.

SCHEDULE THE TASK(1) TO RUN AT EVERY SECOND.

CALL CHECK(IFL,2,0)

IF(IFL.EQ.1) GO TO 499

WRITE(1,57)

LOAD THE TASK(2) INTO HP-2250.

SET THE FIRST ORDER LAG FILTER CONSTANTS

CHECK THE DAT A BUFFER IS JUST REWOUND.

CHECK THE POWER INTERRUPT IN THE PRESSURE TRANSMITTER

FINISH THE FIRST ELEMENT OF FIRST ORDER LAG

FILTER THE CONTROLLED VARIABLES BY FIRST ORDER LAG FILTER.
CHANGE THE MAXIMUM/MINIMUM OUTPUTS TO SOLIDS FLOW CONTROL VALVES AT HIGH TEMPERATURE

* IF B26(3)>3000 THEN V54=V56 ENDIF,'/
* IF B26(4)>4500 THEN V53=V55 ENDIF,'/

FILTER THE OTHER DATA BY AVERAGE FILTER.

* V5=1-V3,IF M6=1 THEN GOTO(IO) ENDIF,'/
* AAV(B17(V5)>V3,B26(5)),'/
* AAVCB19(V5),V3,326(6),AAV(B1KV5J,V3,B26(7)),'/
* AAV(B12(V5)>V3,B26(8)>,AAV(31(V5),V3,B26(9))),'/
* itAV(B25(V5),V3,B26(10)>,AAV(B3<Vj>,V3,B26(ll ))),'/
* AAV(B2< V5),V3,B26(10)>,AAV<B30>5),V3,B26(11)>,'/
* LABEL<10),V2=0,V3=0,V6=0,PAUSE,'/

CALIBRATE THE FUNCTION CARDS.

* ' IF V8=20 THEN CLB<1) ENDIF,'/
* ' IF V8=23 THEN CLB(2),M8=Q ENDIF,*'/

CHECK ANY EMERGENCY IN THE PROCESS

IF THERE IS ANY EMERGENCY, SET FLAG, INDICATOR AND ALARM.

* ' IF V83<>1 THEN GOTO(4) ENDIF,'/
* ' M10=8,REPEAT(3),V10=V10+1,IF B28(<V10)=1 THEN,'/
* ' 00(8,1)1 ENDIF,NEXT,LABEL(4),V10=fl,REPEAK8) ,'/
* ' V10=V10+1,IF 32B(V10)=1 THEN GOTO<5) ENDIF,NEXT,'/
* ' 00(8,1)0,LABEL(5),V10=0,'/

IF PRESSURE IS OUT OF LIMIT, STOP FEEDING

* ' P.REPEAT(4) ,V10=V10 + 1 ,IF B23(V10) = 1 THEN D0(8 ,7)1,'/
* ' ELSE DO<8,7)0 ENDIF NEXT,'/

SET INDICATORS ON THE CONTROL PANEL, IF TEMPERATURE IN THE RECOVERY SECTION IS TOO LOW.

* ' IF B28(9)=1 THEN D0(8,10)1 ELSE D0(8,10)0 ENDIF,'/
* ' IF B28(10)=1 THEN D0(8,11)1 ELSE D0<8,11)0 ENDIF,'/
* ' IF B28(11)=1 THEN D0(8,12)1 ELSE D0(8,12)0 ENDIF,'/

DECIDE THE CONTROLLED VARIABLES FOR PRESSURE TRANSDUCERS AND S.C.R (TEMPERATURE CONTROLLERS)

* ' PAUSE,V23=U23+1,V19=V19+1,'/
* ' V13=Q,REPEAT(3),V13=V13+1,B35(V13)=B33(V13)-B26(V13),'/
* ' IF V19<3 THEN B34(V13) =B35(V13),B41(V13)=B34(V13),'/
* ' ENDFD,IF V19)3 THEN V19=3 ENDIF,'/
* ' IF V13<3 THEN GOTO<11) ENDIF,'/
* ' IF V23<4 THEN GOTO<12) ENDIF,V23=0 ,'/
* ' LABEL(11),V27=2*B34(V13),'/
* ' V27=0~V27,V27=V27+B35(V13),V27=V27+B4KV13),'/
0293 ' V27=V27*840<V13), V27=V27/5, V27=V27*B37(V13>, />
0300 ' M25=B35<V13)-B34(V13), V25=V25/2, />
0301 ' V14=B35(V13)-B34P13), V15=5*V25, V14=V14*B37<V13), />
0302 ' IF V13=3 THKN V15=V15/5, V27=V27*5 ENDIF, />
0303 ' U16=V14+V15, V16=V16+V27, V16=V16*B37(V13), />
0304 ' B39(V13)=V16*B36(V13), V17=B39(V13)/5, />
0305 ' V17=M17*B38(V13), IF V13=3 THEN GOTO C9) ENPIF, />
0306 ' IF M17=0 THH.N GOTO <6) ENDIF, />
0307 ' IF V17=0 THEN GOTO <6) ENDIF, />
0308 ' IF V13=1 THEN IF V20>7600 THEN GOTO (6) ENDIF ELSE IF, />
0309 ' V20*6400 THEN GOTO <6) ENDIF, IF V13=1 THEN IF /
0310 ' V20<5400 THEN GOTO <6) ENDIF ELSE IF V20<5000 THEN, />
0311 ' GOTO <6) ENDIF ENDIF, />
0312 ' IF B39<V13)=0 THEN IF V13=1 THEN V18=V20/10, V18=V18+41, />
0313 ' V18=V18/10, V18=V18/10, V20=V20+y18 ELSE V18=V21/20, />
0314 ' V18=V18*37, U18=V18/5, V18=V18-7Q4, V18=U21=V21+V18 ENDIF, />
0315 ' ELSE IF V13=1 THEN V18=V20/20, V18=V18*29, V18=V18/5, />
0316 ' V18=V18-839, V20=V20-V18 ELSE V18=V21/10, V18=V18+27, />
0317 ' V18=V18/10, V18=V18+5U, V18=V21- V18 ENDIF ENDIF, />
0318 ' IF V13=1 THEN V20=V20+B39<13) ELSE V21=V21+B39<V13), />
0319 ' ENDIF, IF V17=0 THEN B39<V13)=B38<13) ENDIF, />
0320 ' GOTO (6), />
0321 ' LABEL(9), V22=V22+B39(V13), />
0322 ' LABEL(8), B4KV13) = B34(V13), />
0323 ' B34<V13)=B35(V13), B38<V13)=B39(V13), LABEL<12), />
0325 ' NEXT, IF V20>3500 THEN M20=S500, B38(V13) = 10 ENDIF, />
0326 ' IF V2K5 THEN V21=V57, B38<2)=10 ENDIF, />
0327 ' K ' IF V20<53 THEN V20-V53-ENDIF, />
0328 ' IF V21=V54 THEN V21-V54 ENDIF, />
0329 ' IF V13=3) THEN V20=V20<13) ELSE V21=V21+B39<13), />
0330 ' IF V22>1010 THEN V22=10000 ENDIF, />
0331 ' IF V22<0 THEN V22=0 ENDIF, />
0332 ' IF V22=0 THEN V22=0 ENDIF, />
0333 ' M50=V20, V51=V21, V52=V2?, OUT(V50), VO<7,1,3), />
0334 ' SCHEDULE THE TASK(2) TO RUN AT EACH 5 SECONDS. />
0335 ' CALL CHECK(IFL,2,0) /
0336 ' IF(IFL.EQ.1) GO TO 499 /
0337 ' WRITE(1,58) /
0338 ' 58 FORMATC TASK(2) HAS BEEN SUCCESSFULLY COMPILED. '///) /
0339 ' WRITEU,60) /
0340 ' 60 FORMATC PLEASE CHECK THE-EQUIPMENT./ /
0341 ' ' IF IT IS READY TO RUN,FEED THE SAND AND TYPE "1"./ /
0342 ' ' OTHERWISE,TYPE "0" TO STOP THE PROGRAM./ /
0343 ' ' IFUS.EQ.O) IS /
0344 ' READU,*) IS /
0345 ' IFUS.EQ.O) GO TO 499 /
0346 ' C /
0347 ' C /
0348 ' START THE TASK<1) TO RUN /
0349 ' C /
0350 ' WRITE(22,205) /
0351 ' 205 FORMATC STARTU) !') /
0352 ' CALL CHECK(IFL,2,0) /
0353 ' IF(IFL.EQ.1) GO TO 499 /
0354 ' C /
0355 ' WAIT FOR 5 SECONDS THEN CHANGE THE SYSTEM TO BE CONTROLLED /
0356 ' C /
0357 ' BY THE COMPUTER /
0358 ' C /
0359 ' CALL EXECU2, 0,2,0,-5)
0359 WRITEC22,211>
0360 211 FORMAT(8,5>1,D0(8,6)1,D0<8,8)1 !')
0361 CALL CHECKUFL,2,0>
0362 IF<IFL.EQ.1) GO TO 499
0363 C
0364 C START THE TASK(2) TO RUN
0365 C
0366 WRITEC22,2Q7>
0367 207 FORMAT(STARTC2) !'>
0368 CALL CHECKCIFL,2,0)
0369 ' IFCIFL.EQ.1) GO TO 499
0370 SCHEDULE DATA ACQUISITION PROGRAM TPDSL TO RUN
0371 CALL EXEC<10, NAMED
0372 CHECK THE STATUS OF TASKS IN THE HP-2250 AT EACH 15 SECONDS
0373 CALL EXEC<12,0,2,15,-15>
0374 CALL CHECKCIFL,3,1)
0375 ' IFCIFL.EQ.1) GO TO 499
0376 CALL CHECKCIFL,3,2>
0377 ' IFCIFL.EQ.1) GO TO 499
0378 GO TO 65
0379 CALL EXECC6,NAME2,2)
0380 STOP
0381 END
0382
0383 SUBROUTINE CHECK(IFL,N,M)
0384 C
0385 * THIS SUBROUTINE IS USED TO CHECK THE STATUS OF THE TASKS IN THE HP-2250.
0386 C
0387 C
0388 DIMENSION ISTAC8)
0389 IFL=1
0400 IF<NC.EQ.2> GO TO 30 0
0401 320 ISTAC1)M
0402 CALL SECW(22,N,ISTA,1)
0403 300. CALL SECW(22,N,ISTAJ8)
0404 306. IFCISTAC1).EQ.0) GO TO 314
0405 307 IF<ISTA(1).EQ.6> GO TO 322
0406 308 IFCISTAC1).EQ.7) GO TO 399
0407 IFCN.EQ.2) GO TO 300
0408 IFCN.EQ.3) GO TO 399
0409 322 WRITE<1,310) ISTAC2),ISTAC3),ISTA<7)
0410 310 FORMAT( ERROR CODE =',14,' AL ',14,' OF COMMANDS. V
0411 311 * ' IN TASK NUMBER ',14>
0412 312 RETURN
0413 314 URITEC1,316)
0414 316 FORMAT('< TASK DOES NOT EXIST I ')
0415 317 GO TO 312
0416 399 IFL=0

D164
APPENDIX E

PROGRAM TO STORE DATA ONTO DISKS

Program TPDSL-is scheduled to run by program DCTPF. It used to read the data kept in the interface, store the useful data onto the disk of the HP-1000 computer, and display a table describing the process. This program is scheduled to run every minute.
**TPDSL**

**PROGRAM TPDSL, HSU-20 JAN. 83**

**C**

**C** THIS PROGRAM IS USED TO READ DATA FROM HP-2250, STORE DATA INTO FILES, AND DISPLAY A TABLE DESCRIBING THE PROCESS. USUALLY, IT IS SCHEDULED TO RUN BY THE PROGRAM DCIPF. **#

**C**

**PROGRAM TPDSL**

**DIMENSION IUC10),11(195,4),I2<780),I2<39,25),W<22>,TMP(23) **

**DIMENSION I.F1<5),LF2<25>,TIMC195),INDC195,2),TIMN<39),IBU<20> **

**DIMENSION NAME1(3),NAME2(3),rSIZE<2),IDCB(144),D<8) **

**DIMENSION ITIHEC5) **

**COMMON ITX(23,12),TMP1(23),IBAD(20)**

**C**

**SET KNOWN DATA**

**DATA NAME1/2HFI,2HLF,2H /**

**DATA NAME2/2HTP,2HDS,2HL /**

**DATA ISIZE/-1,128/ .---**

**DATA U/22*l./**

**DATA D/.12497E-2,—.25094E+1, .40074E-3,—.12342E+1,.12365E-2, **

**—-.24922E+1,.37244E-3,—.12046E+1/ **

**DO 24 1=1,20**

**24 IBAD(I)=0**

**WRITE<1,14) **

**14 FORMATC ENTER * OF FILE TO STORE DATA, TWO INTEGER !) **

**READ(1,58> NAMEK3)**

**CREAT A FILE TO STORE DATA AND OPEN THAT FILE.**

**CALL CREAT<IDCB,IERR,NAMK3,ISIZE,3,-21,-9)**

**IF(IERR.LT.0) GO TO 90**

**CALL OPENPUDC8,IERR,NAMK3,0,-21,-9)**

**IFUERR.LT.0) GO TO 90**

**SET KNOWN VARIABLES**

**J0=195**

**J1=780**

**J2=39**

**NL1=0**

**INC=0**

**11I=0**

**RLT IS THE LAST SECOND TO TAKE THI. DATA FROM HP-2250 EACH TIHF.**

**READ(1,67)> RLT=0**

**WRITEU,67)**

**FORMAT(' ENTER FLOW RATE <LB/HR) ')'**

**READ(1,M TMP1(22))**

**THP<22)=THP1<22>**
INPUT THE STATUS OF THERMOCOUPLES

WRITEU,13>
FORMATC ENTER .BAD TH.' .'5.' '<0 TO CONTINUE.' >
READ(NTH) TH
IF(NTH.LT.0) GO TO 33
IBAD(NTH)=1
GO TO 23

INPUT THE DATA GAP TO STORE THE DATA
WRITEC1,43>
FORMATC ENTER TEMP. ,PRESS. AND TIME GAP TO STORE DATA ')
READ(*) GAPT,GAPP,GAPTI

INPUT THE WEIGHT OF THERMOCOUPLES
WRITEU,B3>
FORMATC ENTER FIRST ORDER LAG FITER CONSTANTS '/
READ(*) A,B,C

WAIT FOR 145 SECONDS

CALL EXECU2,0,Z,0,-145)
SCHEDULE THE FOLLOWING PROGRAM TO RUN AT EACH 55 SECONDS
CALL EXEC<12,0,2,55,-55)

THE FOLLOWING STATEMENTS ARE USED TO READ THE BUFFERS IN HP-2250,
THE DATA CONTAIN PRESSURE AND TEMPERATURE.

DO 10 1=1,25
IF(I.EQ.1) OR (I .EQ.10) GO TO 10
IF(U.EQ.1f1) OR (<I.EQ.20>) GO TO 10
IF(U.EQ.23) GO TO 10
IF(I.EQ.22) GO TO 11
IV(I)=I
IV(2)=J2

DO 10 1=1,25
IF(I.EQ.b) OR (I .EQ.10) GO TO 10
IF(U.Eq.1f1) OR (<I.EQ.20>) GO TO 10
IF(U.EQ.23) GO TO 10
IF(I.EQ.22) GO TO 11
IV(I)=I
IV(2)=J2
CALL SECW(22, 6, IV, 2)  
CALL SECR(22, 6, J2, I2, I)  
GO TO 10  

IV(1) = I  
IV(2) = J1  

CALL SECU(22, 6, IV, 2)  
CALL SECR(22, 6, I2, J1)  
CONTINUE

DO 18 I = 1, 2  
IV(1) = I5  
IV(2) = J0  
CALL SECU(22, 6, IV, 2)  
CALL SECR(22, 6, I1, J0)  
CONTINUE

IM = 0  
RPT = 0  
IS = 0  
N = 1, J0  
RNT = 3.6E+3*IT<4*N-3)+IT<4*N-2)+E-3*IT<4*N-1)+E-6*IT<4*N)  
TIM(N) = RNT/60.  
IF(ABS(RNT - RLT).LT.0.1) IS = N  
IF(RPT.GT.RNT) IH = N - 1  
RPT = RNT  
CONTINUE

IM = 1  
IF(IS.EQ.I) IFLA5 = 0  
IF(IS.EQ.I) IM = J0  
IH = IM + 2  
IS = IM1  
RFL = 3.6E+3*IT<4*IM-3)+IT<4*IM-2)+E-3*IT<4*IM-1)+E-6*IT<4*IM)  
JX = J0  
DO 80 M = 1, 2  
LX = L  
IF(IS.LT.IH) LX = IM - IS + 1  
IF(IS.LT.IH) LX = IM - IS + 1 + JX  
L2 = LX  
K = IS - 1  
DO 70 N = 1, LX  
K = K + 1  
IF(K.GT.JX) K = 1  
IND(N, M) = K  
CONTINUE

IF(M.GT.J) GO TO 80  
JX = J2  
L1 = LX  
IH = IFIXUM/5.)
IFCIH.LE.0) IM=J2
0179  IS=IFIX(IS/5.)     " 
0180  IF(IS.LE.0) IS=J2 
0181  IF((IS-IM).EQ.1.OR. (IS-IM).EQ.<J2-1)) IS=IM+2
0182  IF(IS-IM).EQ.0) IS=IH+2
0183  IF<IS.GT.J2) IS=IS-J2
0184  CONTINUE
0185  C FILTER THE BIG NOISE OUT
0186  C
0187  C MIT=100
0188  DO 19 N=1,20
0189  IF(N.GT.16) MIT=150
0190  LU=L1-1
0191  DO 15 1=1,LU
0192  IF<ABS(ai(IND<1,1),N)-I1<IND<1 + 1,1),N)).GT.MIT) II<INDCI+1,1),N)=
0193  >IM<IND<1,1),N)
0194  CONTINUE
0195  L22=L2-1
0196  DO 45 1=1,L22
0197  IF<ABS<IND<1,2),N)-I2<IND(1+1,2),N)).GT.MIT) I2<IND<1+1,2),N)=
0198  *I2<INDCI,2),N>
0199  CONTINUE
0200  45 CONTINUE
0201  19 CONTINUE
0202  C FILTER THE DATA BY FIRST ORDER 'LAG FILTER
0203  
0204  C DO 44 N=1,4
0205  IF(IFLAG.EQ.O) GO TO 42
0206  _ I1<IND(1,1),N)=LF1<N)
0207  IFCN.LE.2) CC=C
0208  IF(N.EQ.3) CC=A
0209  IF(N.EQ.4) CC=B
0210  CONTINUE
0211  47 LFKN)=I1UND(L1,1),N)
0212  CONTINUE
0213  44 CONTINUE
0214  DO 48 N=1,23
0215  IFCN.EQ.21) GO TO 48
0216  IFUNFLAG.EQ.O) GO TO 46
0217  I2<IND(1,2),N)=LF2<N>
0218  46 CONTINUE
0219  IFU.LE.16) CC=C
0220  IFCN.EQ.17) CC=A
0221  IF(N.EQ.19) CC=B
0222  J=IND(M,2)
0223  JM1=IND(H-1,1>
0224  I1<J,N)=IFIX<CC*I1<J,N> + <1-CC>*IKJM1,N)><1
0225  CONTINUE
0226  LF2<N)=I2<IND<1,2),N>
0227  CONTINUE
0228  48 CONTINUE
0229  DO 38 M=1,L2
0230  IFCN.EQ.19) CC=B
0231  J=IND(M,2)
0232  JM1=IND(H-1,2)
0233  I2<J,N)=IFIX<CC*I2<J,N> + <1-CC>*IKJM1,N)
0234  CONTINUE
0235  49 CONTINUE
0236  LF2<N)=I2<IND<1,2),N>
0237  CONTINUE
0238  48 CONTINUE
0239  C REARRANGE THF DATA BUFFER TO BE STORED
0240  C 
0241  DO 38 M=1,L2
0242  J=IND(M,2)5
0243  I2<IND<M,2),5)=I1(J,1)
0232  I2CIND(H,2>,10)=IHJ,2)
0240  I2(IND(H,2),18)=I1(J,3>
0241  • T2<IND<M,2>,20)=IMJ,4>
0242  TIMUND(H,2>>=TIH(J)
0243  38 CONTINUE
0244  IF(NL1.GE.1) GO TO 56
0245  C
0246  _C CONVERT THE RAW DATA INTO TEMPERATURE (DEG.C) AND PRESSURE (PSI)
0247  -C
0248  DO 16 1=1,16
0249  16 THP(I+1)=I2(IND(1,2),I)/10.
0250  K=0
0251  DO 86 1=1,7,2
0252  K=K+1
0253  66 THP(K+17)=D(I)*I2(INI>(1,2),K+16)+D<I+1>
0254  rMP(23)=TIHN(IND(1,2))
0255  IF(III.NE.1) GO TO 56
0256  NL1=NI_1+1
0257  TMP(1)=FLOAT(NL1)
0258  C
0259  C STORE THE DATA INTO DISKS
0260  C
0261  CALL URITF(IDC8,IERR,THP,46>
0262  IF(IERR.LT.0) GO TO 90
-0263  C _>"
0264  C "DETERMINE THE DATA TO BE STORED BY DATA GAP
0265  C
0266  56 - DO 26 N=1,L2
0267  IPR=0
0268  DO 36 1=1,16
0269  TP=I2<IND(N,2>,I)/10.
0270  DIF=ABS(TMP(I-t-1>-TP)
0271  GAP1=GAPT'/U(I)
0272  " IF(DIF.LE.GAP1) GO TO 36
0273  IF(IBAD<1>.EQ.1) GO TO 36
0274  IPR=1
0275  36 CONTINUE
0276  K=0
0277  DO 66 N=1,7,2
0278  K=K+1
0279  TP=D(I)*I2(IND(N,2),K+16)+D(I+1>
0280  DIF=ABS(TMP(K+17)-fp)
0281  GAP1=GAPP/W(K+16)
0282  IF(DIF.LE.GAPP1) GO TO 66
0283  IF(IBAD(K+16).EQ.1) GO TO 66
0284  IPR=1
0285  66 CONTINUE
0286  DIF=(TIMNCIND(N,2>)-THP(23>)
0287  GAP1I=GAP1/W<22>
0288  IF<DIF.GT.GAP1I) IPR=1
0289  IF(U.II.NE.1) IPR=0
0290  IF(IPR.NE.1) GO TO 26
0291  C
0292  _C CONVERT THE RAW DATA INTO TEMPERATURE (DEG.C) AND PRESSURE (PSD
0293  C
0294  DO 17 1=1,16
0295  17 _HP(I+1)=I2(IND(N,2),I>/10.
0296  K=0
0297  DO 27 1=1,7,2
0298  K=K+1

D171
0299  27   TMPCK+17)=D<12<INTMN,2>,K+16)+D<1+1>
0300   THP(23)=TIMN(IN(N,2))
0301   .
0302   NL1=NL1+1
0303   m?(l)»FLOAT(NL1)
0304   C
0305   C   STORE THE DATA INTO DISKS
0306   C
0307   CALL URI<T(IF,IC,ERR,TMP,46)
0308   IFCIERR.LT.0) GO TO 90
0309   CONTINUE
0310   C
0311   C   STORE THE LAST DATA INTO THE BUFFERS TO BE DISPLAYED BY A TABLE
0312   C
0313   IF<INC.EQ.12) GO TO 57
0314   DO 31 1=1,16
0315   THPK1+1) = D<IND(L2,2),K+16>+D(1+1>
0316   TTX(I+1,INC) = IFIX(THPKI-H))
0317   CONTINUE
0318   K=0
0319   DO 41 1=1,7,2
0320   K=K+1
0321   TMP1(K+17)=D<12<IND(L2,2),K+16>+D(H-1>
0322   TTX(K+17,INC) = IFIXCTMP1(K+17>"100)
0323   CONTINUE
0324   ITX(22,INC)=IFIX<THPU22)*100>
0325   ITX(1,INC)=NL1
0326   " ITX<22,INC)=IFIX<TMPM22>*10>
0327   " ITX<23,INC)*IFIX<THNN<IND(L2,2>>
0328   THPK1 )=FLOAT(NL1)
0329   TMP1<23)=TIMN<IND(L2,2>)
0330   C
0331   C   DETERMINE THE STATUS OF THE TABLE BY THE REAL-TIME DIGITAL INPUTS
0332   C
0333   IF1=0
0334   IF2=0
0335   DO 25 N=1,L2
0336   IF<12<IND<N,2),E4).GT.100) IF1=1
0337   IF<12<IND<N,2),25).GT.100> IF2=1
0338   CONTINUE
0339   ITF=IF1+2+IF2
0340   IFCITF.EQ.2) 111=1
0341   C
0342   C   CALL THE TABLE DISPLAYING PROGRAM
0343   C
0344   CALL DISP(I,ITF)
0345   C
0346   C   REARRANGE THE BUFFERS FOR DISPLAYING IN THE TABLE
0347   C
0348   IFUNC.LT.12) GO TO 71
0349   DO 77 1=1,23
0350   DO 87 M=1,11
0351  87   ITX<1,M) = II"X(I,H+1>
0352  77   CONTINUE
0353  71   CONTINUE
0354   C
0355   C   IFCITF.EQ.3) CO TO 35
0356   C
0357   C   RETURN TO EXEC CALL FOR ANOTHER RECYCLE!
0358   C
0359   GO TO 30
C INDICATE THE END OF DATA IN THE DISKS
0360 C
0361 35 TMP<1)= -1.
0362 CALL URirF(IDC,B,IERR,THP,46>
0363 IFUERR.LT.0) GO TO 90
0364 INPUT THE COMMENTS DESCRIBING THE DATA STORED
0365
0366 WRITEU,();
0367 FORMAT( ENTER * OF COMMENT LINES U<0 TO STOP))
0368 READ<1,*-> LIC
0370 IBU(i)=NL1
0371 IBIK2)=23
0372 IBU<3>=LIC
0373 CALL URITMIDCB,IDKR,IBU,20)
0374- IFUERR.LT.0) GO TO 90
0375 ~IF<.LIC.LT.0) GO TO 88
0376 OO 78 1 = 1,1.10
0377 WRITEU,68)
0378 68 FORMAT( ENTER A COMMENT (39 CHAR. MAXIMUM) )
0379 READ(1,58) (IBIKM),H-1,20)
0380 58 FORMAT<20A2)
0381 CALL URITFUDCB,IDRR,IBU,20>
0382 C IF(IERR.LT.0) GO TO 90
0383 78 CONTINUE
0384 C CHOP THE UNUSED TRACKS IN THE DISKS
0385 C
0386 88 CALL 0 C F(IDC>B,IE:RR71J1,1 JB.1 J2,J SEC)
0388 IFUERR.LT.0) GO TO 90
0389 --- ITR = (JSEC/2)-UR3+l>
0390 CALL CLOSE(IDC3,1FRRR^ITR)
0391 IFdERRRR.LT.0) GO TO 90
0392 GO TO 99?
0393 90 CALL CLOSE,UDCBUERRR)
0394 91 WRITEI55)IERRR,JERR«
0395 55 FORMAT< FMP-ERROR= ',14,4 X,' CLOSING ERROR= ',14)
0396 999 CALL EXEC(6,NAME2,2)
0397 END
0398 END*
LITERATURE CITED


Auldridge, L. "Oil Reserves Reduced, but Production Climbs." *Oil and Gas Journal* 74 (Dec. 27, 1976): 105.


