WRI 09-R007

DEVELOPMENT OF THE WRITE™ PROCESS FOR PIPELINE-READY HEAVY OIL

TOPICAL REPORT

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And
U.S. Department of Energy
National Energy Technology Laboratory
Morgantown, West Virginia

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Task 51

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DISCLAIMER

ABSTRACT

Work completed under this program advances the goal of demonstrating Western Research Institute's (WRI's) WRITE™ process for upgrading heavy oil at field scale. MEG Energy Corporation (MEG) located in Calgary, Alberta, Canada supported efforts at WRI to develop the WRITE™ process as an oil sands, field-upgrading technology through this Task 51 Jointly Sponsored Research project. The project consisted of 6 tasks: (1) optimization of the distillate recovery unit (DRU), (2) demonstration and design of a continuous coker, (3) conceptual design and cost estimate for a commercial facility, (4) design of a WRITE™ pilot plant, (5) hydrotreating studies, and (6) establish a petroleum analysis laboratory.

WRITE™ is a heavy oil and bitumen upgrading process that produces residuum-free, pipeline ready oil from heavy material with undiluted density and viscosity that exceed prevailing pipeline specifications. WRITE™ uses two processing stages to achieve low and high temperature conversion of heavy oil or bitumen. The first stage DRU operates at mild thermal cracking conditions, yielding a light overhead product and a heavy residuum or bottoms material. These bottoms flow to the second stage continuous coker that operates at severe pyrolysis conditions, yielding light pyrolyzate and coke. The combined pyrolyzate and mildly cracked overhead streams form WRITE™'s synthetic crude oil (SCO) production.

The main objectives of this project were to (1) complete testing and analysis at bench scale with the DRU and continuous coker reactors and provide results to MEG for process evaluation and scale-up determinations and (2) complete a technical and economic assessment of WRITE™ technology to determine its viability. The DRU test program was completed and a processing envelope developed. These results were used for process assessment and for scale-up. Tests in the continuous coker were intended to determine the throughput capability of the coker so a scaled design could be developed that maximized feed rate for a given size of reactor. These tests were only partially successful because of equipment problems. A redesigned coker, which addressed the problems, has been build but not operated.

A preliminary economic analysis conducted by MEG and an their engineering consultant concluded that the WRITE™ process is a technically feasible method for upgrading bitumen and that it produces SCO that meets pipeline specifications for density. When compared to delayed coking, the industry benchmark for thermal upgrading of bitumen, WRITE™ produced more SCO, less coke, less CO_2 per barrel of bitumen fed, and had lower capital and operating costs. On the other hand, WRITE™'s lower processing severity yielded crude with higher density and a different product distribution for naphtha, light gas oil and vacuum oil that, taken together, might reduce the value of the SCO. These issues plus the completion of more detailed process evaluation and economics need to be resolved before WRITE™ is deployed as a field-scale pilot.

EXECUTIVE SUMMARY

A preliminary economic analysis conducted for MEG Energy Corporation (MEG) by Triumph EPCM (Triumph) concluded that Western Research Institute's WRITE™ process is a technically feasible method for upgrading bitumen and that it produces synthetic crude oil (SCO) that meets pipeline specifications for density. When compared to delayed coking, the industry benchmark for thermal upgrading of bitumen, WRITE™ produced more SCO, less coke, less CO_2 per barrel of bitumen fed, and had lower capital and operating costs. On the other hand, WRITE™'s lower processing severity yielded a crude with higher density and a different product distribution for naphtha, light gas oil and vacuum oil that, taken together, might reduce the value of the SCO. These issues plus the completion of more detailed process evaluation and economics need to be resolved before WRITE™ is deployed as a field-scale pilot.

MEG continues to support WRITE™ technology and has funded the construction of 5-bpd, engineering-scale, distillate recovery and continuous coker reactors. MEG is also funding tests in these reactors that will provide the information needed for a detailed analysis of the process.

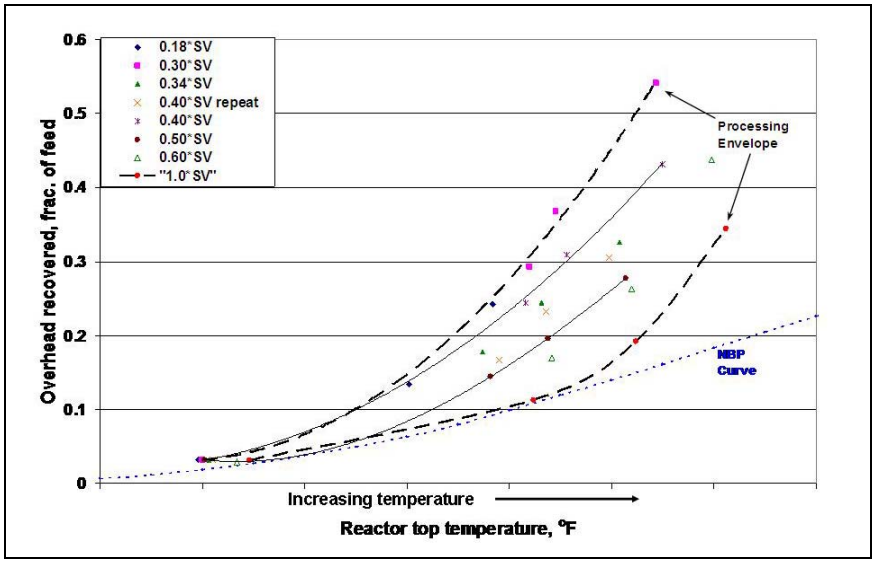
WRITE™ is a field-deployable heavy oil and bitumen upgrading process that produces residuum-free, pipeline ready oil from heavy material with undiluted density and viscosity that exceed prevailing pipeline specifications. WRITE™ uses two processing stages to achieve low and high temperature conversion of heavy oil or bitumen. The first stage distillate recovery unit (DRU) operates at mild thermal cracking conditions, yielding a light overhead product and a heavy residuum or bottoms. These bottoms flow to the second stage continuous coker that operates at severe pyrolysis conditions, yielding light pyrolyzate and coke. The combined pyrolyzate and mildly cracked overhead streams form WRITE™'s SCO production.

Work completed under this Task 51 Jointly Sponsored Research project advances the goal of demonstrating WRI's WRITE process for upgrading heavy oil and bitumen at field scale. MEG located in Calgary, Alberta, Canada co-sponsored the effort, the objectives of which are:

- Complete testing and analysis at bench scale with WRI's Distillate Recovery (DRU) and continuous coker units and provide results to MEG's engineers (and their consultants) for process evaluation and scale-up determinations
- Complete a technical and economic assessment of WRITE™ technology to determine its viability
- Develop reference designs that allow implementing the WRITE™ process at commercial scale
- Complete the design of a field-scale pilot facility to demonstrate WRITE™
- Establish and staff a petroleum analysis laboratory to support oil upgrading research at WRI

The program comprises six tasks: (1) optimization of the distillate recovery unit (DRU), (2) demonstration and design of a continuous coker, (3) conceptual design and cost estimate for a commercial facility, (4) design of a WRITE™ pilot plant, (5) hydrotreating studies, and (6) establish a petroleum analysis laboratory. The work outcomes are discussed below.

The objective of DRU testing was to determine conditions that optimized overhead yield. The range of tests conducted with the bench-scale DRU yielded 30 to 50wt% (relative to feed) of a light distillable overhead that meets pipeline specifications for density and viscosity, while maintaining a consistent quality that is relatively insensitive to process severity. The figure below shows the range of reactor temperature and residence time (expressed as normalized space velocity [SV]) conditions for all tests conducted in the bench-scale DRU. The curves shown as dashed lines represent the minimum and maximum range of conditions explored. Taken collectively, these data define the processing envelope for the bench-scale DRU. The quantity of overhead produced correlated positively with increased temperature and reduced space velocity but was independent of sweep gas composition. (The use of sweep gas is essential to WRITE™ processing). The dependence of yield on increased temperature and residence time suggests that DRU processing is kinetically controlled, and analysis of data demonstrated that overhead production followed first order kinetics, which is expected for hydrocarbon reactions. We concluded that the DRU functions as a mild thermal cracking process because of the overhead oil's light distillate nature and yield follows reaction kinetics.



Processing Envelope Resulting from DRU Testing

Tests conducted in WRI's bench-scale continuous coker were expected to provide information regarding its maximum conversion rate as a function of temperature. This relationship will aid in designing a reactor that maximizes throughput at a minimum size. Tests conducted up to 875°F demonstrated that throughput of DRU bottoms increases with temperature. However, the tests could not be completed because the coker's recovery system flooded at higher temperatures. This shortcoming has been addressed in a revised 5-bpd design.

The results from DRU and continuous coking tests were provided to MEG's process engineers and consultants. These data aided in the development of a reference design for the DRU and an improved design for the continuous coker that allowed higher temperature operation. The reference design for the DRU implements WRITE technology with processes and control methods used in commercial scale equipment. MEG used these designs to construct the 5-bpd engineering-scale MDRU and continuous coker reactors. The MDRU currently operates at WRI, producing additional process data for use in feasibility and scale-up calculations.

The reference design was also used to develop a design for a field-scale DRU pilot that would be sited at MEG's production facility in Canada. The continuous coker is not currently included in the design because insufficient data and experience has been developed at this time to demonstrate its viability as a commercial process, although provision has been made for its inclusion. A design bid memorandum is completed for the majority of the facility that does not include the reactor. This design will be completed when the decision on throughput capacity of the facility is finalized.

Hydrotreating studies were conducted to determine the extent of hydrotreating required to enable the naphtha fraction of DRU produced overhead oil to meet minimum pipeline specifications, corresponding to 1gm of bromine per 100gm of oil. Hydrotreating conditions varied from 550 to 700ºF and pressures from 1300 to 1950 psi. All conditions reduced the bromine number to a value less than one. The condition that exhibited the least hydrogen uptake of 60scf/bbl (and therefore considered optimum) occurred at a temperature and pressure of 500ºF and 1950 psi. Recent developments in catalyst technology should allow equivalent reduction in bromine number with lower severity processing.

MEG and their consultant Triumph performed a technical evaluation and economic screening of a presumed WRITE™ field-scale upgrading complex that processes 100,000-bpd of bitumen. The study compared WRITE™ to delayed coking, the industry accepted standard for thermal upgrading of bitumen. The study confirmed the technical feasibility of using WRITE™ as a process for upgrading bitumen. The study also found advantages and challenges for WRITE™, as discussed earlier in this summary. The study noted important differences between WRITE™ and delayed coking that are summarized in the following table.

Important Differences Between WRITE™ and Delayed Coking

Item	Delayed Coking	WRITE™ Process	Comment
Yield of SCO	Base	Same	Expect WRITE™ to be 3 lv% higher. Requires steady-state pyrolyzer operation.
Gravity	29.2	24.7	Delayed Coking SCO is lighter
Product Quality	Base	Heavier, higher sulphur	Heavier, higher sulphur SCO product from WRITE™
Coke Make	Base	Base – 20 wt%	Lower coke make with WRITE™
Fuel Gas Make	Base	Base – 40%	Lower intensity cracking with WRITE™
CO_2 production	Base	Base – 30 wt%	Lower CO_2 production with WRITE™
Make – Up Water	Base	Base – 20 wt%	Lower make-up water use for WRITE™
Capital Cost	Base	Base – 24%	Lower capital costs for WRITE™
Fixed Cost	Base	Base – 24%	
Technical Risk	Low	Medium	The pyrolysis unit of the WRITE™ process is unproven
other	Commercial	Not commercial	Delayed coking is industry proven

The positive indicators for WRITE™ notwithstanding, the study advised that a comprehensive test program needs to be conducted to provide additional information for more detailed process evaluations and associated economics. These evaluations must be completed before WRITE™ is deployed as a field-scale pilot. The proposed test program would include operating the 5-bpd engineering-scale reactors with bitumen produced from MEG's production facilities, characterizing the performance of the continuous coking reactor, and operating the continuous coker for extended periods of time to gain operating experience and to produce sufficient coke for combustion and gasification studies. The study also emphasized that finding cost effective and environmentally acceptable reuse strategies for coke is the key to the success of carbon rejection processes, such as WRITE™.

A petroleum analysis laboratory was established and staffed to support heavy oil upgrading activities at WRI. The laboratory currently supports the operation of the 5-bpd engineering-scale reactor. Analyses performed by the laboratory include bromine number, atmospheric distillation of petroleum products, shear and kinematic viscosity, density, micro carbon residue, determination of asphaltene concentration, simulated distillation, volatile matter in petroleum coke, elemental analysis (excluding oxygen), and hydrocarbon types.

INTRODUCTION

The U.S. remains dependent upon foreign sources of oil even after some 30 years since the oil embargo of the 1970's. Currently, the U.S consumes 20.7 million barrels per day (b/d) of the worldwide consumption of 84 million b/d and is projected by the Energy Information Administration to consume another 5.4 million barrels per day of oil globally by 2025. This is in light of increased international demand for oil by China and India of another 7.8 million barrels per day of oil by 2025 (Clark, 2007).

A Federal Task Force on Unconventional Fuels (established under the Energy Policy Act of 2005) concluded that the high cost and volume of oil imports have worsened the nation's trade deficit, weakened the dollar against other currencies, and put national security and economic stability at risk (Clark, 2007). The Task Force recommended the development of a domestic unconventional fuels program, taking advantage of oil shale, tar sands, coal-to-liquids, and heavy oil resources, the major resource being oil shale in Colorado, Utah and Wyoming. The estimated growth of an unconventional fuels program by 2035 was only 7.5 million barrels per day, including an optimistic 2.6 million from coal-to-liquids (CTL) and an increase of 1.3 million barrels per day from EOR via CO_2 injection. This barely keeps pace with the growth for demand, and leaves the US in the same position in 2035 as it is today as far as oil imports.

The continuing trend of high oil imports from unfriendly and unstable regions of the world argues for other sources for energy security. One possibility is increasing the use of the imports from friendly neighbors, such as Canada and their oil sands resources. The Canadian oil sands contain an estimated 2.5 trillion barrels of bitumen, 20% more than the total oil shale, tar sands and heavy oil reserves in the US combined. Assuming 40% in situ recovery via steam assisted gravity drainage (SAGD) would yield approximately 1 trillion barrels of recoverable heavy oil, enough to cover total oil needs of the US under current consumption rates for over 50 years.

The Canadian oil sands industry is rapidly increasing its production to meet the demands for oil. However this growth will increase the strain on existing resources and infrastructure. Significant increases in production of non-upgraded bitumen will place heavy demands first on condensate and then on syncrude for use as diluent. The fact that synbit and dilbit blends require between 30 to 50% diluent will lead to decreased efficiency in pipeline transport because a substantial fraction of capacity will be moving recycled diluent. The increasing number of SAGD projects will strain the natural gas system to deliver the fuel needed for steam generation and other process needs.

Currently, the limited coking capacity of US refineries presents an impediment to marketing heavy oil produced in Canada. A potential method for addressing this issue is the development and deployment of cost effective field upgrading technologies, such as WRITE™. Field upgrading refers to processes sited at the point of production that upgrade the bitumen to a higher value product slate. The upgraded product exhibits a reduced viscosity that makes the

crude amenable for pipeline transport without the addition of diluent. Current field upgrading involves the modification of the product's H/C ratio either by rejecting carbon via a coker or by adding hydrogen or both. With cokers, the coke disposition needs to be understood and its potential use as a fuel for SAGD to supplement the limited supply of natural gas. In the case of hydrogen addition, a source of hydrogen must be found, which may come from methane reforming or coke gasification.

OBJECTIVES

Work completed under this program advances the goal of demonstrating Western Research Institute's (WRI's) thermal enhancement process (WRITE™) for upgrading heavy oil at field scale. MEG Energy Corporation (MEG) located in Calgary, Alberta, Canada supports efforts at WRI to develop the WRITE™ process as an oil sands, field-upgrading technology through this Task 51 Jointly Sponsored Research project. Objectives for Task 51 are:

- Complete testing and analysis at bench scale with WRI's Distillate Recovery (DRU) and continuous coker units and provide results to MEG's engineers (and their consultants) for process evaluation and scale-up determinations
- Complete a technical and economic assessment of WRITE™ technology to determine its viability
- Develop reference designs that allow implementing the WRITE™ process at commercial scale
- Complete the design of a field-scale pilot facility to demonstrate WRITE™
- Establish and staff a petroleum analysis laboratory to support oil upgrading research at WRI

BACKGROUND AND TECHNICAL APPROACH

The primary objective of this program is to provide sufficient experimental data and technical analysis to design a demonstration-scale WRITE™ facility (for upgrading heavy oil or bitumen) that can be placed at MEG's production site in Canada. Working toward accomplishing this objective required the completion of a number of subtasks that will be described in this report.

WRITE™ Processing Concept

WRITE™ is a field-deployed heavy oil and bitumen upgrading process used to produce residuum-free, pipeline ready oil from heavy material with undiluted density and viscosity that exceed prevailing pipeline specifications. WRITE™ uses two processing stages to achieve low and high temperature conversion of heavy oil. The first stage distillate recovery unit (DRU)

operates at mild thermal cracking conditions, yielding a light overhead product and a heavy residuum. These bottoms flow to the second stage continuous coker that operates at severe pyrolysis conditions, yielding light pyrolyzate and coke. The combined pyrolyzate and mildly cracked overhead streams form WRITE™'s synthetic crude oil (SCO) production. The WRITE™ process' functional steps for upgrading a diluted bitumen stream are shown as a block diagram in Figure 1.

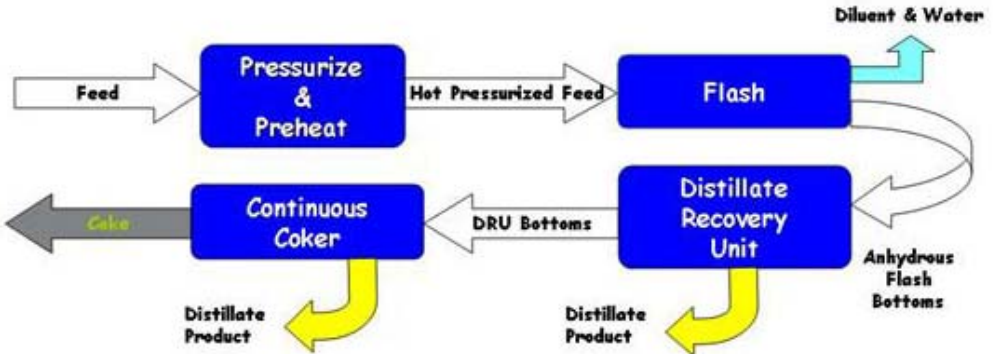


Figure 1. Processing Diagram for WRITE™

Work Plan for Development of the WRITE™ Process

Development of the WRITE™ process is expected to proceed in three phases. Phase 1 involves the completion and analysis of bench-scale tests, the design of a pilot facility for field testing, and development of preliminary economics for a commercial facility. Work outcomes completed under Phase 1 are described in this report. Phases 2 and 3 involve the construction and operation of a field-scale pilot plant that will be sited at a production facility using actual run bitumen. The Phase 1 program for WRITE™ process development comprises six tasks:

1. Optimization of the distillate recovery unit. Complete testing with the bench-scale DRU using raw and undiluted Cold Lake, Athabasca bitumen. In cooperation with MEG's process engineering consultants, analyze results from current and previous tests to generate information for technical evaluation and scale up. Evaluate chemical engineering unit operations to determine how WRITE™ could best be implemented for pilot, demonstration and commercial facilities and translate these findings into a reference design.
2. Demonstration and design of a continuous coker. Conduct a limited series of tests with bottoms from processing Cold Lake bitumen and obtain sufficient operating experience and process data to develop a preliminary reference design for the continuous coker. Where possible, implement changes to the existing bench-scale coker and continue testing to obtain additional data. In cooperation with MEG's consults, evaluate the process and generate information for scale up. Evaluate alternative methods of coking and develop a final reference design as in Task 1.

3. Conceptual design and cost estimate for a 20000 bpd facility. Use information from Tasks 1 and 2 to develop a conceptual design of a commercial facility. Develop capital and commercial costs for the facility to determine the viability of using the WRITE™ process to upgrade bitumen.
4. Design the pilot plant. Using reference designs developed earlier develop a detailed design for a pilot plant. The design will include complete process layout including P&ID's, a bid package for construction and a pre-contract estimate of capital and operating costs.
5. Hydrotreating studies. Conduct hydrotreating studies to determine the minimum conditions necessary to stabilize olefins, diolefins and other unsaturated compounds in synthetic crude produced by the DRU and continuous coker such that pipeline specifications are met.
6. Establish a petroleum analysis laboratory. Establish and staff a petroleum analytic laboratory to provide cost effective and timely analytical results to support the development of ongoing and future hydrocarbon recovery and conversion technologies by WRI and its collaborators.

Lead responsibilities for the tasks were allocated as follows. WRI assumed lead for conduct of the test programs in the bench-scale reactors and the hydrotreating studies. MEG, their consultants, and WRI cooperated in evaluating process data and developing reference designs for the DRU and continuous coker. MEG and their consultants assumed the lead for the conceptual commercial design and economics as well as design of the pilot plant. MEG and WRI shared responsibilities on the petroleum analysis research laboratory. MEG acquired the services of SNC-Lavalin to evaluate results from the DRU and continuous coker and to develop the reference designs for same. Triumph EPCM (Triumph) conducted the preliminary commercial-scale economic screening studies and compared the performance of WRITE™ with other heavy oil upgrading technologies. Triumph also developed the design for the pilot plant facility.

Bench-Scale Equipment Used for Testing

WRI has used its bench-scale test equipment described below to determine the compositions and yields of products expected when upgrading various heavy crudes with the WRITE™ process. This equipment consists of a nominally one barrel facility designed to simulate the performance of the DRU and a six-inch inclined rotary screw reactor designed to simulate the continuous coker. Both pieces of equipment are located at WRI's Advanced Technology Center (ATC) in Laramie, Wyoming.

DRU test facility

A schematic of the bench-scale equipment used to simulate the DRU is shown in Figure 2. Heavy oil or bitumen flows from a feed tank through a pump that pressurizes the material into an electrically heated feed pre-heater (Stripper Unit 1). A pressure let-down valve

(flash valve) controls the pressure. The separator removes the water as vapor where it subsequently condenses as overhead in KO-1 as Product 1. Substantially water-free material flows successively from the bottom of the flash tank through four Stripping Units. Each unit is an electrically heated vessel with its own temperature controller, sweep gas provisions, and equipment for product recovery. Glycol-cooled heat exchanger units trap the lowest boiling oil fractions. Reactor heating is done with calrod-type electrical elements that are completely submerged in the oil. Gases produced in any of the heated vessels can be sampled and analyzed. The material of construction used throughout the system is type 316 L stainless steel. This selection was in part dictated by the fact that some of the feedstocks used in earlier investigations contained high concentrations of chlorides and sulfur. Previous refinery experience indicated that type 316 L is adequate for this service. All flows in and out of the bench-scale test unit are monitored and continuously logged by computer, as are all temperatures and pressures.

Reactor temperature control for the individual stages is done with thermocouples that contact the electrical elements at the center of their length. Oil exit temperature for each reactor unit is measured using a thermocouple located on the transfer line between reactor units. In the latter stages of testing under this program, we measured oil temperatures for units 3, 4, and 5 by locating thermocouples in the oil bath at the center point relative to reactor's length and radially at the mid-point between the outer-most heater element and inside wall.

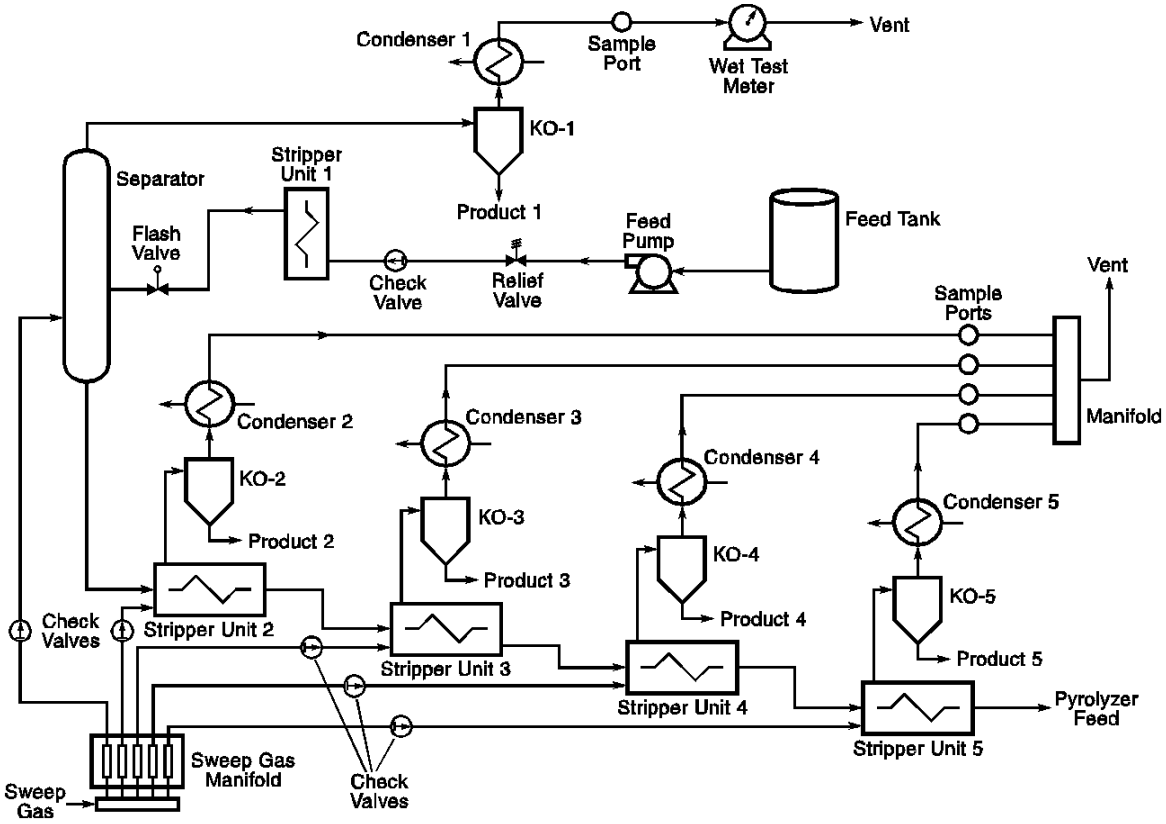


Figure 2. Schematic of the WRITE™ Process Bench-Scale Flash-Stripper Unit.

Continuous coker test facility

A 6-inch twin rotary screw reactor simulates processing in the continuous coker. The equipment consists of an electrically heated twin screw that can process heavy crudes, and bottoms material from the DRU at three progressively higher temperatures along the reactor's length. The temperatures of the three heating zones are controlled with individual clamshell heaters that encircle the barrel of the screws. Temperature control is accomplished by thermocouples located on the surface of the reactor's barrel. Interior temperatures are measured using thermocouples that penetrate to a point just inside the interior wall of the reactor. The overhead liquid product recovery train consists of condensers and knock-out pots with provisions for gas sampling. A final stage, glycol-cooled heat exchanger recovers the lowest boiling oil fraction. A gear pump feeds the screw. Liquid level is maintained by injecting oil at the point corresponding to the desired location in the reactor. Samples of processed solids may be collected during the test. A computer system controls and records the temperatures, pressures and other related conditions. A schematic of the reactor is shown in Figure 3.

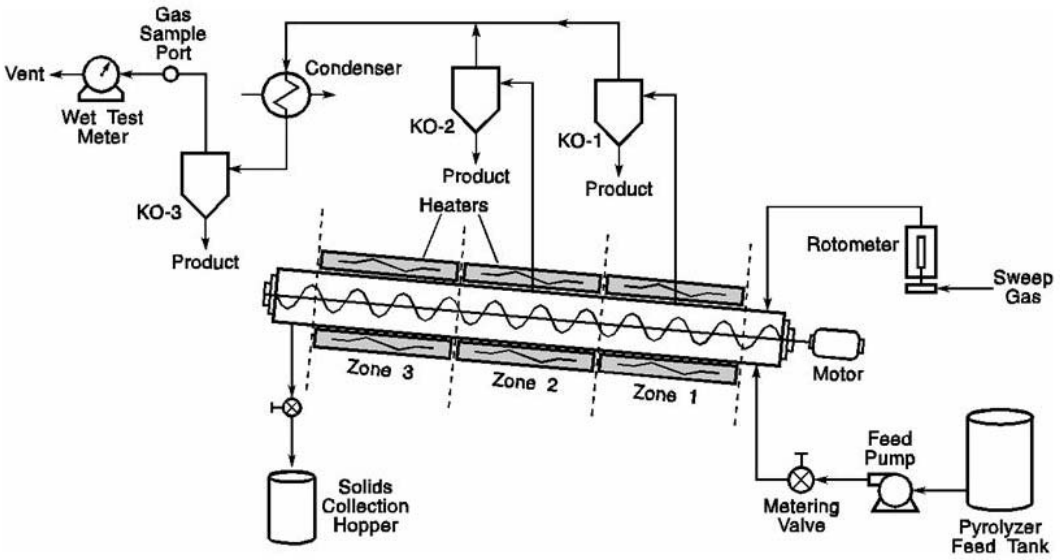
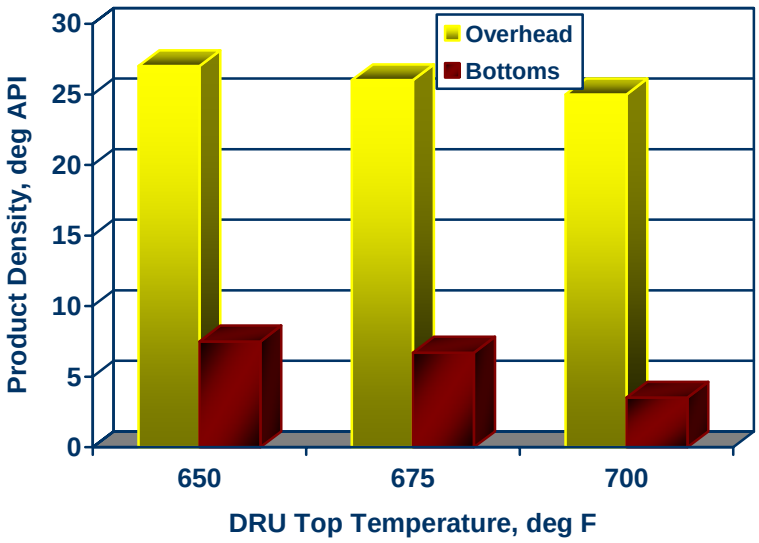


Figure 3. Schematic of the WRITE™ Process Bench-Scale Continuous Coker.

DRU OPTIMIZATION TESTS

While performing early evaluations of the WRITE™ process, the distillate recovery unit was presumed to function as a vacuum still to thermally separate the higher and lower boiling fractions of the incoming feed. Operating in this fashion, the unit should collect approximately 20% of Cold Lake bitumen as an overhead distillate while the remaining 80% would be bottoms to be fed to the continuous coker. During the conduct of a JSR program with the National Centre for Upgrading Technology (NCUT) to assess the stability and compatibility of oils produced by the WRITE™ process (Brecher, 2008), hereafter referred to as Stability Program, WRI observed overhead yields in excess of 30wt% when operating its bench-scale reactor that was designed to simulate the operating conditions of the DRU.



Of equal significance, product quality of the overhead distillate, shown in the pictorial above, as measured by density and viscosity, remains nearly constant when processing temperature increases. Conversely, quality of the bottoms product, measured relative to these same two parameters, decreases with increased temperature.

Differential balances of the incrementally produced oil, compared to its normal boiling curve, indicated that the DRU products are enriched in material boiling at temperatures less than 850ºF and depleted in materials boiling higher than that temperature. This performance indicated that the DRU functions not only as a device for physically separating the oil on the basis of its constituent's boiling points, but also as a chemical reactor, operating under mild oil pyrolysis or thermal cracking conditions. Operating as a reactor, the yield of distillate collected overhead will depend on the DRU's temperature profile as well as other processing characteristics such as the oil's residence time, sweep gas composition, and sweep gas rate.

Conduct of the DRU optimization test program

Testing under Task 51 expands on data obtained from the previously conducted Stability Program. Tests under the current program characterized DRU performance over a wider range of severity and determined (with the inclusion of earlier test results) the operating envelope for upgrading bitumen under mild thermal cracking conditions. The term "mild conditions" as used here implies that no coke is formed. "Severity" refers to conditions that drive reactions in a kinetic sense, which requires increased temperature or decreased space velocity or both. For consistency with the Stability Program, testing was done with both undiluted and diluted (dilbit) bitumen crudes from EnCanna's Foster Creek operations (see Table 1 for a typical oil analysis). Independent variables for the test program included sweep gas composition, temperature, and space velocity, which is the inverse of residence time. Testing comprised three series:

- The first series, run with undiluted bitumen, explored the effects of sweep gas composition and space velocity.
- The second series run with diluted and undiluted bitumen, varied temperature and space velocity.
- The third series, run with dilbit, explored a range of reduced space velocities.

MEG and WRI conducted an additional test series to confirm overhead product yield for an extended period of operation. These tests were totally funded by MEG.

To maintain consistency among tests conducted under Task 51 and all previous studies, the reactor residence times, expressed in terms of space velocity, were normalized relative to the maximum feed rate used in the Stability Program, hereafter designated as SV. The value of SV also represents the maximum rate employed in DRU testing with the bench-scale reactor.

Table 1. Typical Analysis of Bitumen Used in DRU Optimization Tests

	With Diluent	Without Diluent
Elemental, wt%		
C	84.05	
H	10.44	
N	0.27	
S	4.27	4.5
Water, wt%	0.372	3.675
PI, wt%	14.1	17.52
HI, wt%	9.89	
TI, wt%	0.01	0.03
MCR, wt%	11.85	13.24
BSW, wt%	0.2	5.3
Pour Point, °C	-15	18
Density (API)	0.9624 (15.39)	0.997 (10.43)
Viscosity, cSt		
60, °C	214.16	2099
80, °C	84.66	502.7
100, °C	41.06	173.5
SARA, wt%		
Asphaltenes (C_5)		
Saturates		17.52
Aromatics		20.60
Polars		47.31
		14.57
Metals, ppm		
Ni		58.4
V		155.8

First Test Series: Test Effects of Sweep Gas and Space Velocity on Product Recovery

Tests in this series used undiluted bitumen as feed material and explored the effects of sweep gas composition and space velocity on product recovery. Initial tests repeated the temperature and high space velocity conditions (1.0*SV) of the previously conducted Stability Program. Tests conducted under the Stability Program used nitrogen as a sweep gas, however current testing used CO_2 or CH_4 (either of which would be available at an actual upgrading facility) to determine the effects of these species on overhead recovery. The test results and replicates are summarized in Tables 2 and 3 for CO_2 and CH_4, respectively.

Table 2. CO_2 Sweep Gas Tests at High Space Velocity (1.0*SV).

Test 1a	Stage 1	Stage 2	Stage 3	Stage 4	Stage 5	Closure, %
Temperature,°F	338	471	612	661	704	
Cumulative Yield,%	3.17	3.48	11.25	19.41	33.95	99.88
Test 1b	**Stage 1**	**Stage 2**	**Stage 3**	**Stage 4**	**Stage 5**	**Closure, %**
Temperature,°F	338	473	612	662	706	
Cumulative Yield,%	2.49	3.11	11.23	19.16	34.44	97.55

Table 3. CH_4 Sweep Gas Tests at High Space Velocity (1.0*SV)

Test 3a	Stage 1	Stage 2	Stage 3	Stage 4	Stage 5	Closure, %
Temperature,°F	333	469	607	657	703	
Cumulative Yield,%	2.13	2.38	11.57	19.08	34.01	98.77
Test 3b	**Stage 1**	**Stage 2**	**Stage 3**	**Stage 4**	**Stage 5**	**Closure, %**
Temperature,°F	342	466	607	657	704	
Cumulative Yield,%	1.91	2.92	11.74	18.48	32.93	98.07

The small differences in yield between the high space velocity tests are statistically within normal experimental variation and therefore do not reflect a conclusive dependence on sweep gas composition. Figure 4 shows overhead recovery for CO_2 and CH_4 sweep gas with process temperature and compares results with a reference test conducted under the Stability Program. The bitumen's normal boiling (NBP) curve, as determined by simulation, is also shown. These results demonstrate an excellent correspondence with the reference test and continue to indicate no effect of sweep gas composition on overhead production.

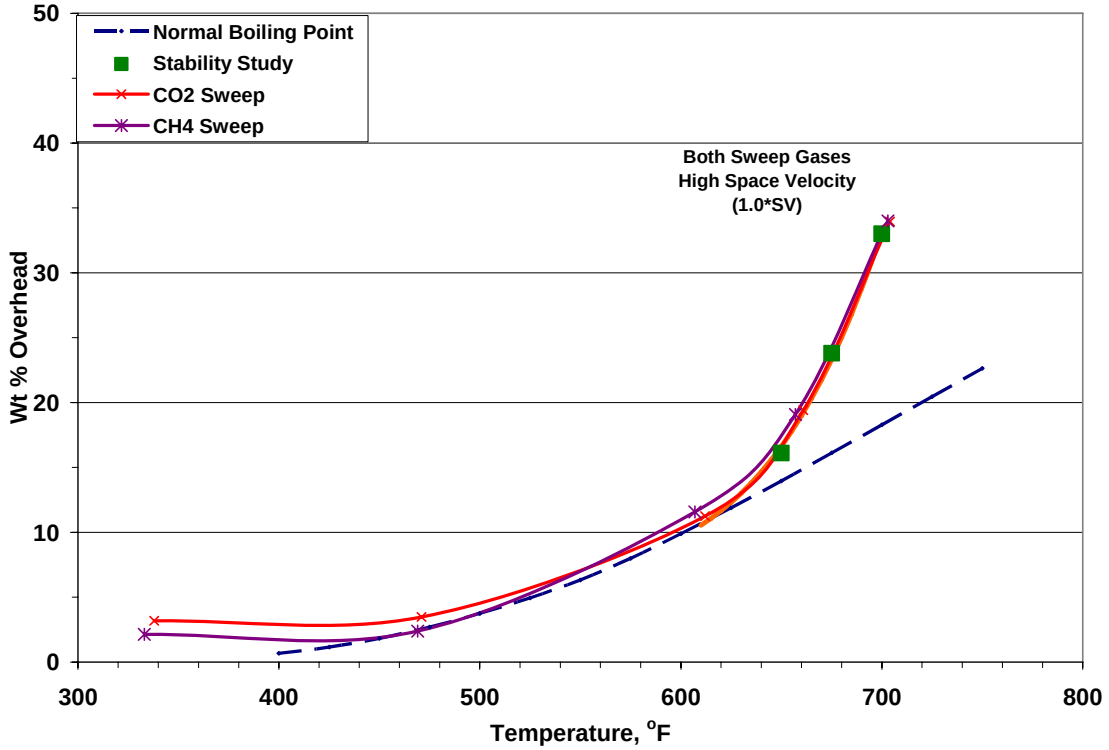


Figure 4. Comparison of overhead recovery when using CO_2, CH_4 and N_2 sweep gas.

Conducting additional tests at lowered space velocity (0.6*SV) yielded results summarized in Tables 4 and 5 for CO_2 and CH_4 sweep gas, respectively.

Table 4. CO_2 Sweep Gas Tests at Lowered Space Velocity (0.6*SV)

Test 2a	**Stage 1**	**Stage 2**	**Stage 3**	**Stage 4**	**Stage 5**	**Closure, %**
Temperature,°F	335	467	621	663	697	
Cumulative Yield,%	1.67	3.39	15.61	25.15	38.87	101.16
Test 2b	**Stage 1**	**Stage 2**	**Stage 3**	**Stage 4**	**Stage 5**	**Closure, %**
Temperature,°F	329	464	622	660	697	
Cumulative Yield,%	2.6	3.08	17.45	26.95	41.43	101.34

Table 5. CH_4 Sweep Gas Tests at Lowered Space Velocity (0.6*SV)

Test 4a	Stage 1	Stage 2	Stage 3	Stage 4	Stage 5	Closure, %
Temperature, °F	337	468	620	658	701	
Cumulative Yield,%	2.12	2.21	17.07	26.24	48.31	97.28
Test 4b	**Stage 1**	**Stage 2**	**Stage 3**	**Stage 4**	**Stage 5**	**Closure, %**
Temperature, °F	337	468	620	658	701	
Cumulative Yield,%	2.68	3.04	17.72	26.82	46.12	97.69

Overhead production at lowered space velocity exhibited somewhat more scatter than at high space velocity but continued to show no conclusive correlation to sweep gas composition. However, the results show a strong dependence of lowered space velocity on increased overhead recovery as seen in Figure 5 that compares averaged recovery curves for the high and lowered space velocity tests.

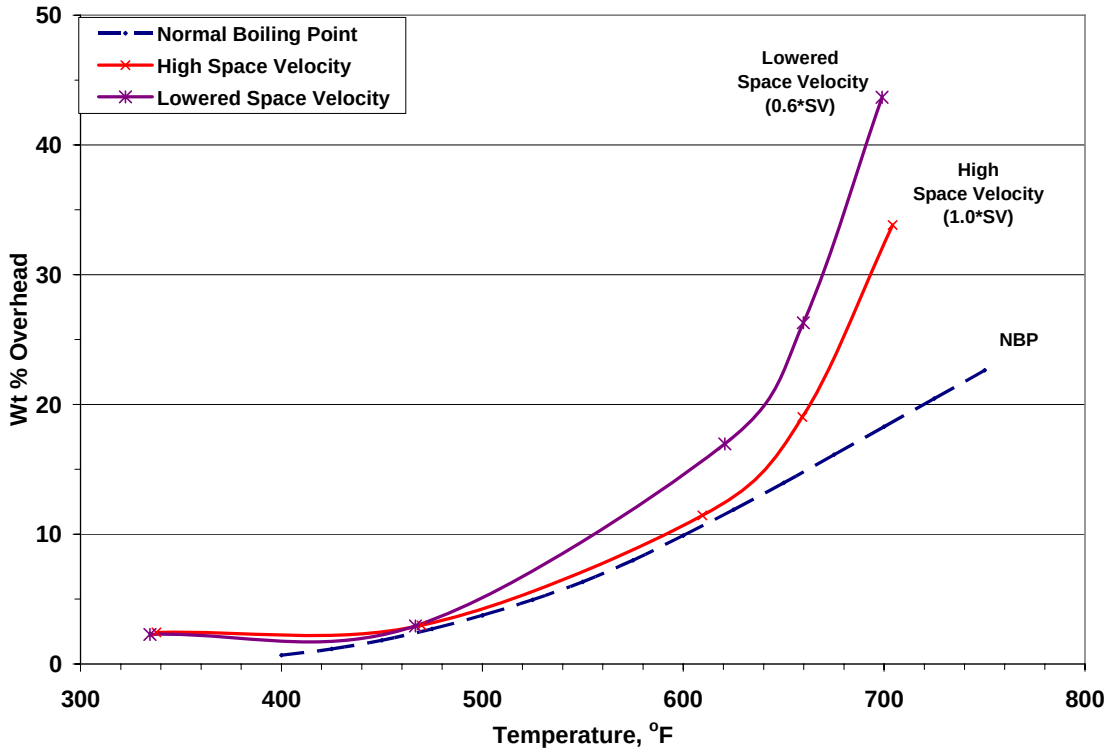


Figure 5. Averaged overhead production curves using CO_2 and CH_4 sweep gas at high and lowered space velocities.

Samples from overhead and bottoms streams were submitted for chemical analyses and the results are summarized in Tables 6 and 7. The differences in properties for the overhead material produced using CH_4 or CO_2 sweep and high or reduced space velocity are small to insignificant (Table 6).

Table 6. DRU Overhead Analyses as Function of Sweep Gas and Space Velocity

Property	High Space Velocity, 1.0*SV		Lower Space Velocity, 0.6*SV	
	CO_2	CH_4	CO_2	CH_4
Carbon, wt%	84.68	84.50	84.38	84.41
Hydrogen, wt%	12.80	12.84	12.91	13.04
Nitrogen, wt%	0.12	0.15	0.12	0.11
Sulfur, wt%	2.75	2.94	2.82	2.81
Diene Value, g I_2/100 g	1.70	1.64	1.49	1.91
Pour Point, °C	-60	-33	-48	-54
Density, °API	24.34	24.07	24.76	24.15
Viscosity @ 20°C, cSt	18.311	17.012	14.453	17.792
P Value	3.1		4.62	

Conversely, the quality of the DRU's bottoms streams negatively correlate to increased residence time as shown in Table 7. Bottoms produced at the lowered space velocity are denser and more viscous than those produced at the higher space velocity. The bottoms also exhibit higher P values (measure of tendency to coke), indicating a greater degree of thermal degradation. Also consistent with their thermal history, bottoms produced at lower space velocity are lower in aromatics and resins but have higher asphaltene concentrations. All of these results are consistent with a higher severity of cracking reactions necessary to produce an increased quantity of light, overhead product.

Analyses of the runs conducted in this test series shows that overall DRU overhead yield depends on stage temperature and space velocity, not sweep gas composition. Future test series used CH_4 as sweep gas because it is readily available at field production facilities and would most probably be used in WRITE™ processing.™

Although overhead production does not depend on gas composition, a preliminary analysis of the product's naphtha fraction (Table 8) might suggest an increase in naphtha at both space velocities when using CO_2 as a sweep gas rather than CH_4. Further testing would be needed to confirm this observation.

Table 7. DRU Bottoms Analyses as Function of Sweep Gas and Space Velocity

Property	High Space Velocity, 1.0*SV		Lower Space Velocity, 0.6*SV	
	CO_2	CH_4	CO_2	CH_4
Carbon, wt%	85.58	84.80	85.26	84.64
Hydrogen, wt%	9.08	9.24	9.24	8.22
Nitrogen, wt%	0.96	0.90	0.59	1.07
Sulfur, wt%	5.17	5.29	5.18	5.28
Diene Value, g I_2/100 g	17.56	13.2	16.29	10.77
Density, °API	5.08	4.69	3.65	1.61
Viscosity @ 100°C, cSt	698	752	1190	1010
P Value	1.82	2.06	1.38	1.12
SARA, wt%				
Saturates	11.93	12.63	12.10	12.20
Aromatics	39.94	39.94	38.28	35.84
Resins	24.80	25.37	23.16	20.25
C_5 Asphaltenes, wt%	23.33	22.06	26.46	31.71
Metals, ppm				
Al	12.4	9.0	10.5	12.8
Ba	2.7	1.6	1.9	2.3
K	5.3	<0.8	1.9	2.3
Ca	68.3	36.9	42.8	55.8
Fe	18.6	7.4	10.5	14.0
Mg	10.6	5.7	7.6	9.3
Mn	<0.9	<0.8	<1.0	<1.2
Na	48.8	33.6	38.1	52.3
Ni	108.2	105.0	118.1	131.4
Si	18.6	10.7	18.1	18.6
V	287.4	283.0	317.1	353.6

Table 8. Properties of the DRU Overhead Naphtha Fractions as Function of Sweep Gas and Space Velocities

Property	High Space Velocity		Lower Space Velocity	
	CO_2	CH_4	CO_2	CH_4
Yield, wt%	8.5	8.0	8.8	7.3
Composition, vol%				
Aromatics	16.0	14.1	19.7	15.5
Olefins	29.0	33.3	26.7	29.5
Saturates	55.0	52.6	53.7	55.0
Bromine Number, g Br_2/100 g	46.2	50.6	48.9	50.6

An additional test was conducted with undiluted bitumen at a space velocity of 0.3*SV but at a lowered stage 5 temperature of 675ºF. Test C was actually conducted under Series 2 tests (Table 9) but included here to compare results with other undiluted bitumens. Plotting results from Test C along with the averaged recovery curves for space velocities of 0.6*SV and 1.0*SV (Figure 5) continue to show that reduced space velocity yields increased overhead production, even at a lowered top stage temperature.

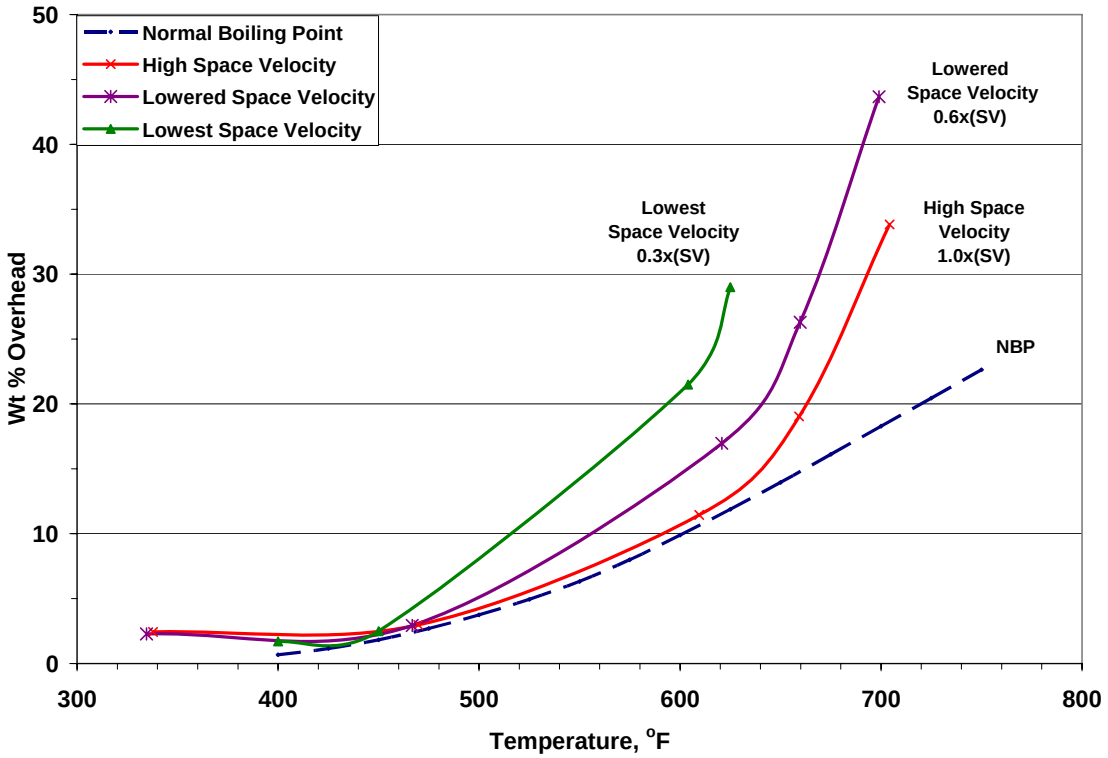


Figure 6. Comparison of overhead production curves for undiluted bitumen produced at space velocities of 0.3*SV, 0.6*SV, and 1.0*SV

Second Series: Test Increased Processing Severity Using Undiluted and Diluted Bitumen

This series of tests was run with diluted and a single undiluted bitumen from EnCanna's Foster Creek operations (Table 1.) Simulated distillations were performed on both crudes and the boiling distribution is shown in Figure 7. As expected, the dilbit has an initial boiling point 300ºF lower than the undiluted bitumen. The curves become generally parallel at temperatures above 500ºF indicating that the diluent has been removed above this temperature. The dilbit reportedly contained a nominal 20wt% condensate, however if the normal boiling point curve for the dilbit is reduced by 9wt%, the boiling point curve of the undiluted bitumen is reasonably approximated (Figure 7).

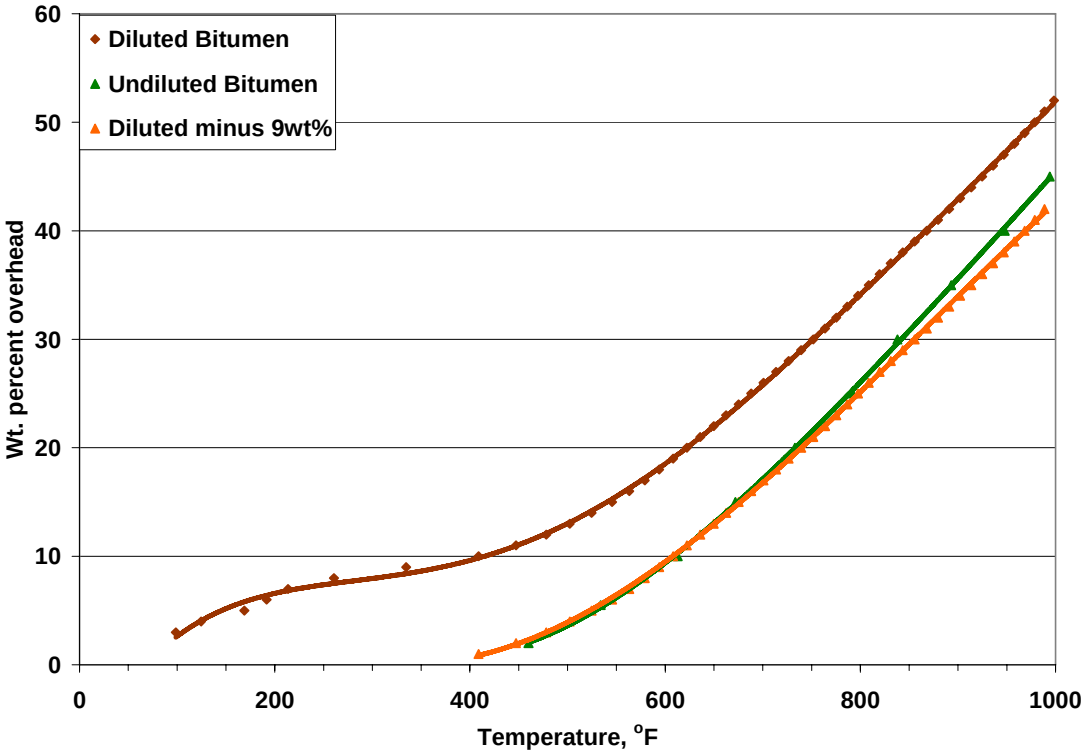


Figure 7. Comparison of simulated distillations from diluted and undiluted bitumens.

Runs were conducted at stage 5 temperatures of 650ºF and 675ºF using both diluted and undiluted feeds. Operating conditions and material balances for three representative tests in this series is summarized in Table 9. It should be noted that process severity based on the combination of temperature and space velocity for tests B2 and C were more severe than for the previous test series that used undiluted bitumen. Representative composite overhead samples accumulated from all reactor stage knock-out pots were sent for analysis. The crude assays from these samples are shown in Figure 8.

Table 9. Mass Balance for Tests Using Diluted and Undiluted Feed

Run	Diluent Y/N	Temp. °F	Feed lbs	Overhead Collected		Bottoms lbs	Loss /Closure Lbs/%	Space Vel. (SV)
				KO 1&2	KO 3-5			
A1	Y	650	70.05	4.89	17.18	45.43	2.55/96.4	0.50
B2	Y	675	36.80	1.65	18.72	10.43	6.00/83.7	0.29
C	N	675	56.44	2.94	22.57	27.03	3.90/93.1	0.30

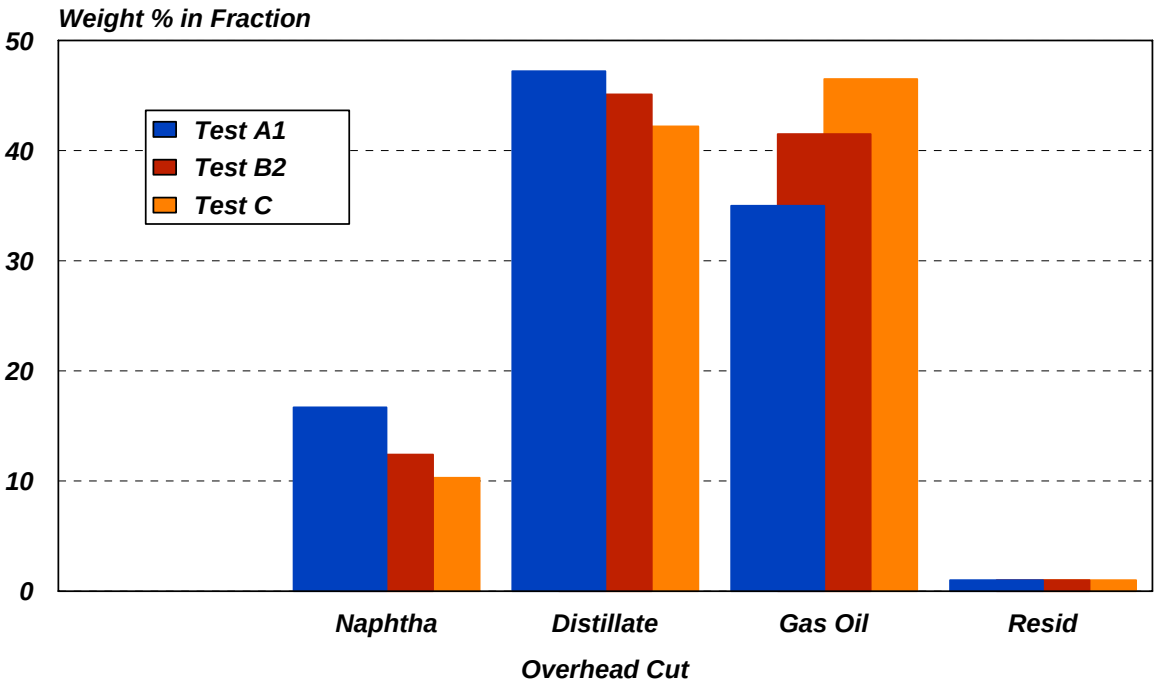


Figure 8. Crude assays of overhead produced from Series 2 Tests A1, B2, and C.

Material balance closures here were not as good as for tests that used undiluted bitumen. Part of the problem lies with the bench-scale DRU's product collection system that was not designed to trap the fraction of diluent existing at room temperature and 12 psia. However, Test C used undiluted feed, and its material balance closure was also less than expected. Figure 8 shows that overhead from Test C is heaviest among the three products reported, suggesting a higher degree of thermal decomposition. So it's possible that Test C also produced a light boiling fraction not trapped by the reactor's knock-outs. Test B2 differed from Test C only in its use of dilbit. Its generally lighter composition, relatively greater amounts of naphtha and lesser amounts of gas oil (compared to Test C), may be consistent with some of the diluent being collected along with the lightest product oils, but also exhibited the lowest recovery of low boiling materials and poorest material balance closure. Because Test A1 operated under the least severe conditions, it produced the least amount of overhead that might contain the greatest

proportion of diluent. Interpretations of the crude assays should bear these considerations in mind.

Tables 11 and 12 summarize the production by stages for Tests A1 and B2, respectively. Plots of as-measured overhead production from Tests A1 and B2 as a function of temperature fell below the NBP curve, which is unexpected given results from all previous tests that consistently showed overhead recoveries higher than those expected by distillation. This and the fact that the material balances for these two tests were deficient by 2.55 and 6.00 lbs, respectively, caused the postulation that unaccounted for material was non-condensable or lost diluent. Given a proper recovery system the material would have been recovered in knock-out pots 1 and 2, where stage 2 temperature was approximately 470ºF. Accordingly, Tables 10 and 11 show both the as-measured overhead by stages, and account for the un-recovered material as diluent that would have been recovered on a 50:50 weight basis in knock-outs 1 and 2. Figure 9 shows the adjusted overhead production curve for Tests A1 and B2 as well as the dilbit's NBP curve. The fact that the overhead production with "diluent added" fall on or above the NBP curve suggests this method of accounting for lost material represents a plausible explanation.

Table 10. Production by Stage for Test A1

	KO #1	KO #2	KO #3	KO #4	KO #5	Total
Total	4.09	0.80	7.88	3.62	5.68	22.07
Percentage	5.84	1.14	11.25	5.17	8.11	31.51
Cumulative	5.84	6.98	18.23	23.40	31.51	
Adding "Lost Diluent"						
Total	5.365	2.075	7.88	3.62	5.68	24.62
Percentage	7.66	2.96	11.25	5.17	8.11	35.15
Cumulative	7.66	10.62	21.87	27.04	35.15	

Table 11. Production by Stage for Test B2

	KO #1	KO #2	KO #3	KO #4	KO #5	Total
Total	1.65	0.00	9.60	2.75	6.37	20.37
Percentage	4.48	0.00	26.09	7.47	17.31	55.35
Cumulative	4.48	4.48	30.57	38.04	55.35	
Adding "Lost Diluent"						
Total	4.65	3.00	9.60	2.75	6.37	26.37
Percentage	12.64	8.15	26.09	7.47	17.31	71.66
Cumulative	12.64	20.79	46.88	54.35	71.66	

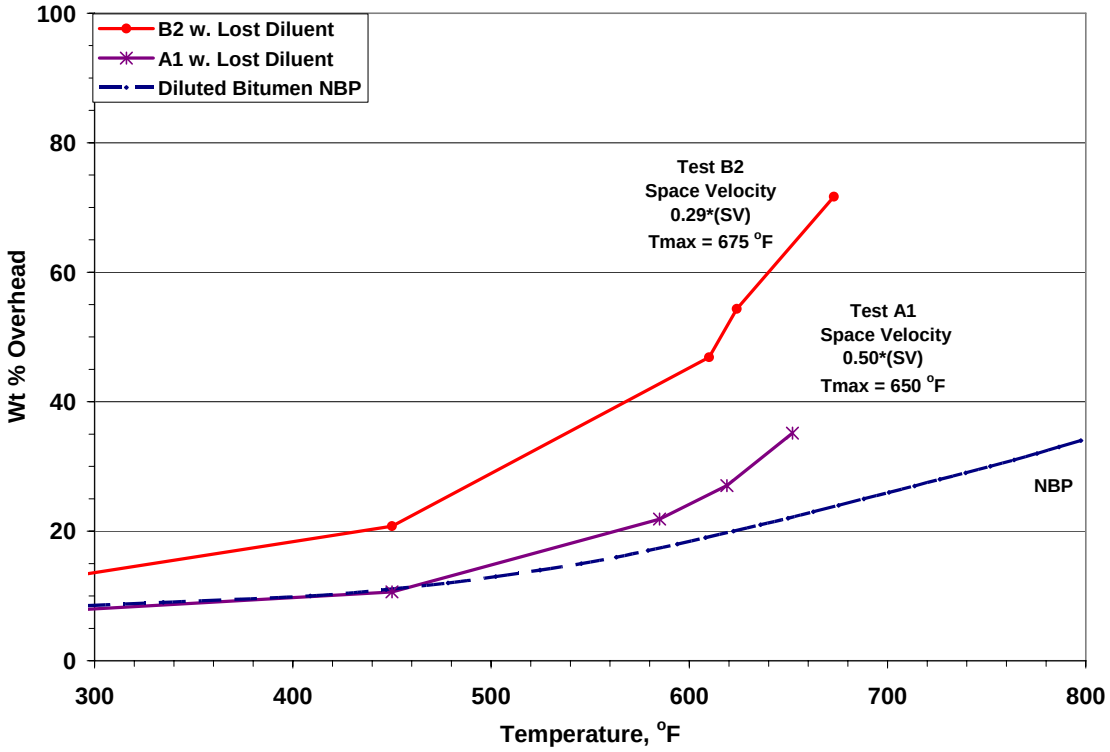


Figure 9. Overhead production for Tests A1 & B1 with "Lost Diluent" added.

For comparison to Test C, overhead production for Tests A1 and B2 were corrected to a diluent free basis using the following estimation procedure. The calculations assumed that material recovered from stages 1 and 2 was comprised mostly of diluent, which could be subtracted from the distribution of overhead product, except for the quantity that resulted from volatilization of the feed bitumen at temperatures below approximately 450 to 470ºF. This quantity was estimated by averaging fractional recovery from knock-outs 1 and 2 in tests using undiluted bitumen. For most tests, the averaged quantity of material recovered reasonably approximated that determined from the normal boiling curve. Overhead recovered from stages 3-to-5 reflected thermal processing of the bitumen so was included in the distribution of overhead. This method resulted in the overhead production curves plotted in Figure 10. As with tests that use undiluted bitumen, these results continue to show that increased severity (measured by increased temperature and reduced space velocity) correlates directly to amount of overhead recovered.

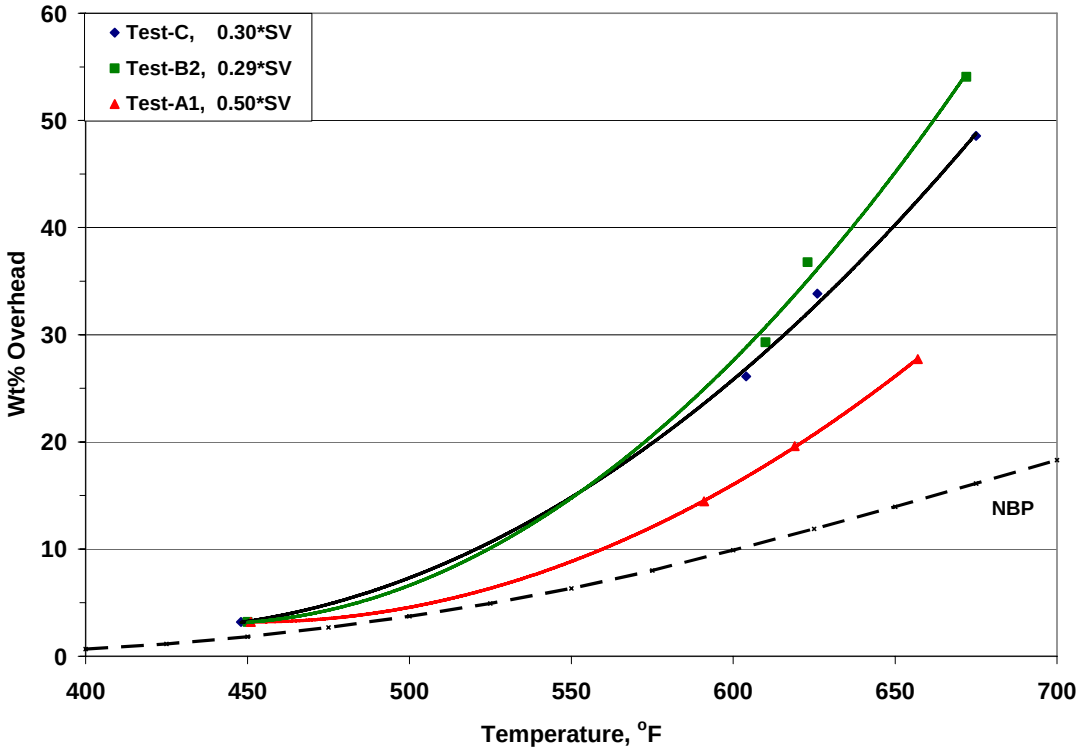


Figure 10. Overhead production from Tests A1, B2, & C on a diluent-free basis.

Third Series: Test at Lower Space Velocities Using Diluted Bitumen

Tests conducted in this series explored overhead recoveries at space velocities and temperatures lower than previously investigated. We ran five tests with space velocities that varied nominally from 18% to 50% of that used in the Stability Study. Before testing on this series commenced, the bench-scale DRU was refurbished as described below.

A peristaltic feed pump was installed into the system and subsequent calibrations verified its ability to deliver dilbit at reduced feed rates needed for the low space velocity test conditions with less than a 1wt% variation in delivery rate.

All five stages were drained and flushed with diesel to ensure residual material was removed. Load cells used to measure feed rate, production rate, and overhead production from individual reactor stages, were recalibrated. Stages one through five, their associated vapor recovery systems, and the shared gas collection system were tested for leaks. A major leak was found and corrected in stage three's vapor recovery plumbing. A leak was also found and corrected in a gas recovery line leading to the gas bubblers, downstream of the knock-outs. These leaks may have contributed to the material balance losses that were noted in Series Two's tests. With the leaks repaired, the system was pressurized to approximately five psig and allowed to remain for several hours to verify system integrity.

In the process of inspecting the interior of reactor stage five, we discovered that coke had completely encased its heating elements and appeared to cover the lower half of the horizontal tank. Stage five had to be cut in half to gain access to the heater element for manual dislodgment of the accumulated coke. The manual cleaning resulted in complete removal of the coke without damaging the heater. As part of the reassembly process, two additional thermocouple ports were added to stage five to allow the measurement of this reactor stage's oil temperature. One thermocouple was located at the mid-point of the reactor (measured lengthwise) and the other approximately three inches from its discharge point. Both were positioned so as to remain completely immersed in oil during testing. Deployment of these thermocouples provided a more precise indication of processing conditions in the DRU's highest temperature reactor stage.

Five tests were conducted with space velocities that varied from 0.18*SV to 0.48*SV. Material balances mostly improved over the previous series. Tests with space velocities from 0.30-to-0.42*SV reported recoveries of 98% or higher, but tests at 0.18*SV and 0.48*SV had recoveries of 94 and 93%, respectively. All tests evidenced a progression of increasing overhead yields with increased severity (i.e. decreased space velocities and increased temperatures), consistent with previous tests. The production curves for the five tests in this series are plotted (on a diluent free basis) in Figure 11.

The production response for the test conducted at 0.48*SV is anomalous because its recovery falls below or on the NBP curve. It is possible that non-condensable vapor escaped stage 1 and 2 knock-outs as was postulated in the second series of tests. However, this test was conducted at the highest space velocity (and probably lowest overall severity) for the series, so that argument is less persuasive. Another possible explanation is that the low relative reactor temperatures coupled with the relatively high liquid rates resulted in a less than optimum overall heat transfer to the oil. This could reduce kinetic rates to the point were cracking reactions were negligible.

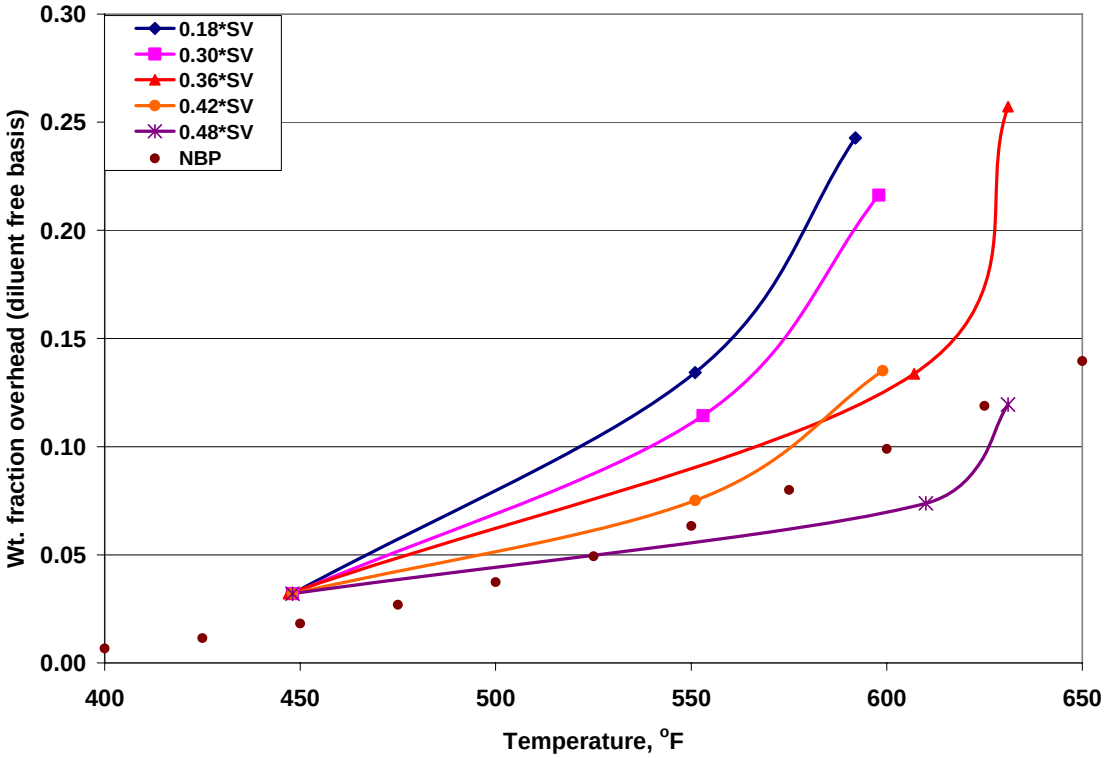


Figure 11. Overhead production from Series 3 tests at space velocities 0.18 to 0.48*SV.

Consistent with earlier observations, DRU distillate overhead quality measured by API gravity and kinematic viscosity is not a strong function of processing conditions (Table 12). For this series of tests, little difference exists in crude assays of the boiling point fractions or among saturate, aromatic and olefin concentrations in various boiling point cuts (Figures 12 through 14, respectively).

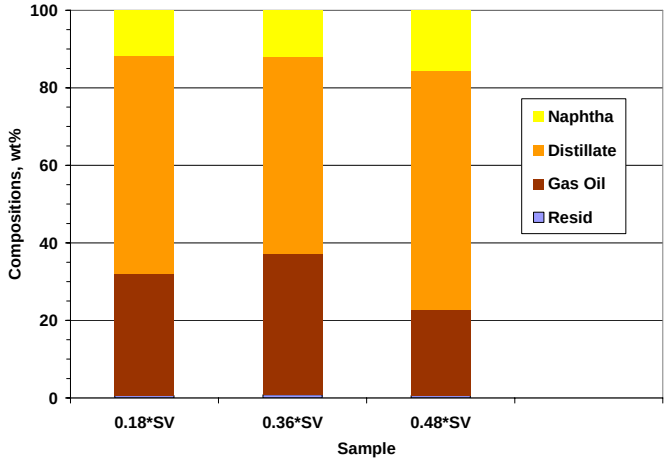


Figure 12. Crude assays of overhead produced from Series 3 tests at 0.18, 0.36, &0.48*SV.

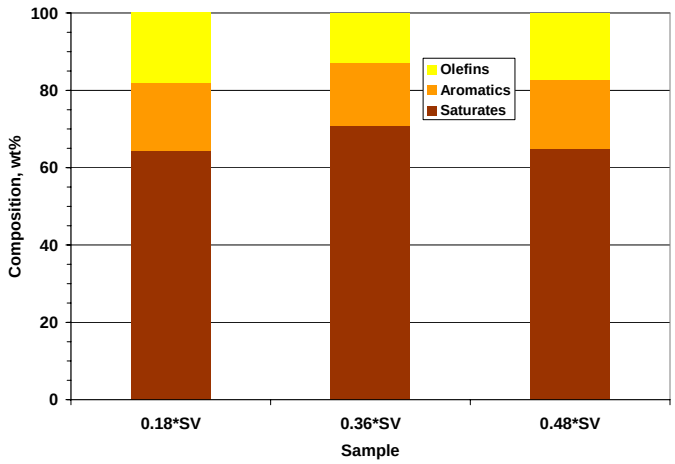


Figure 13. Assay of naphtha fraction from overheads produced at 0.18, 0.36, &0.48*SV.

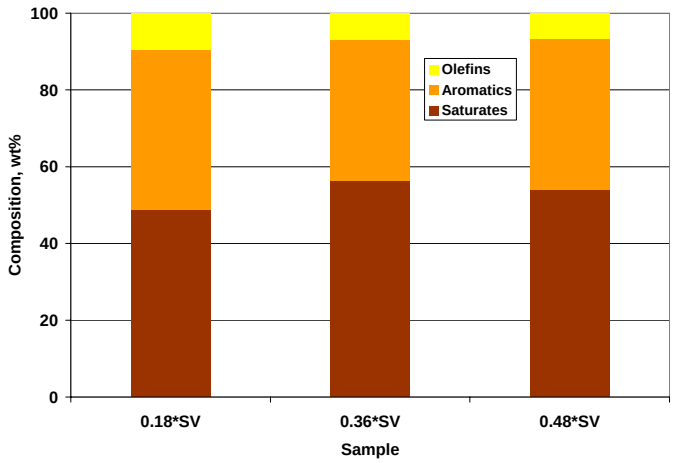


Figure 14. Assay of distillate fraction from overheads produced at 0.18, 0.36, &0.48*SV.

Table 12. Properties of Products from Series Three Tests

Properties of the DRU Overheads				
Analysis	**0.18*(SV)**	**0.30*(SV)**	**0.36*(SV)**	**0.48*(SV)**
API Gravity, deg	27.25	27.93	27.98	29.21
Pentane Insoluble, wt%	.005	.015	.02	0
Toluene Insoluble, wt%	0	.01	0	0
Pour Point, °C	-48	<-80	-60	<-80
TAN, mg KOH/g oil	0.44	0.64	0.469	0.538
Kinematic Viscosity, cSt				
20 °C	8.817	8.239	7.066	6.519
40 °C	4.818	4.566	4.475	3.843
50 °C	3.783	3.618	3.597	3.049
Properties of the DRU Bottoms				
Analysis	**0.18*(SV)**	**0.30*(SV)**	**0.36*(SV)**	**0.48*(SV)**
Density, g/cc	1.086	1.132	1.134	1.134
API Gravity, deg	-1.21	-6.50	-6.72	-6.72
I_N	41.6	43.65	49.0	42.7
S_{BN}	98.9	126.6	128.4	120.4
P-Value	2.38	2.90	2.62	2.82

Extended Production Tests

MEG's consultants recommend additional testing to define, over a more specific range, the DRU's operating envelope and to demonstrate its long term performance characteristics. This data was necessary to complete the reference design for the DRU. The work was accomplished in two test sequences, both completely funded by MEG.

Before testing began, additional modifications were made to the bench-scale DRU. As was done earlier in reactor stage 5, a thermocouple was placed at the midpoint of both reactor stages 3 and 4 to measure respective oil bath temperatures. This change allowed a more detailed characterization of average oil bath temperature in the DRU's high-temperature stages where chemical reactions will occur. In addition, thermocouples were placed in the vapor space of knock-out pots 1-through-5 to measure their respective condensation temperatures. These temperatures, with the use of vapor-liquid equilibrium calculations, helped quantify the amount of hydrocarbon vapor that escaped the DRU's recovery system.

Survey and extended production test sequences were conducted to acquire specific process information requested by MEG. The survey tests comprised a set of relatively short duration runs, conducted at increasing top stage temperatures and a specified space velocity, with the goal of finding the condition that maximized overhead yield. Production testing was conducted at the optimum condition found in the sequence of survey tests. The objective was to operate continuously for 10 days to ensure that the DRU had achieved steady and stable conditions. Characterizations of the production streams and DRU operating parameters would then reflect the performance of a continuous flow reactor operating at steady state conditions.

The runs in this test series were successful and provided the needed information. Yields of overhead from the production test were consistent with those found in the previous test series and the Stability Program. The material balance for the production test was 98wt%.

Summary of Results from DRU Testing

The range of reactor temperature and residence time conditions for all tests conducted in the bench-scale DRU are summarized in Figure 15. The overhead recoveries shown in the plot are expressed on a diluent free basis. The curves shown as dashed lines represent the minimum and maximum space velocities explored. Taken collectively, these data define the processing envelope for the bench-scale DRU that will be of use for scaling to a larger size reactor.

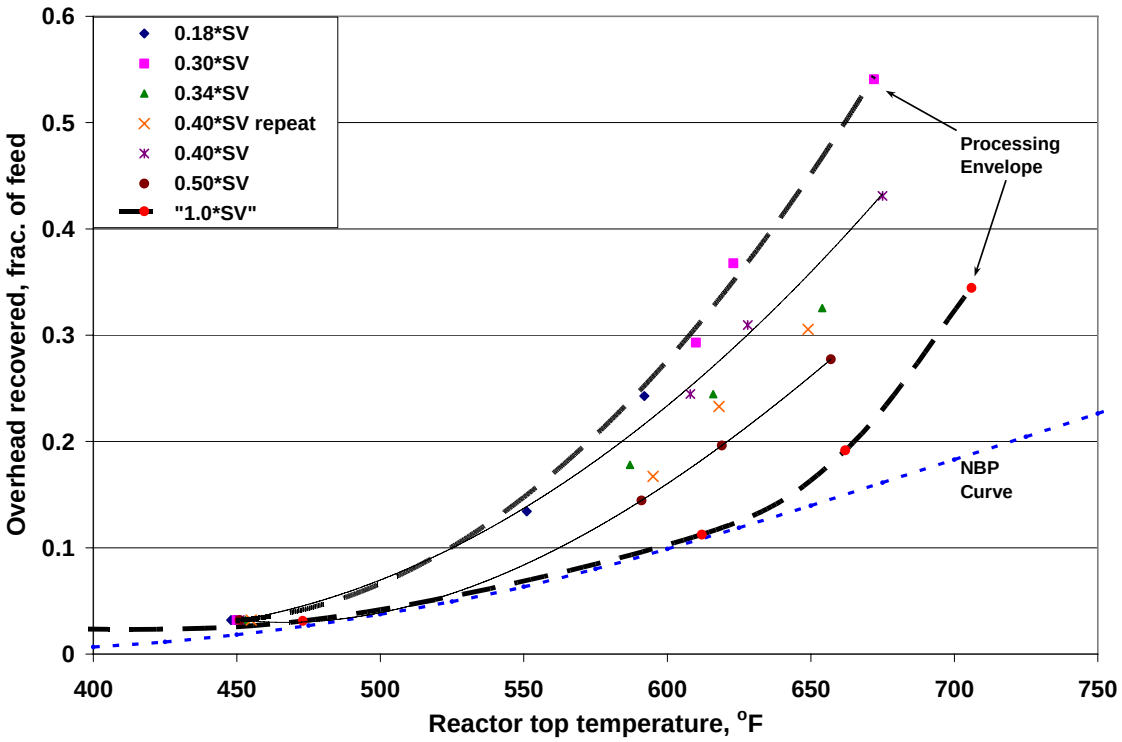


Figure 15. Processing envelope determined from DRU testing.

We developed kinetic relationships for overall DRU conversion based on oil temperature data (from tests for which that data was available). Kinetic relationships derived from three top stage oil temperatures ranging from least to most severe and corresponding overhead production shows the reactions are first order, as expected for oil cracking (Figure 16). The Arrhenius fit to the data is also shown in Figure 16. The activation energy is approximately 59kcal/mole, which is within the range of values determined by Hayashitani (1978) in his study that determined cracking kinetics of Athabasca oil sands oil.

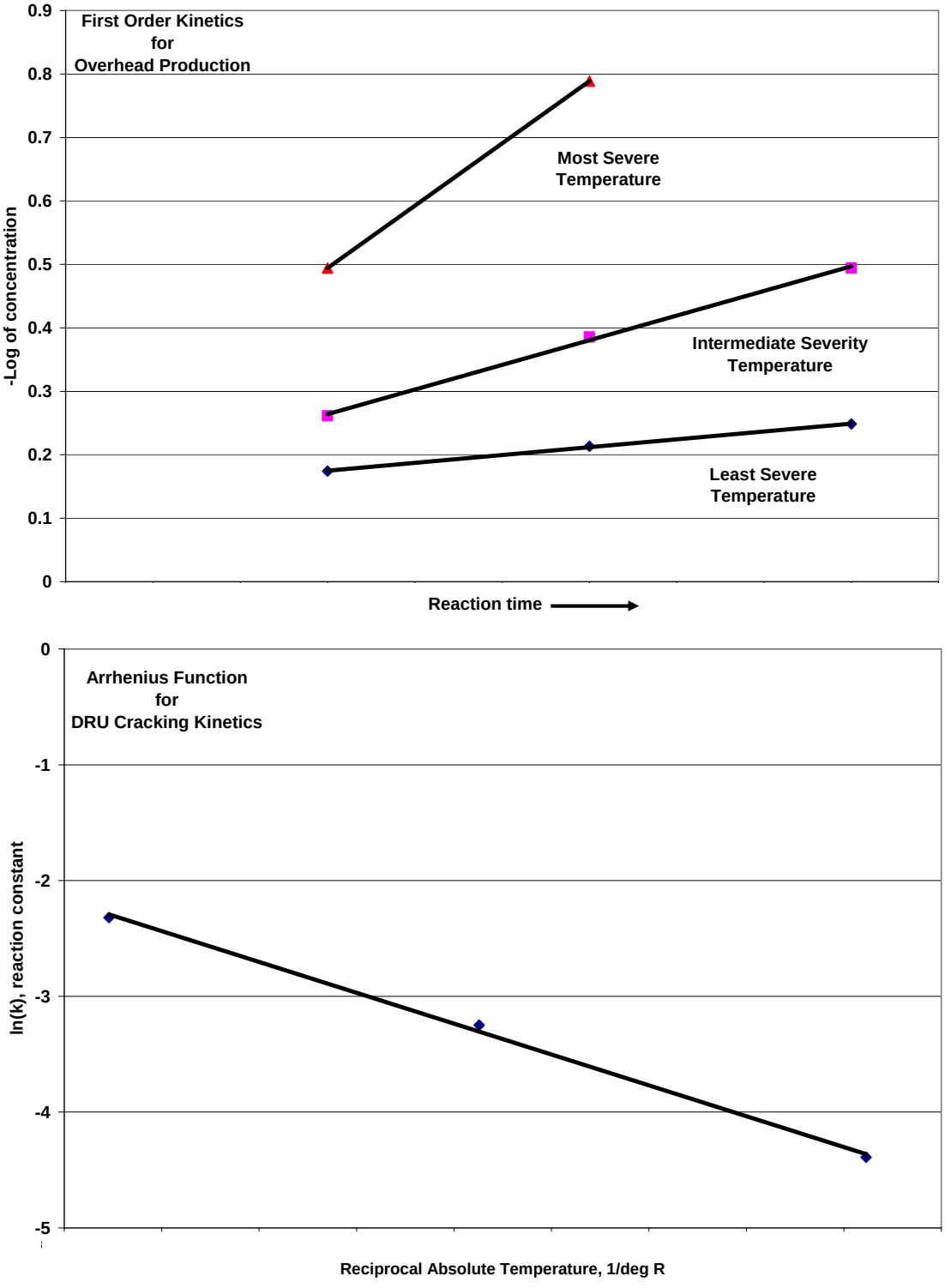


Figure 16. First order kinetics and Arrhenius relationship developed from DRU tests.

We determined the following from DRU testing:

1. The overhead oil produced by the DRU meets pipeline specifications for density and viscosity. The API gravity and kinematic viscosity of the overhead product show little variation for the wide range of processing conditions used in this study. Oil density and viscosity varied from 27 to 30 ºAPI and 6 to 9 cSt at 68ºF, respectively. The DRU overhead product easily conforms to Canadian pipeline specifications of 19º API gravity or lighter and a viscosity at pipeline temperatures of 350 cSt or less.

2. The observed yields of overhead from the DRU ranged from 30 to 50wt%. This is significantly higher than the 20wt% indicated by boiling considerations alone.

3. The DRU functions as a mild thermal cracking reactor with performance described by first-order kinetics, which accounts for its increase recovery compared to distillation. Overhead yields generally increase with increasing temperature and decreasing space velocity (i.e. increased residence time).

4. The yields of distillate product from the DRU exhibited very little, if any dependence on sweep gas composition for the N_2, CO_2, and CH_4 gases employed in testing.

5. The range of compositions for the DRU overhead oil are 10-15wt% naphtha (boiling lower than 392ºF), 50 to 60 wt% distillate (boiling between 392 and 662ºF), 25-30wt% gas oil (boiling between 662 and 914ºF), and less than 1wt% resid (boiling higher than 914ºF). In general the naphtha fraction tended to slightly increase with increasing processing severity and the distillate tended to decrease, which is consistent for thermal cracking reactions. Consistent with kinetics, processing severity increases with increasing temperature and decreasing space velocity or both.

6. The naphtha fraction of the product oil contains 10-20wt% olefins, 15-20wt% aromatics, and 65-70wt% saturates. The distillate fraction contains 5-10wt% olefins, 35-40wt% aromatics, and 50-55wt% saturates. The olefins, particularly in the naphtha fraction, will require hydrotreating to meet pipeline specifications for bromine number.

7. Bottoms from the DRU become heavier and more viscous as processing severity increased. These results were consistent for thermal cracking of heavy oils that result in concentrating the ultra-heavy asphaltenes in the non-distillable residual fraction.

CONTINUOUS COKER TESTS

The continuous coker is the second, high severity stage of the WRITE™ process. The bottoms produced from the DRU are a heavy, viscous co-product containing high concentrations of metals and other heteroatoms such as nitrogen and sulfur. As a result, the bottoms have limited options for use. Typically this material is processed by either (1) direct combustion as a fuel or (2) severe temperature pyrolysis. Applications for direct fuel combustion are limited by the viscous nature of the bottoms and maintaining a reasonable viscosity for pumping and storage. A reasonable viscosity requires 350ºF, which is costly and adversely affects the economics of the process.

The second alternative is to process the bottoms under more severe conditions to produce additional overhead product and coke. The coke can then be used as a fuel and the overhead blended with overhead from the DRU and sent to the pipe line as SCO. The severe processing of the bottoms represents coking conditions that can be performed in equipment such as a delayed coking unit or WRI's continuous coker.

The continuous coker has been demonstrated in earlier studies as a viable alternative to delayed coking. However, additional data is required to define the coker's operating envelope and allow scaling the design to commercial applications. This Task's objectives were to obtain data to determine (1) the rate of bottoms throughput for the continuous coker as a function of temperature, (2) the product slate, and (3) the extent of volatile material removal from the coke. The feedstock for these tests was bottoms produced from Cold Lake bitumen that was processed in the bench-scale DRU.

The study was conducted using an inclined 6-inch, twin screw pyrolyzer initially developed for use as the high temperature stage of the Recycle Oil Pyrolysis and Extraction (ROPE) process (Cha et. al. 1987). ROPE co-processed hydrocarbon bearing materials such as tar sands and oil shale with waste oil to produce a light distillate overhead. The twin screw pyrolyzer is shown schematically in Figure 3 and in a photograph in Figure 17. The only modification made to the unit was the addition of a level control loop in the feed system, which maintained a constant liquid level of bottoms in the continuous coker.

The continuous coker was heated by external heaters that for this series of tests were divided into two zones. The first zone, located at the lowest third of the reactor, and referred to as the bottoms pyrolysis zone, heated the pool of bottoms material in the continuous coker. It is believed to be the region where the majority of the coking takes place. The second or drying zone encompassed the remainder of the screw length and was used to volatilize the final traces of liquid material from the coke.

The first series of tests were designed to establish the maximum throughput of bottoms in the continuous coker. This information is required to define the size of a continuous coker needed per quantity of feed. The higher the throughput that can be achieved, the smaller the size

coker required for commercial-scale processing, which relates to lower capital and operating costs.

Figure 18 summarizes the results from the throughput studies. We define throughput as the feed rate in pounds per hour required to maintain the liquid level in the bottoms pyrolysis zone. Examination of these results shows the expected response of feed rate increasing as a function of increased temperature. The results provided in Figure 18 can be fit with an exponential function. During these experiments, it was not possible to increase the temperature of the bottoms pyrolysis zone above 875ºF because the reactions became too violent above and the reactor's performance unstable.

The cause of the instability of the reactor was a shortcoming in its design. The six-inch pyrolyzer was designed as a second, high-temperature processor fed directly with solid material from a first stage. To accommodate the solid feed, an eight inch length of 10-inch diameter pipe was welded on the top of the screw barrel at the bottoms-feed end of the unit. This pipe was fitted with a flange, which we blanked off so the unit could operate as a stand alone continuous coker. During high temperature operation, the dead volume in the pipe filled with produced vapor, and when the pressure in this volume exceeded system pressure, the vapor would in-rush into the bottoms pool. This in-rush of vapor entrained the bottoms, which carried over into the overhead collection system. In addition, the vapor movement forced bottoms up into zone 2, where the temperature was significantly higher. With the bottoms at high temperature, they underwent rapid pyrolysis. The additional pyrolysis in zone 2 increased the vapor production rate, contributing to additional entrainment of bottoms into the overhead collection system. At the lower temperatures studied, the vapor production rate was significantly lower, which was accommodated by the unit. The problems with entrainment and associated instability in operation of the six-inch continuous coker resulted in the design and fabrication of a new reference design for the continuous coker (discussed elsewhere in this report).

Data from the throughput studies was also used to determine the product slate as a function of process temperature. The distribution of the products is an important factor in sizing a commercial unit to accommodate the overhead, coke and gas production. Figure 19 provides the product slate (overhead, coke and gas) as a function of process temperature.

Examination of the results in Figure 19 show the overhead production decreased as a function of increased temperature. The data for a processing temperature of 825ºF may indicate a slight increase in the overhead yield as compared to the data at 782ºF. However, this experiment was not duplicated because of the decision to begin design of a new continuous coker. At this time the data point at 825ºF is of interest and will require further evaluation when testing resumes. We also noted that coke production increased with increased processing temperature. Conversely, produced gas exhibited a decrease.

The third area investigated with the continuous coker was the extent of removal of volatile material from the produced coke. In order to use the coke as a source of process fuel, it is necessary to maintain a level of volatiles necessary to promote coke ignition. Alternatively, if

the coke is to be land filled or produced for other uses, such as adsorbent, then it is desirable to minimize the volatile content.

The experiments were conducted by initiating the operation of the continuous coker using a zone 1 temperature of 850ºF and zone 2 temperature of 1000ºF. The temperature of zone 1 was maintained at 850ºF while the temperature of zone 2 was increased from 1000 to 1300ºF in 100ºF intervals. Samples of the coke were collected at each temperature and analyzed. For this study, the volatiles content in the coke was defined as the concentration of toluene soluble material remaining in the coke.

Table 13 summarizes the results for the coke drying study and shows a rapid decrease in the volatiles content from 1000 to 1100ºF, after which the volatiles content reaches a minimum. Because combustion studies have not been conducted on the produced coke, the concentration of volatile material needed for ignition is not known. The results indicate the possibility of some level of control of the volatiles concentration in the coke. However, the control range may not be sufficient to provide an optimum fuel product. If the coke product is to be land filled or used for other purposes requiring minimal volatiles content, the results indicate there should be no problem removing the volatile material.

Figure 17. The 6-inch pyrolyzer used to study the continuous coker technology.

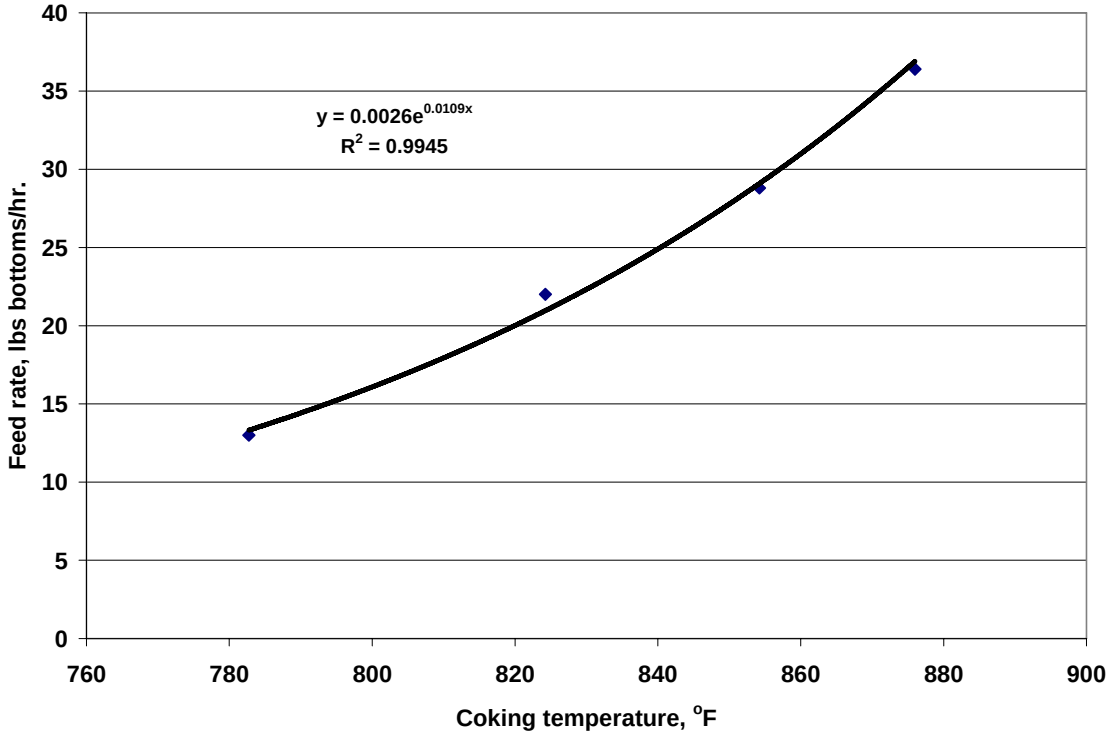


Figure 18. Maximum feed rate of bottoms achieved to the pyrolyzer as a function of coking (pyrolysis) temperature.

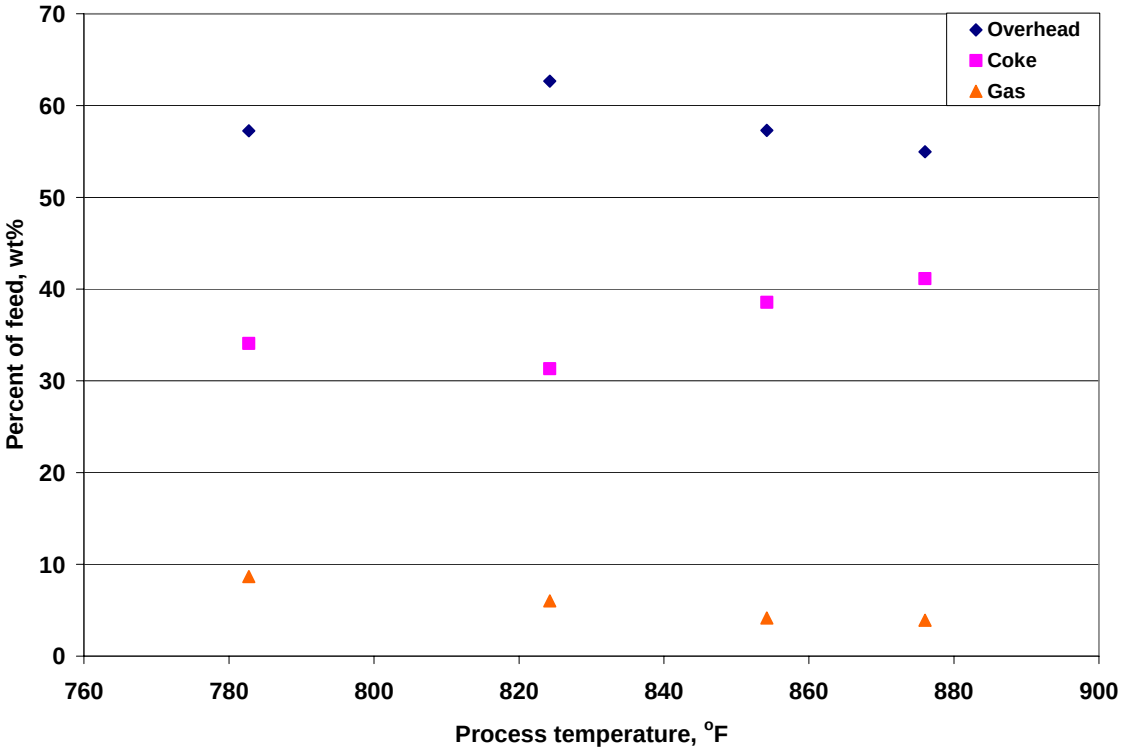


Figure 19. Product distribution from continuous coker as a function of coking (pyrolysis) temperature.

Table 13. Volatiles (Toluene Soluble) Content of Produced Coke as a Function of Zone-2 Temperature.

Zone 2 Temperature (°F)	Toluene Extractables in Coke, (%)
1001	30.9
1102	14.7
1202	0.1
1305	0.2

REFERENCE DESIGNS FOR DRU AND CONTINUOUS COKER

The reference designs reflect all the experience in process research and engineering gained to this point pertaining to the WRITE™ process. These designs employ WRITE™ technology but accomplish oil upgrading by using processing equipment and control methods identical or similar to that found in scaleable commercial designs. WRI's bench-scale reactors are valuable for exploring reaction concepts and developing fundamental engineering relationships of process conditions vs. yield, but are not easily or efficiently implemented at commercial scale. In addition, it was necessary to design and construct the next generation of engineering-scale DRU and continuous coking reactors that operate at higher throughput but use the same "commercial type" process equipment as would the field-scale pilots. These engineering-scale reactors would be used to verify the results from bench-scale testing, generate larger quantities of products for more detailed engineering evaluations, and simplify scale-up to field pilots.

DRU Reference Designs

The design developed by MEG's process engineering consultants and WRI personnel uses a scheme that sources process heat to the bitumen stream externally to the DRU reactor (Figure 20). Bitumen is first pumped through an oil furnace where it is heated to reaction temperature, and then flows into a vessel where it remains for a residence time necessary to complete the bitumen upgrading. A recycle of oil between the reactor vessel and furnace provides additional process heat if needed. A gas stream continuously purges the reaction vessel to promote the rapid removal of overhead products out the top of the reactor. After exiting the reactor, the overhead is cooled then fractionated into various streams, a portion of which is hydrotreated to reduce olefin content. A recycle stream composed of the non-condensable gases provides process heat for the furnace and reactor purge gas. The heavy, non-volatile fraction exits the bottom of the reactor where it is sent for additional processing. For the feed side of the DRU, a pressurized stream of diluted bitumen heats as it picks up recuperated energy from the oil furnace, then flows through flash separation stages where diluent and water are removed, yielding the anhydrous bitumen feed stream. The diluent is returned for reuse. This design offers

simplicity as well as efficiency and is readily constructed using process equipment commonly available for refinery applications.

The commercial concept was adapted to a 5-bpd, engineering-scale design shown in Figure 21 and maintained the same processing concepts and type of equipment. The design uses individual electric heaters for reactor oil, recycle oil, purge gas, and dilbit streams. Overhead oil is not fractionated or hydrotreated and the separated non-condensable gas stream is flared not recycled. The dilbit flash uses a single stage separation, instead of a more sophisticated multi-stage design. The 5-bpd engineering-scale facility has been constructed with MEG funding and sited at WRI's Heavy Oil Technology Center (HOTC) where it is currently in operation (Figure 22).

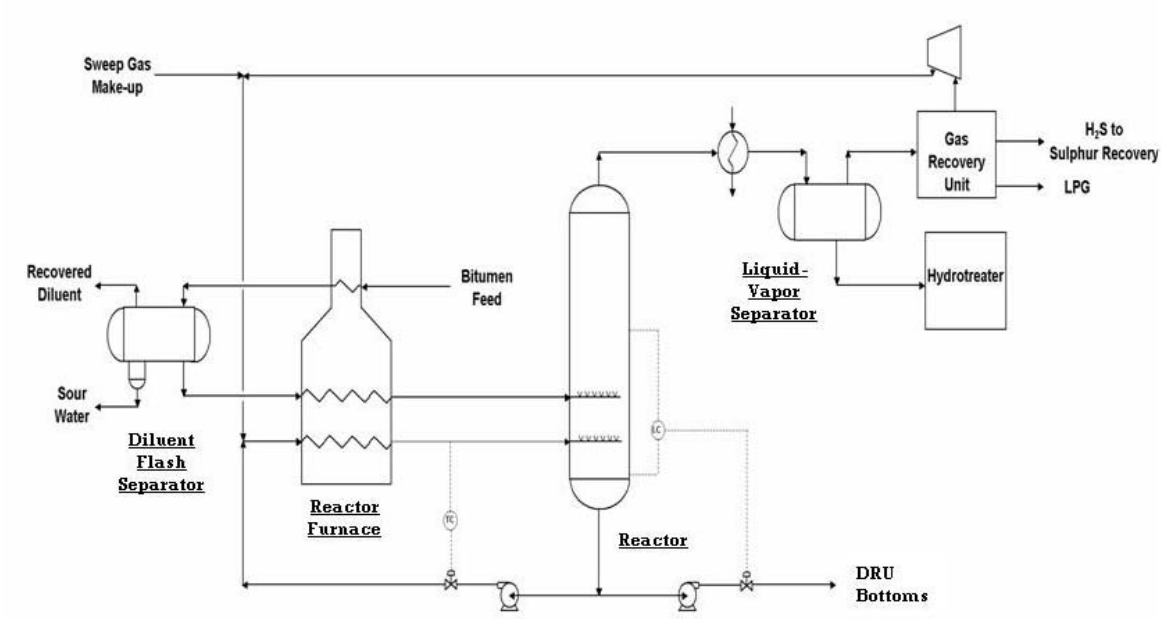


Figure 20. Commercial reference design for implementing WRITE™ Process' DRU.

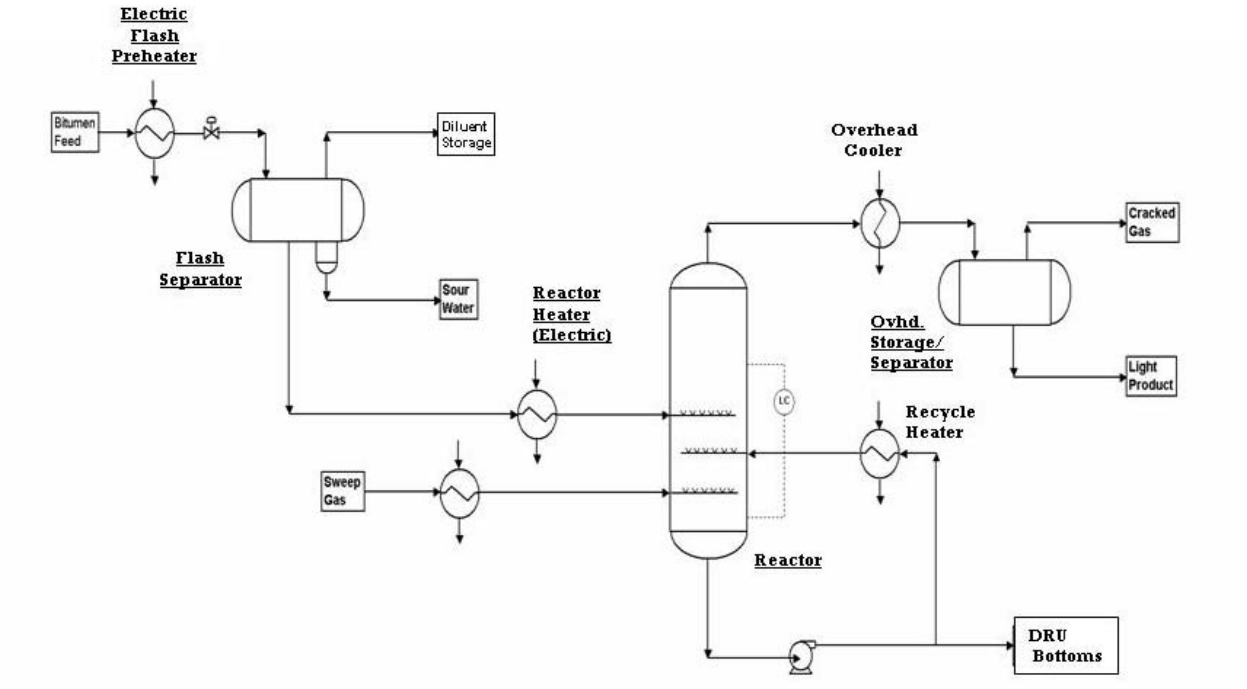


Figure 21. Engineering-scale design for 5-bpd DRU reactor based on reference design.

The 5-bbl DRU is built on two levels to conserve floor space, with a footprint of 8x16ft. The upper level contains tankage for recovery of stripped bitumen and DRU overhead. This level also houses the dilbit stripper and associated heater. The lower level houses the reactor, the bitumen feed tank, the bottoms tank, and associated heaters and transfer pumps. The reactor is located at the rear of the skid in this photo, so can not be seen. The equipment in the foreground at the right of the photo is the 2-inch continuous coker. The DRU is currently in operation at WRI.

Figure 22. Photo of the 5-bbl DRU reactor skid in place at WRI's HOTC engineering lab.

Continuous Coker Reference Design

A commercial design concept for the continuous coker has not been completed because additional process data is required. However, an engineering-scale design was developed as a revision to WRI's existing 6-inch bench-scale coker (Figure 17). The revised design uses a 2-inch diameter, twin-auger reactor to match the 5-bpd throughput of the DRU described earlier. A large volume of disengagement is provided at the region of the coker where oil pyrolysis occurs, to accommodate the rapid evolution and expansion of hydrocarbon vapors. This region of increased volume should eliminate the problems with screw flooding, experienced when the 6-inch coker was operated at temperatures above 875ºF. An improved hydrocarbon vapor recovery system is also specified to maximize recovery of overhead pyrolyzate. The system provides increased volumetric capacity and increased cooling through the use of a glycol-cooled heat exchanger specially designed to condense hydrocarbon vapors. The heat exchanger is located upstream of the final knockout pot. To provide additional safety for operations personnel and reduce the possibility of bridging at the reactor outlet, a discharge auger is specified to convey the produced coke to sealed collection containers. The auger also provides heat exchange for solids cooling. Figure 23 is a flow schematic of the engineering-scale, 2-inch continuous coker showing all major components. Figure 24 is a photo of the continuous coker under construction. Note the large triangular shaped vapor disengagement section in the center of the photo and the coke auger to the left, partially obscured by the upper section of coker's 2-inch screw barrel. The white vertical tank behind and partially obscured by the disengagement section is knockout-1. Construction of the 2-inch continuous coker is now complete and it is sited at WRI's HOTC building.

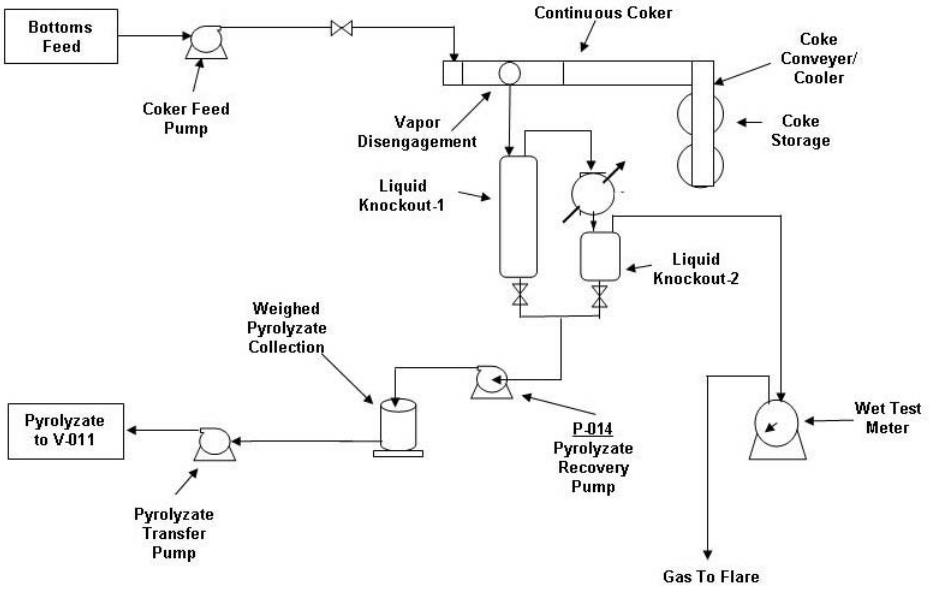


Figure 23. Flow schematic of engineering-scale continuous coker.

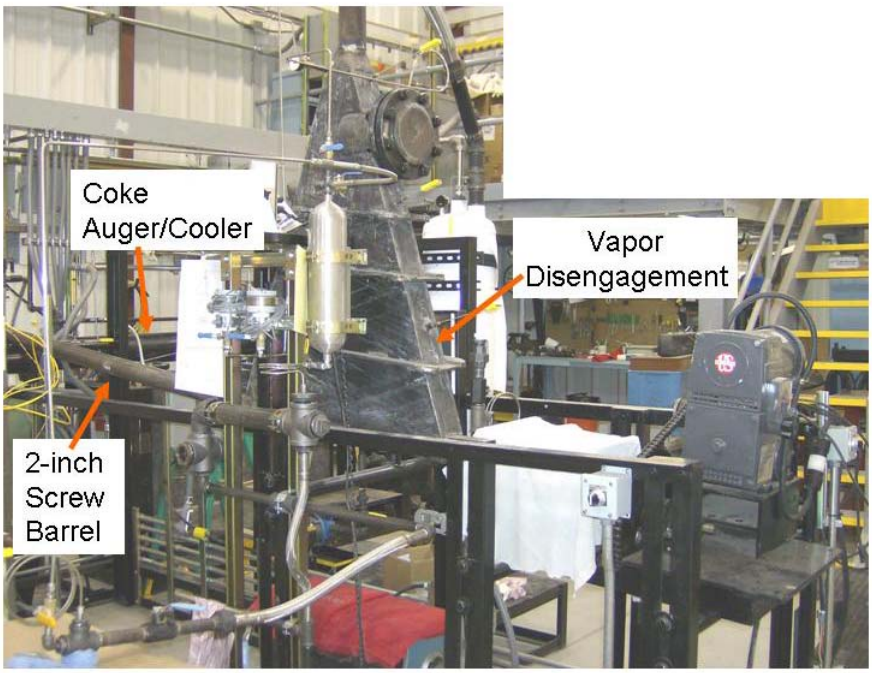


Figure 24. Photo of 2-in. engineering-scale reactor under construction.

DESIGN OF THE PILOT PLANT

A detailed design for a pilot-scale WRITE™ DRU has been completed and a bid package for construction and operation prepared. The pilot facility is designed to accomplish the following:

- confirm that the design criteria generated from the engineering-scale 5-bpd DRU reactor testing can be successfully scaled up to a commercial operating plant
- Obtain a more accurate prediction of operating costs
- Identify and resolve any associated operating problems
- Confirm yield and quality of the products generated from the process

Current plans call for the location of the pilot at MEG's production facility in Canada. The equipment is designed to operate continuously for 365 days per year for five years which should allow sufficient time for completion of pilot- and field-scale testing for the DRU. The design has sufficient flexibility to handle a range of operating conditions. The instrumentation specified will have sufficient complexity to allow detailed mass and energy balances needed for commercial plant design. The planned plant layout for the DRU reactor portion of the facility is shown in Figure 25.

The pilot design does not currently include the continuous coker because insufficient data and experience has been developed at this time to demonstrate its viability as a commercial process, although space has been provided for its inclusion in the plot plan.

The design of the DRU pilot generally follows the commercial reference design shown in Figure 20 and includes (1) diluent stripper, (2) reactor, (3) dedicated tank storage systems for feed and products, (4) blending system for producing mixtures of overhead products with bitumen or DRU bottoms to produce dilbit, (5) flare system, and (6) H_2S removal system. The produced sweet gas from the process will be recycled to reduce natural gas costs. The pilot facility will conform to all applicable construction, operating, and environmental codes.

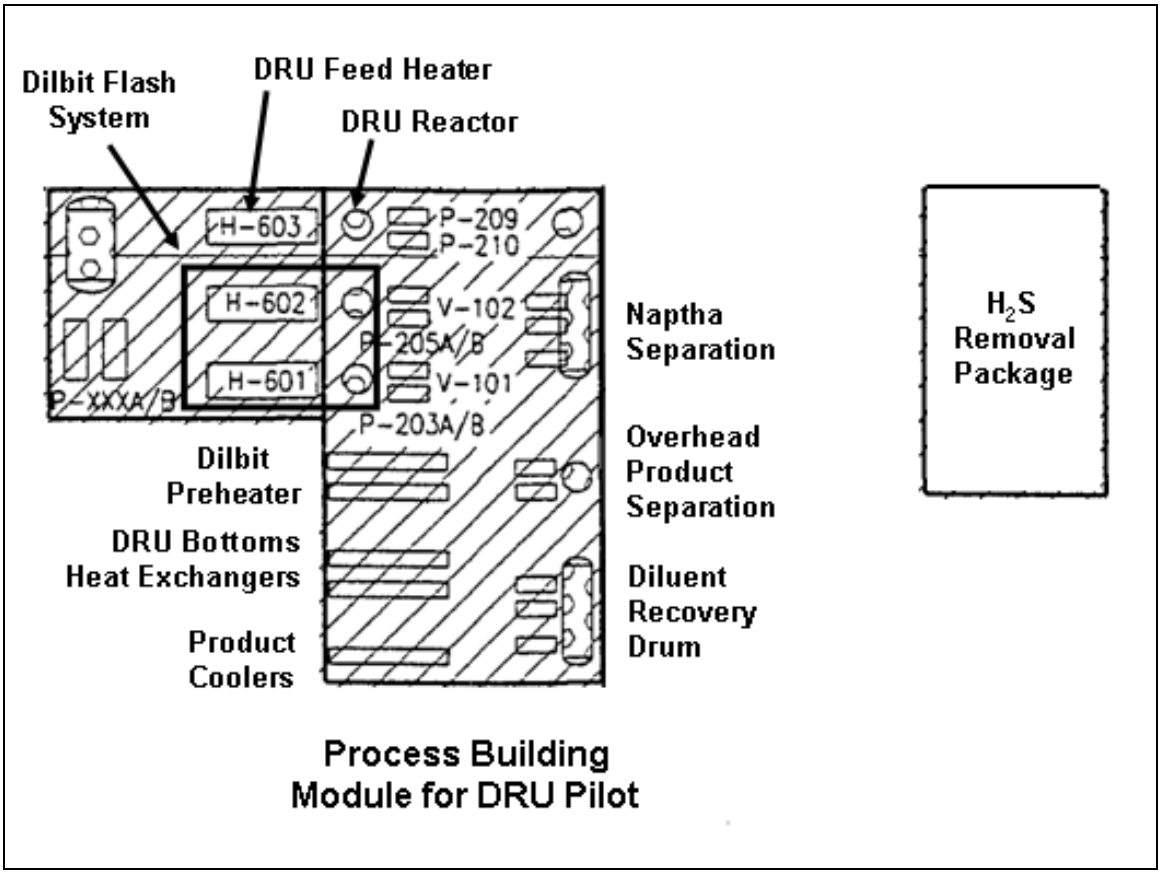


Figure 25. Planned equipment layout for DRU pilot at MEG's production facility.

HYDROTREATING STUDIES

Processes that pyrolyze oil, such as WRITE™, produce an overhead product from pyrolytic or cracking reactions in the absence of a hydrogen atmosphere. As a result, the overhead contains olefinic (alkene) compounds. Pipeline experience has shown that oils containing high concentrations of olefins in the naphtha fraction can cause gum formation, which may result in plugging of the pipeline. When this occurs, the plug must be removed causing down time and clean up expenses. To prevent this from occurring, the pipeline industry has

established limits on the amount of olefinic material that can be present in the naphtha fraction of oil entering the pipeline. This level, referred to as bromine number, is 1 gram of bromine reacted by 100 grams of the naphtha fraction. The solution to meeting the bromine number requirement for pyrolysis produced products is to hydrotreat the naphtha fraction to reduce the olefin concentration.

This task's objective was to identify the minimum hydrotreating conditions required to reduce the olefin content of the naphtha fraction from DRU product oils to meet pipeline specification (i.e. bromine number less than 1). To achieve this objective, the hydrogen consumption must be minimized to only saturate the olefinic bonds without using hydrogen for other chemical reactions (i.e. saturating aromatic rings or removal of heteroatoms). At the same time a high yield of liquid product must be maintained (minimize gas production).

The hydrotreating studies were conducted in a Chemical Data Systems (CDS) Model 810 Micro-Pilot Plant Reactor available at WRI (Figure 26). The system has the following capabilities:

- Maximum operating pressure – 1975 psi
- Maximum operating temperature - 1200ºF
- 5-cc catalyst bed
- 'Trickle flow' reactor

The hydrotreating studies were conducted over a range of conditions to provide the data to determine the minimum conditions and hydrogen uptake necessary to reduce the olefin content to below a bromine number of 1. The range of hydrotreating conditions employed for this study was as follows:

- Catalyst – Shell 424 (5 cc bed volume)
- Temperature range – 550 to 700ºF
- Hydrogen pressure range – 1300 to 1950 psig
- Liquid feed rate – 5 mL/hr
- Hydrogen flow rate – 4990 scf/bbl

The naphtha fractions were prepared by atmospheric distillation of the produced overhead at atmospheric conditions. For this study, the naphtha fraction was defined as the fraction of the produced overhead distilling between the initial boiling point and 225ºF. After the naphtha fractions were collected they were stored under a nitrogen blanket in a refrigerator to preserve the olefinic content.

Overhead samples produced from several bench-scale tests were available for use in this study. The tests producing these samples were conducted to evaluate the effect of residence time on the process. Thus, the independent variable was residence time expressed as normalized space velocity. Table 14 lists the samples available plus the elemental composition and the

bromine number of the naphtha fraction from each sample. The bromine number for the different samples range from 3.7 to 6.3 and all of the values were above the acceptable limit for pipeline quality. The wide range of bromine numbers were believed to be due to the fact that the samples were not "fresh" when the analyses were performed and some of the olefins polymerized to gum with standing. From this list of available samples, the sample produced at a reactor space velocity of 0.30*SV was selected for the hydrotreating study because it had the highest bromine number (6.3) and was presumably the most difficult of the sample set to hydrotreat.

The results of the hydrotreating study are presented in Table 15. The information contained in the Table includes the hydrotreating temperature and pressure, the mass balance, hydrogen balance, hydrogen consumption, liquid product yield and the bromine number of the hydrotreated product. Examination of the results in Table 15 shows that all of the hydrotreating conditions reduced the bromine number to a value below 1. Hydrogen consumption ranged from 60 to 140 scf/bbl.

The least hydrogen consumption was observed at the lowest processing temperature (550ºF) and the highest pressure (1950 psi) used in the study. This observation indicates that, with the Shell 424 catalyst, the lower temperature minimizes cracking reactions and the higher pressure indicates that sufficient hydrogen must be present to hydrogenate the olefins. These conditions are considered optimum (for this catalyst) to provide adequate reduction of the olefins without cracking other components or saturating aromatic rings. However, all of the conditions studied provide sufficient olefin reduction, liquid product yield, and hydrogen consumption to be acceptable for providing a product with a low bromine number.

It should also be noted that recent advances in development of hydrogenation catalysts has made available new catalysts that are capable of hydrogenating olefins under even mild conditions. Use of the newer catalysts is expected to provide olefin reduction with minimal hydrogen consumption using extremely mild processing temperatures. These results indicate the olefinic content of the produced overhead can be readily reduced with mild hydrotreating.

Figure 26. The Chemical Data Systems (CDS) Model 810 Micro-Pilot Plant Reactor used to conduct the hydrotreating studies.

Table 14. List of Samples from Bench-Scale DRU Tests Available for Use in Hydrotreating Study.

DRU Space Velocity (hr^{-1})	Carbon (wt%)	Hydrogen (wt%)	Nitrogen (wt%)	Sulfur (wt%)	Bromine Number (g Br/100 g oil)
0.48*SV	85.21	10.86	nd	1.89	5.4
0.42*SV	85.23	10.62	nd	1.97	4.9
0.36*SV	85.20	10.78	nd	2.32	6.3
0.30*SV	84.44	10.46	nd	2.05	3.4
0.24*SV	84.68	10.78	nd	2.23	4.9
0.18*SV	84.90	10.76	nd	2.48	3.7

nd is not detected

Table 15. Results Obtained from Hydrotreating the Naphtha Fraction from Bench-Scale DRU Produced Overhead.

Hydrotreating Temperature (°F)	Hydrogen Pressure (psi)	Mass Balance Closure (%)	Hydrogen Balance Closure (%)	Hydrogen Consumption (scf/bbl)	Product Yield (wt%)	Bromine Number (g Br/100 g oil)
700	1950	95.1	102.0	140	93.8	0.3
650	1950	99.8	100.8	122	98.7	0.2
600	1950	99.6	100.7	104	98.7	0.7
550	1950	97.4	101.1	60	95.9	0.2
700	1600	98.6	101.4	129	96.0	0.3
650	1600	99.2	101.6	111	97.5	0.5
600	1600	96.3	101.7	99	95.2	0.2
700	1300	98.2	102.0	139	96.6	0.3

COMMERCIAL AND TECHNICAL ASSESSMENT OF THE WRITE™ PROCESS

A preliminary assessment of WRITE™ conducted by MEG and their engineering consultant Triumph EPCM concluded that the process successfully converts bitumen to a SCO that meets pipeline gravity specification. (See the report "WRITE™ Process Commercial Assessment" attached to this document.) The assessment also concluded that WRITE™ is a technically feasible method of upgrading bitumen; however, additional data from a more comprehensive test program, currently underway and funded by MEG, is needed to complete a detailed commercial evaluation of the process.

Process and Economic Analyses

As part of a comprehensive commercial upgrading study conducted for MEG by Triumph, the WRITE™ field upgrading process was evaluated as a means for upgrading bitumen into SCO that meets minimum pipeline specifications. Material and utility balances were developed for a WRITE™ field upgrading complex using product yield and quality data obtained from tests conducted in WRI's bench-scale equipment.

WRITE™ was compared to delayed coking (DC), the standard reference of comparison for thermal upgrading of bitumen. Delayed coking uses two processing steps: First, a separation column removes the bitumen's distillable fraction as overhead. Second, the bottoms from distillation feeds through a high-temperature furnace then into a stationary vessel (coker), where the material reacts, under severe pyrolysis conditions, to form a light distillable oil and coke. After the reaction completes, the coke must be manually removed. Overhead product streams from distillation and coking combine to form SCO.

The study found advantages and challenges for WRITE™ when compared to delayed coking. WRITE™ produced more SCO (76wt% vs. 74wt%) and less coke (15wt% vs. 19wt%) than delayed coking (Figure 27 compares selected products on a weight basis). WRITE™ was also projected to produce 30wt% less CO_2 per barrel of bitumen fed (Figure 27). In addition, a WRITE™ facility had 24% lower capital and fixed costs than a comparable delayed coking

facility. Table 16 compares capital and utility costs between WRITE™ and delayed coking. These cost estimates are +/- 50% and calculated on a third-quarter, 2008 basis using Canadian dollars.

On the other hand, WRITE™ produced a higher density SCO (24.7 vs. 29.2 ºAPI) with a different product split that included less naphtha, less light gas oil (LGO), more vacuum gas oil (VGO), and resid (Figure 28). It should be noted that the bottom fraction shown in the WRITE™ product oil (Figure 28) probably resulted from an operational upset that "bumped" a small quantity of this material into the overhead light oil. WRITE™ processing by its nature would not be expected to produce a resid fraction in its overhead product. Neglecting its small resid fraction, an SCO with the boiling distribution produced by WRITE™ might be more expensive to refine than that produced by delayed coking, but this was not addressed in the study. Table 17 highlights important differences between delayed coking and the WRITE™ process in a field upgrading application.

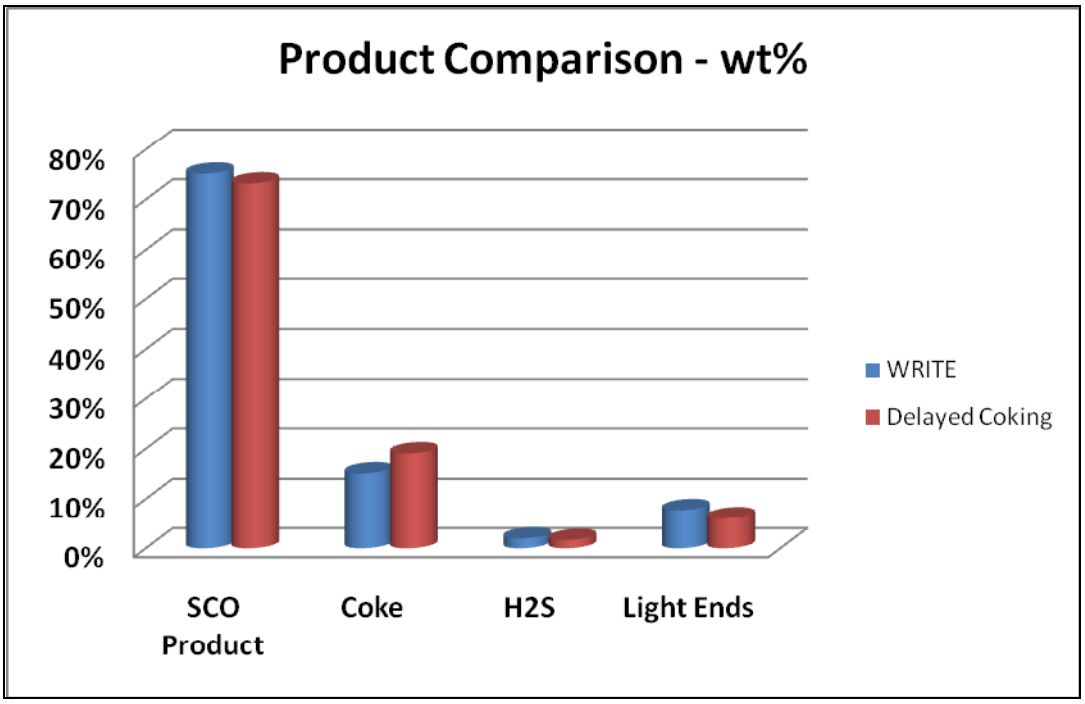


Figure 27. Comparison of products made from WRITE™ and Delayed Coking on weight basis.

Table 16. Capital and Utility Cost Comparisons for WRITE™ and Delayed Coking Facilities.
Basis: 100,000 BPD bitumen

	1	2	2-1	
CAPITAL COSTS	**Delayed Coking**	**WRITE™**	**difference**	**Comment**
Capital Cost (*MM$*)	4,092	3,123	-969	
$C /bbl Bitumen per day	40,917	31,226	-24%	WRITE™ is cheaper
Utility Estimates				
Variable Cost (*Per bbl Bitumen*)				
Make-up Water (*lb*)	158	127	-31	WRITE™ has lower water use
Make-up Diluent (*bbl*)	0.00925	0.00925	0	
CO_2 Emission (*lb*)	56.4	43.4	-13	WRITE™ has lower CO2
Catalysts & Chemicals (*$*)	0.040	0.046	0.006	
Fixed Cost (*k$/day*)	346	264	-82	WRITE™ has lower fixed costs

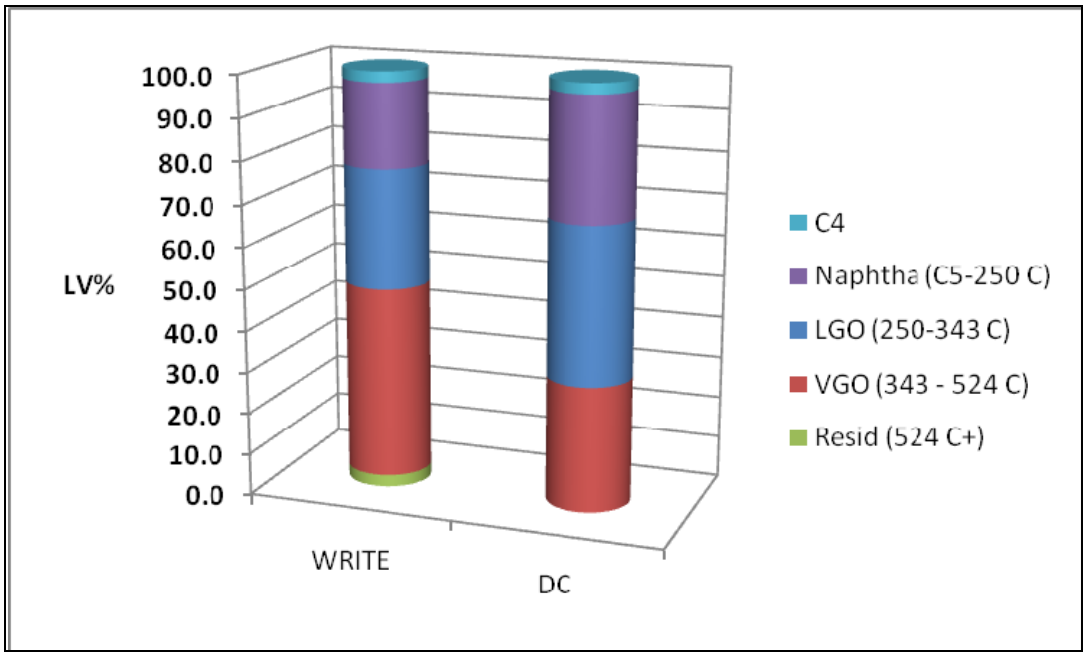


Figure 28. Comparison of boiling distributions from WRITE™ and Delayed Coking SCO.

Table 17. Important Differences Between WRITE™ and Delayed Coking.

Item	Delayed Coking	WRITE™ Process	Comment
Yield of SCO	Base	Same	Expect WRITE™ to be 3 lv% higher. Requires steady-state pyrolyzer operation.
Gravity	29.2	24.7	Delayed Coking SCO is lighter
Product Quality	Base	Heavier, higher sulphur	Heavier, higher sulphur SCO product from WRITE™
Coke Make	Base	Base – 20 wt%	Lower coke make with WRITE™
Fuel Gas Make	Base	Base – 40%	Lower intensity cracking with WRITE™
CO_2 production	Base	Base – 30 wt%	Lower CO_2 production with WRITE™
Make – Up Water	Base	Base – 20 wt%	Lower make-up water use for WRITE™
Capital Cost	Base	Base – 24%	Lower capital costs for WRITE™
Fixed Cost	Base	Base – 24%	
Technical Risk	Low	Medium	The pyrolysis unit of the WRITE™ process is unproven
Other	Commercial	Not commercial	Delayed coking is industry proven

Feasibility of WRITE™ for Commercial Applications

The process and economic analysis (discussed above) indicates that WRITE™ technology offers potential benefits in a field upgrading application compared to delayed coking. The lower processing severity used in WRITE™ provides comparable yields of SCO to delayed coking, while producing less coke by-products. WRITE™'s lower capital costs imply favorable economics. (It should be noted however that results from the process and economic analysis assumed the continuous coker operated under steady state conditions, which has not been demonstrated in bench-scale testing.) Its potential benefits notwithstanding, considerable work remains before WRITE™ can be deployed as a field-scale pilot at MEG's production facility:

The study indicated that assumptions of WRITE™'s yields and performance will require confirmation and validation using bitumen from MEG's SAGD production site. Moving from WRI's bench-scale to MEG's engineering-scale (5 bpd) equipment, with a comprehensive experimental and analytical program, will confirm process yield and quality estimates. In addition, data from the engineering-scale reactors will provide information for a detailed technical and economic analysis that *must* precede the deployment of WRITE™ as a field-scale pilot.

Recall the engineering-scale reactors were based on reference designs that employ WRITE™ technology but accomplish oil upgrading by using processing equipment and control methods identical or similar to those found in scaleable commercial designs. Therefore, test results obtained from the 5-bpd equipment should provide a reasonable basis for projecting performance and economic factors to pilot and commercial scale.

The DRU poses a low risk for projecting to commercial and demonstration scale because of its extensive testing at bench-scale, its simplicity, and the consistent results obtained to date in the engineering-scale reactor. Conversely, the continuous coker has seen relatively little development. So a higher level of uncertainty exists regarding this reactor's performance, cost, and reliability.

The study indicated that an effective disposal or reuse of the coke by-product is the key to the successful commercial application of a carbon rejection process, such as WRITE™. MEG's production site currently has no access to coke disposal outlets, so exploring options for reuse is mandatory. The engineering-scale continuous coker will need to operate for extended periods to produce sufficient quantities of pyrolysis coke for alternative fuels testing in combustion or gasification that can generate "clean energy" to support SAGD steam production with CO_2 sequestration potential.

MEG continues to support the development of WRITE™ technology at the 5-bpd engineering scale. The goals of continued WRITE™ process development are:

- Define more precisely the WRITE™ process' operating envelope with regard to yields and SCO quality
- Build improved confidence with extended operation of the continuous coker
- Provide the process basis for a factored, equipment-based, capital cost estimate necessary for deploying WRITE™ as a field-scale pilot (accuracy of +30/-15%).

ESTABLISH A PETROLEUM ANALYSIS LABORATORY

The purpose of this task is to establish and staff a petroleum analytical lab to support the development of ongoing and future hydrocarbon recovery and conversion technologies by WRI and collaborators. The laboratory is intended to support current and future WRI research activities, not to compete with commercial entities. In the early stages of WRITE™ process development, MEG and WRI experienced high analytical costs and inevitable delays when sending samples for analysis. Consequently, MEG and WRI extended the Task 51 JSR agreement to establish and operate a petroleum analysis laboratory. MEG's funding partially offset their long-term analytical costs and provided WRI with the necessary analytical equipment. WRI provided the space to house the analytical lab in its HOTC building, located at WRI's Advanced Technology Center. Matching funds from DOE were used to staff and operate the lab for one year.

A laboratory to support the analytical needs of process development cannot provide every conceivable petroleum analysis method. The scope of the analytical services for the laboratory was initially defined in consultation with SNC-Lavalin and MEG personnel. Three criteria were established for the selection of analytical procedures provided by the laboratory. First, those analyses that provide data for control and monitoring of the process being developed; second, those procedures that provide data required to complete material balance closure; and three, procedures which specify short shelf life for the chemical integrity of the species being analyzed. Based on these criteria, thirteen analytical procedures were identified for inclusion in the laboratory's mission. These methods are established American Standard Tests and Measurements (ASTM) procedures for analysis of petroleum and related products and are listed in Table 18.

Set up of the laboratory space required providing the area with water and sewer services from elsewhere in the building, providing additional electrical service, and installing laboratory bench space. Concurrently with setting up the laboratory space, all of the analytical equipment and associated supplies were requisitioned. When the analytical instrumentation arrived, factory representatives conducted equipment setup and training of laboratory personnel.

The analytical laboratory was operational when experimentation began on the 5-bpd, engineering-scale DRU. During a typical 10-day operation of the DRU facility, a total of 5 experimental conditions are studied. Samples from five separate process streams are collected for analysis. Each stream requires analyses by at least eight different ASTM methods. This results in over 200 individual analyses that are performed by the laboratory for each 10-day pilot plant test period. These analyses are performed during and after the test period and the turn around time for any particular analysis is less than two weeks. The rapid turn around of the analytical results has made planning of experiments more efficient than it was when out side laboratories were used for the analytical work. Two additional analytical methods are being added to those performed by the laboratory, total acid number (TAN) and olefins by NMR (nuclear magnetic resonance). As the development of the process continues, it is anticipated additional analytical methods will be added. Figures 29 and 30 show some of the equipment available in the laboratory.

Table 18. ASTM Analytical Methods Currently Performed in the Laboratory

ASTM Method	Method Title
D1159	Bromine Number
D86	Atmospheric Distillation of Petroleum Products
D5018	Shear viscosity
D445	Kinematic Viscosity
D2320	Relative Density of Solid Pitch (Pycnometer Method)
D 4052	Density by Digital Density Meter
D1298	Density by the Hydrometer Method
D4530	Micro Carbon Residue (equivalent to Conradson Carbon Residue)
D6560	Determination of Asphaltene Concentration
D2887 and D5307	Simulated Distillation
D6374	Volatile Matter in Green Petroleum Coke
D5291 and D1552	Carbon, Hydrogen, Nitrogen and Sulfur
D2207	Hydrocarbon Types

Figure 29. Balances, drying oven, MCR and density meter available in the laboratory.

Figure 30. Simulated distillation equipment available in the laboratory.

CONCLUSIONS AND STATUS

An economic screening completed by Triumph EPCM for MEG concluded that the WRITE™ process is a technically feasible method of upgrading bitumen and that it produces a SCO that meets pipeline specifications.

Tests conducted with WRI's bench-scale DRU yielded 30 to 50wt% (relative to feed) of distillable overhead, considerably higher than the 20wt% recoverable by distillation. The quantity of overhead produced increases with increased temperature and reduced space velocity.

The DRU's overhead meets Canadian pipeline specifications for density and viscosity, while maintaining a consistent quality that is relatively insensitive to processing severity. For the range of conditions studied, the produced overhead's density and viscosity ranged from 27 to 30ºAPI and 6 to 9 cSt at 68ºF, respectively. These oil properties compare favorably to Canadian specifications of 19º API gravity or lighter and a viscosity of 350 cSt or less.

Tests conducted in WRI's bench-scale continuous coker were expected to provide information regarding its maximum conversion rate as a function of temperature. Having this relationship would allow a reactor design that maximizes throughput at a minimum size. Tests conducted up to 875ºF demonstrated that throughput of DRU bottoms increases with temperature. However the planned tests could not be completed because the coker's recovery system flooded at higher temperatures. This shortcoming has been addressed in a revised 5-bpd design.

The results from DRU and continuous coking tests were used to generate a commercial reference design for the DRU and an improved 5-bpd continuous coker. These designs were used to complete a preliminary design for a field-scale DRU pilot that would be sited at MEG's production facilities. A similar design could not be produced for the continuous coker because of insufficient data and operating experience at this time.

MEG also used the reference designs to construct 5-bpd, engineering-scale DRU and continuous coking reactor systems. These reactors are located in WRI's Heavy Oil Technology Center (HOTC) building and will be used to provide more detailed information for scaling WRITE™ to a field-scale pilot.

An economic screening and technical evaluation of the integrated WRITE™ process was performed by MEG and its consultants. The study used delayed coking as the basis for evaluation, the industry accepted method for thermal upgrading of bitumen. The study found advantages and challenges for WRITE™: WRITE™ was projected to produce more SCO, less coke, less CO_2 per barrel of bitumen fed, and had lower capital and operating costs. On the other hand, WRITE™'s lower processing severity yielded a material with higher density and a different product distribution for naphtha, light gas oil and vacuum oil that, taken together, might reduce the value of the SCO.

The study advised that a comprehensive test program needs to be conducted to provide additional information for more detailed process evaluations and associated economics. These evaluations must be completed before the integrated WRITE™ process is deployed as a field-scale pilot. The recommended program includes operating the 5-bpd engineering-scale reactors with bitumen produced from MEG's production facilities; better characterizing the performance of the continuous coking reactor, and operating the continuous coker for extended periods of time to gain operating experience and to produce sufficient coke for combustion and gasification studies.

MEG continues to support the WRITE™ process by supporting testing at the 5-bpd, engineering-scale.

A petroleum analysis laboratory was established and staffed to support heavy oil upgrading activities at WRI's HOTC. The laboratory currently supports the test program conducted in the 5-bpd engineering-scale reactor.

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WRITE Process Commercial Assessment

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MEG Energy Corp.

Date: March 31, 2009

WRITE Process Commercial Assessment

Basis: Commercial Upgrader Study – Phase 1A Report
Prepared by: Triumph EPCM, January 2009

Executive Summary

MEG Energy Corp. (MEG) is a bitumen production company with plans for in excess of 200,000 barrels per day of production from the oil sands reserves south of Fort McMurray, Alberta, Canada. MEG has been supporting the development of the WRITE process, since 2003, for potential application in field upgrading of Steam Assisted Gravity Drainage (SAGD) produced bitumen.

An experimental program using the WRITE process has been conducted at the bench scale (1 BPD) level. The results from this experimental program have established "proof of concept" of the WRITE process for upgrading bitumen using the combination of the thermal cracking and pyrolysis. The early indications of attractive process yields have lead to additional study work of the WRITE process in a field upgrading application.

A commercial study of upgrading technology has been conducted by Triumph EPCM. In this study, a field upgrading complex using the WRITE process has been developed and compared to a complex using delayed coking, the most widely used thermal conversion process in the oil sands industry.

The WRITE process appears to offer a yield advantage over a commercial upgrading complex based on delayed coking technology. The low pressure operation of WRITE, coupled with lower coke production, produce higher liquid product volumes. These indicators suggest that the WRITE process has good performance potential. However, the experimental program was unable to achieve stable operation of the pyrolyzer, creating problems in reconciling the material balance. The process yields of the WRITE process obtained from the 1 BPD pilot are believed to be understated.

Validation of the WRITE process capabilities can be achieved through the completion of a comprehensive experimental program at the 5 BPD scale, followed by a thorough technical and economic evaluation of the data.

Background

As part of a commercial upgrading study conducted for MEG Energy Corp. by Triumph EPCM (Calgary Alberta), the WRITE Field Upgrading process has been evaluated as a means for upgrading bitumen into synthetic crude oil (SCO) meeting minimum pipeline specifications.

The application of the **W**estern **R**esearch **I**nstitute **T**hermal **E**nhancement (WRITE) Process in a field upgrading complex has been assessed and evaluated in comparison to a Delayed Coking reference case. The field upgrading process is sized for 100,000 BPD of bitumen feed.

Method

Using WRITE process yield and quality estimates from the bench scale (1 BPD) work conducted in 2003 – 2004, material and utility balances have been developed for a field upgrading complex.

The field upgrading complex includes primary and secondary upgrading processes necessary to produce an upgraded bitumen product that meets pipeline specifications for gravity, viscosity and olefin content (bromine number).

In this assessment, the performance of the WRITE process is compared to a Delayed Coking based field upgrading complex using screening level technical and economic measures.

Assessment of the WRITE Process

The block flow diagram of the WRITE process in a 100,000 *BPD* field upgrading application is shown in Figure 1.

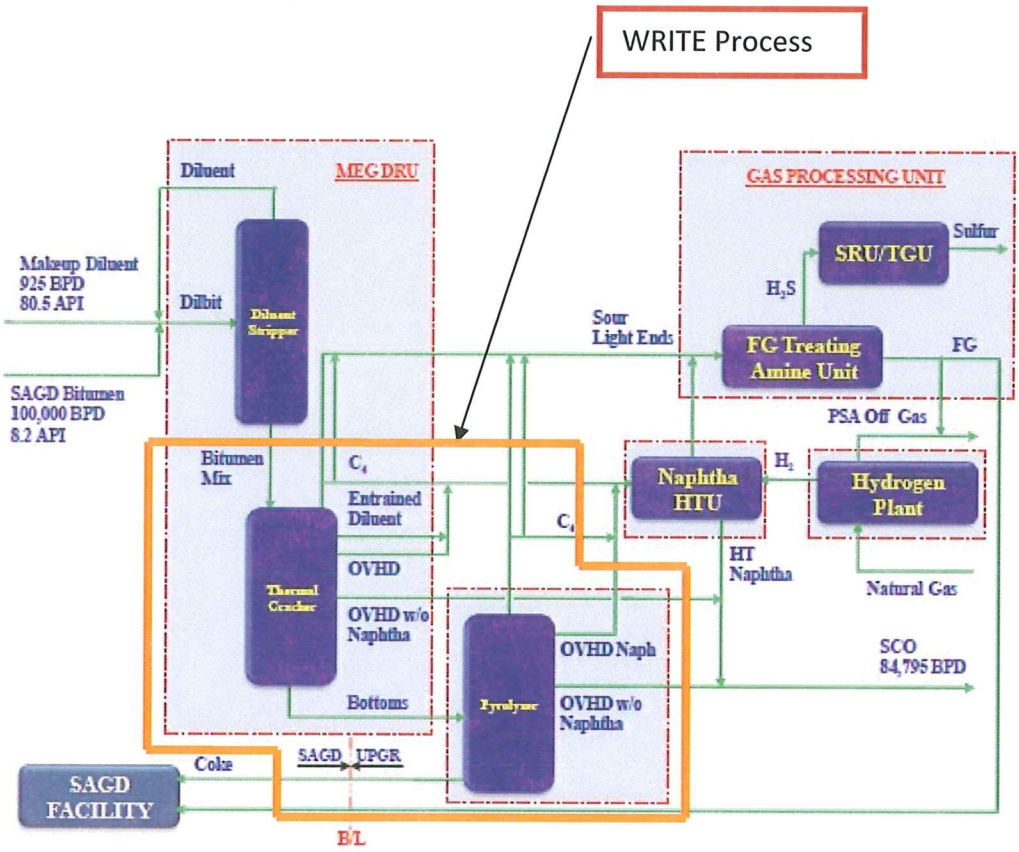


Figure 1 – Block Flow Diagram of the WRITE process (Field Upgrading)

Process Description

The WRITE process two processing stages in a field upgrading application for processing bitumen:

- Mild thermal cracking of raw bitumen followed by
- Severe thermal cracking (pyrolysis) of the heavy bottoms product

In a commercial field upgrading application, the WRITE process would convert SAGD produced bitumen into sour synthetic crude oil (SCO) meeting pipeline transport specifications. The thermal cracking and pyrolysis units of the WRITE process would be supported by the necessary process units for preparing bitumen feedstocks, product stabilization, and by-product separation.

Supporting units for the WRITE process in this application include:

- Diluent stripping
- Naphtha hydrotreating (for olefins control)
- Hydrogen plant
- Gas processing
- Sulphur plant

Diluent Stripping Unit

In this field upgrading application, 100,000 *BPD* SAGD produced bitumen which is diluted with sufficient lighter hydrocarbons for field separation. Diluents are typically naphtha or natural gas condensate required to meet the SAGD processing requirements. This diluted bitumen (118,500 *BPD*) is fed to the first stage of the process, the Diluent Stripper. Diluent is recovered from the overhead of the Diluent Stripper and returned to the SAGD facilities.

Mild Thermal Cracking Unit

Diluent Stripper bottoms (100,925 *BPD*) feed the Mild Thermal Cracking unit (reactor), the first step of the WRITE process. The bottoms from the Thermal Cracker (34,216 *BPD)* are routed to a Pyrolyzer, the second step of the WRITE process.

The Diluent Stripper Overhead stream (14,161 BPD) is primarily cracked naphtha. This naphtha (combined with pyrolyzer naphtha) is sent to an 800 psig hydrotreater for stabilization, desulphurization, and denitrification. This naphtha will be treated sufficiently to meet pipeline quality specifications for olefin content (bromine number).

Pyrolyzer (Severe Thermal Cracking)

Mild Thermal Cracker bottoms are fed to the pyrolyzer (34,216 BPD). This is a severe thermal process with continuous coke removal using an inclined twin screw system. Products from the Pyrolyzer are sour light ends, naphtha, cracked blend (gasoil and diesel boiling range), and coke.

Sour light ends are treated in the gas processing unit. Cracked naphtha is hydrotreated. Cracked blend (gasoil and diesel) is routed to SCO blending.

Naphtha Hydrotreater

The naphtha hydrotreater consists of two reactors in series with a capacity of 19,177 *BPD*. The first reactor is dedicated to di-olefin saturation, and the second reactor, to desulphurization and denitrification. Detailed information should be confirmed by licensors at next stages. The naphtha hydrotreater will be operated at 800 *psig* and a liquid hourly space velocity of 2.5 h^{-1}. The unsaturated light ends will be hydrotreated for stability. The desulphurization and denitrification level can be adjusted according to specific requirement.

Hydrogen Plant

A hydrogen plant will be required to achieve the hydrogen generation capacity of 16.3 *MMSCFD*, which is required for hydrotreating. The steam methane reformer associated with pressure swing adsorption (PSA) operation produces high purity hydrogen. The conventional hydrogen plant will use the following facilities:

- Natural Gas Desulphurization
- Steam Methane Reformer
- Two stage shift conversion
- CO_2 absorption
- Hydrogen compression

Gas Processing Unit

The unit includes an Amine Unit and a Sulphur Recovery Unit (SRU)/Tail Gas Unit (TGU). The Amine Unit removes 13,432 *kg/hr* of H_2S from 65,880 *kg/hr* of sour light ends to produce 701 *Megawatts (MW)* of fuel gas. The SRU/TGU converts H_2S to produce 303 *tonnes/day* of sulphur.

Process Performance

The WRITE process yield estimate is based on bench scale (1 BPD) operation of the mild thermal cracking unit and the pyrolyzer. The combined liquid product yield calculated from the experimental program is believed to low. The pyrolyzer did of achieve the required level of steady state operation to have confidence in the process yields. This unstable operation of the pyrolyzer has resulted in yields that are suspect as they did not show the yield benefits from low pressure operation.

As both the pyrolyzer and delayed coking processes are both severe thermal cracking technologies, the performance expectations of the pyrolyzer should follow industry guidelines for delayed coking operations. Liquid product yields are expected to increase by 1 lv% for each 10 psi drop in coking pressure.

The pyrolyzer will operate at lower pressures than commercial delayed coking (excess of 30 psi lower). There is an expectation of up to 3 lv% yield increase compared to commercial delayed coking which operates at 40+ psig.

Notwithstanding the confidence issues with the WRITE process yields; this upgrading configuration has been developed and estimates of production yields and qualities of synthetic crude oil (SCO) have been produced. The yield of product synthetic crude oil is 84.8 lv% with a gravity of 24.7 °API. Pyrolysis coke rejected is 15 wt% of the bitumen feed.

	Volume		Weight	
Summary:	**bpd**	**lv%**	kg/hr	wt%
in				
Bitumen	100,000		671,046	99%
hydrogen			1,731	0%
diluent loss			4,090	1%
Total			676,867	100%
out				
SCO	84,795	84.8%	508,764	76%
coke			102,223	15%
other			65,880	10%
Total			676,867	101%

Table 1 – WRITE Process Material Balance

SCO Quality

The SCO produced from a commercial upgrader incorporating the WRITE process is expected to be 24.7 °API with 2.5 wt% sulphur. Nitrogen levels will be 0.11 wt%. This would be characterized as a heavy, sour synthetic crude oil (SCO).

The final SCO blend is predicted to have the following composition and properties:

Item	units	Value
volume	BPD	84,795
API	°API	24.74
Sulphur	wt%	2.52
Nitrogen	wt%	0.11
volume fractions		WRITE
C4	lv%	2.60
Naphtha (C5-250 °C)	lv%	20.09
LGO (250-343 °C)	lv%	28.57
VGO (343 - 524 °C)	lv%	45.88
Resid (524 °C+)	lv%	2.86
		100.00

Table 2 – WRITE Process SCO Quality

WRITE Process SCO Composition

The composition of the SCO from the WRITE field upgrading complex is shown in the Figure 2.

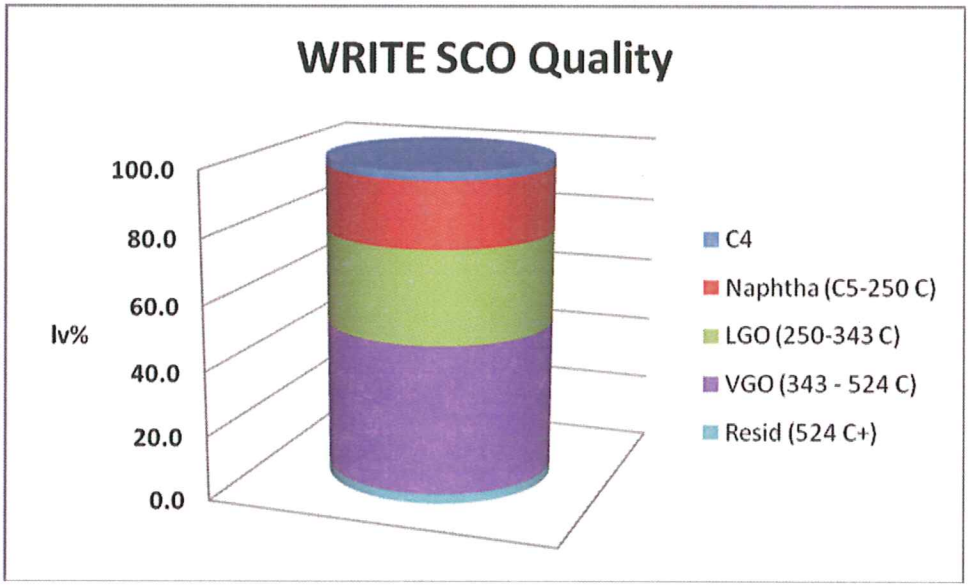


Figure 2 – WRITE Process SCO composition

Observation: WRITE SCO contains some residual (524 ^{o}C+) material.

Utility and Cost Estimates

Utility Balance

The utility balances have been developed to reflect the needs of the commercial upgrading complex based on the WRITE process.

In addition to the primary upgrading process, there is a utility supply/demand associated with supporting units, including:

- Sulphur Plant
- Amine Absorption and Regeneration
- Sour Water Stripper
- Cooling Towers
- Power Plant
- Deaerators (water treatment)
- Demineralization Plant (water treatment)
- Steam Letdowns
- Steam Turbine Generators
- Process Boilers

The utility balances for the WRITE process configuration is included as Attachment 2. The basis for this utility balance is summarized in Appendix 1.

Capital Cost

The capital cost estimates for the upgrading complex. There are order of magnitude capital cost estimate (+/- 50%), Class V, curve type estimates for a remote Canadian location, developed by Triumph EPCM.

Details on the development of the capital cost estimate are included in Appendix 2.

WRITE Process

CAPITAL COSTS	WRITE
Capital Cost (*MM$*) (+/- 50%) *note 1	3,123
$/bbl Bitumen per day	31,226
Utility Estimates	
Variable Cost (*Per bbl Bitumen per day*)	
Make-up Water (*lb*)	127
Make-up Diluent (*bbl*)	0.00925
CO_2 Emission (*lb*)	43.4
Catalysts & Chemicals ($)	0.046
Fixed Cost (*k$/day*)	264

**Note 1: $ Canadian, Remote site location*

Table 3 – Capital Cost and Utility Summary

Capital costs are based on Q3 2008 estimates in $ CDN for a field upgrading complex in a remote Canadian location.

Delayed Coking Field Upgrading

For a Field Upgrading application, a Delayed Coking based complex producing sour synthetic crude oil (SCO) would have a process flow diagram (PFD) as illustrated in Figure 3.

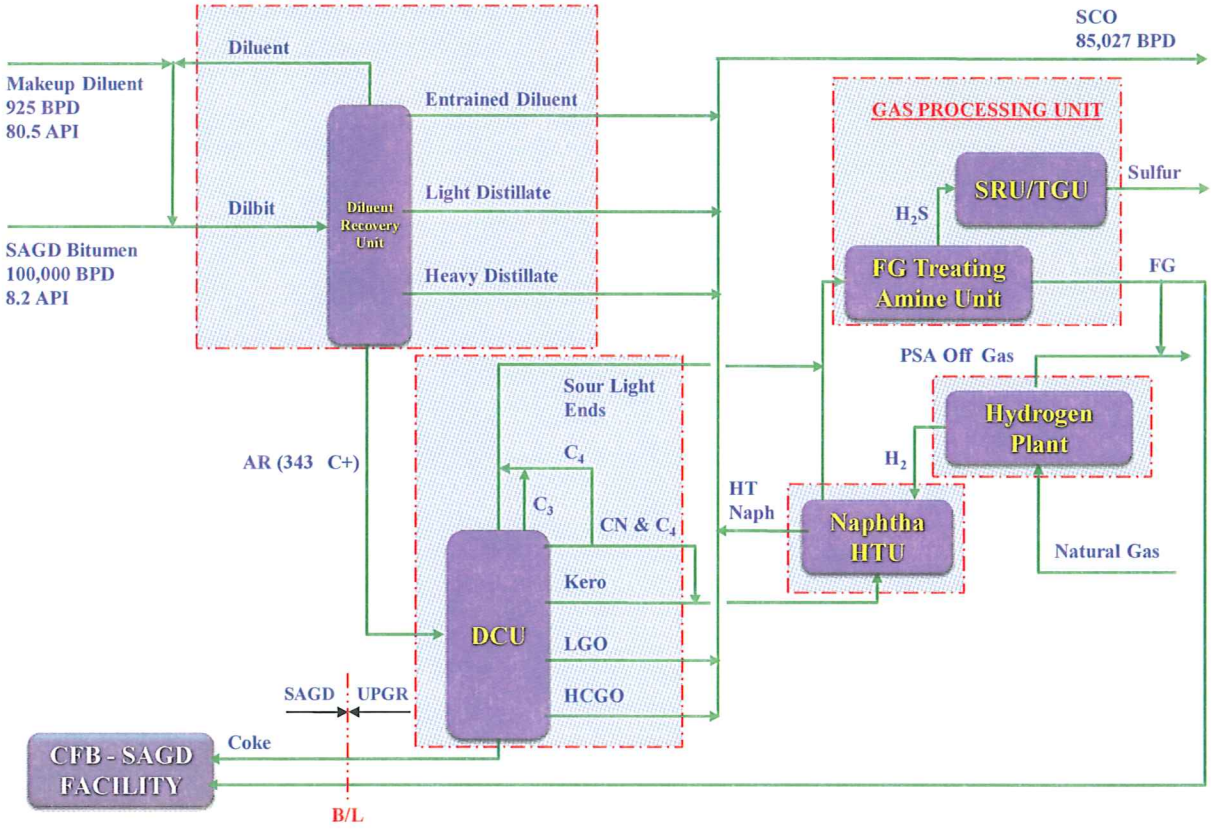


Figure 3 – Delayed Coking Based Field Upgrading Complex

Process Description

The field upgrading complex based on the Delayed Coking process has two primary processing steps:

- Initial fractionation and separation of diluent and atmospheric products from bitumen
- Delayed Coking of DRU bottoms - Atmospheric Residue (AR)

The feed is diluted bitumen from the SAGD production facility. The primary product is sour synthetic crude oil (SCO).

Supporting units for the Delayed Coking process in this application include:

- Naphtha hydrotreating (for olefins control)
- Hydrogen plant
- Gas processing
- Sulphur plant

Diluent Recovery Unit

In this field upgrading application, 100,000 *BPD* of SAGD produced bitumen which is diluted with sufficient lighter hydrocarbons for field separation. Diluents are typically naphtha or natural gas condensate required to meet the SAGD processing requirements. The Diluent Recovery Unit (DRU) is a single atmospheric distillation tower with a preflash tower, and able to process 118,500 BPD of dilbit at 16.0 ^{o}API.

Diluent is recovered (95 vol%) from the overhead which is returned to SAGD facility. The cut point of Atmospheric column Residue bottoms (AR) was set at 343^{o}C (650 ^{o}F) to produce a stream of light gas oil totaling 12,000 BPD for SCO storage blending. The remaining 88,000 BPD of atmospheric residue (AR) is sent to the Delayed Coker for conversion.

Delayed Coking Unit

The delayed Coker Unit is able to process 88,000 *BPD* of AR. It is arranged as two pairs of coke drums (total 4) each with an independent fired heater and operated at a pressure of 40 *psig* and a cycle time of 14 *hrs*. Residue is heated to high temperature and cracked inside Coker Drums. Sour light ends are extracted from the Coker fractionator overhead. 26,078 *BPD* cracked naphtha is routed to naphtha hydrotreater. 45,850 BPD gas oil is directed to SCO storage. 3,102 *tonnes/day* of coke is available to fuel steam generation for Steam Assisted Gravity Drainage (SAGD) bitumen production.

Naphtha Hydrotreater

The naphtha hydrotreater consists of two reactors in series with a capacity of 26,078 *BPD*. The first reactor is dedicated to di-olefin saturation, and the second reactor, to desulphurization and denitrification.

The details of the coker naphtha hydrotreating process will need to be confirmed by hydrotreating process licensors at next stage of design development. The naphtha hydrotreater are anticipated to operate at 800 *psig* with a liquid hourly space velocity of 2.5 h^{-1}.

The unsaturated light ends will be hydrotreated for stability. The desulphurization and denitrification levels of the hydrotreating process can be adjusted to meet specific quality targets.

Hydrogen Plant

A hydrogen plant will be required to achieve the hydrogen generation capacity of 24.2 *MMSCFD*, which is required for hydrotreating. The steam methane reformer associated with pressure swing adsorption (PSA) operation produces high purity hydrogen. The conventional hydrogen plant will use the following facilities:

- Natural Gas Desulphurization
- Steam Methane Reformer
- Two stage shift conversion
- CO_2 absorption
- Hydrogen compression

Gas Processing Unit

The Gas Processing Unit includes an Amine Unit and a Sulphur Recovery Unit/Tail Gas Unit. The Amine Unit removes 11,196 *kg/h* of H_2S from 52,381 *kg/h* of sour light ends to produce 533 *MW* of fuel gas. The SRU/TGU converts H_2S to produce 252 *tonnes/day* of sulphur.

Process Performance

The Process Flow Diagram for the Delayed Coking based field upgrading complex is included as Attachment 3.

Delayed Coking Process

	Volume		Weight	
Summary:	**bpd**	**lv%**	kg/hr	wt%
in				
Bitumen	100,000		671,046	99%
hydrogen			2,567	0%
diluent loss			4,090	1%
Total			677,703	100%
out				
SCO	85,027	85.0%	496,058	74%
coke			129,264	19%
other			52,381	8%
Total			677,703	101%

Table 4 – Delayed Coking Material Balance

Note: Higher hydrogen requirement is for treating the higher naphtha volumes produced from delayed coking.

Delayed Coking SCO Quality

The SCO produced from the Delayed Coking process is expected to be 29.2 °API with 2.5 wt% sulphur. Nitrogen levels will be 0.06 wt% (600 ppm).

Item	Units	Value
volume	BPD	85,027
API	°API	29.18
Sulphur	wt%	2.45
Nitrogen	wt%	0.06
volume fractions		
C4	lv%	2.56
Naphtha (C5-250 °C)	lv%	29.40
LGO (250-343 °C)	lv%	37.77
VGO (343 - 524 °C)	lv%	30.27
Resid (524 °C+)	lv%	0.00
		100.00

Table 5 – Delayed Coking SCO Product

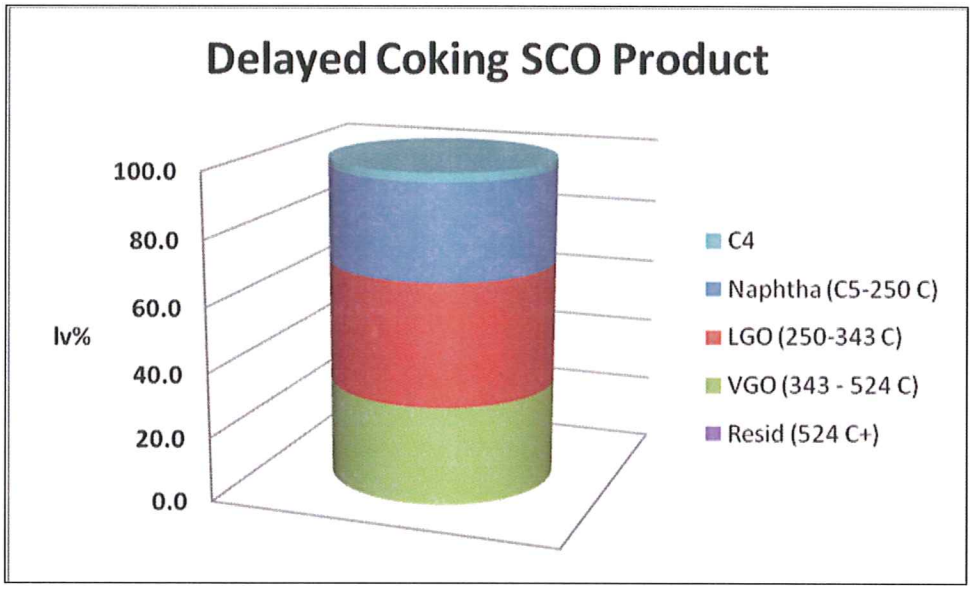


Figure 4 – Delayed Coking SCO Product

Observation: SCO is bottomless

Utility and Cost Estimate

The utility balances for this configuration are included as Attachment 4.

An order of magnitude capital cost estimate has been produced for this configuration.

Delayed Coking

CAPITAL COSTS	Delayed Coking
Capital Cost (*MM$*) *note 1	4,092
$/bbl Bitumen per day	40,917
Utility Estimates	
Variable Cost (*Per bbl Bitumen*)	
Make-up Water (*lb*)	158
Make-up Diluent (*bbl*)	0.00925
CO_2 Emission (*lb*)	56.4
Catalysts & Chemicals ($)	0.04
Fixed Cost (*k$/day*)	346

*note 1 - $ Canadian, remote site location

Table 6 – Capital Cost and Utility Summary

Process Comparison

In a field upgrading application for processing 100,000 BPD of bitumen, the WRITE process is compared with a delayed coking based complex and summarized in Table 7.

SCO Product	Units	1 WRITE	2 DC	2-1 difference	comment
Volume	BPD	84,795	85,027	232.0	comparable
Gravity	API	24.7	29.18	4.5	lower gravity for WRITE
Sulphur	wt%	2.52	2.45	-0.1	higher sulphur for WRITE
SCO Composition					
C_4	lv%	2.6	2.56	0.0	
Naphtha [C_5 – 250 °C]	lv%	20.1	29.4	9.3	less naphtha for WRITE
LGO [250 °C – 343 °C]	lv%	28.6	37.77	9.2	less LGO for WRITE
VGO [343 °C – 524 °C]	lv%	45.9	30.27	-15.6	more VGO for WRITE
Resid [524 °C+]	lv%	2.8	0	-2.8	resid portion remains in WRITE
SIDE PRODUCTS					
Coke/Residue	tonnes/day	2,453	3,102	649.0	less coke for WRITE
Sulphur	tonnes/day	303	252	-51.0	
Fuel Gas	MMBTU/hr	234	385	151.0	
CO_2	tonnes/day	232	344	112.0	lower CO2

Table 7 – Comparison of WRITE vs Delayed Coking

Observation: Based on initial steady state thermal cracking operation and unsteady state pyrolyzer testing, the associated results imply that the WRITE process will produce comparable yields SCO product to delayed coking. This SCO product is heavier (lower gravity 24.7 °API vs 29.2 °API). The WRITE process produces less coke.

There is an expectation that the liquid yield for the WRITE process of up to 88 lv% will be achievable with a properly sized, controllable pyrolyzer.

Product Yield Comparison

On a mass basis, the following figure shows that the WRITE process produces more SCO product and less coke.

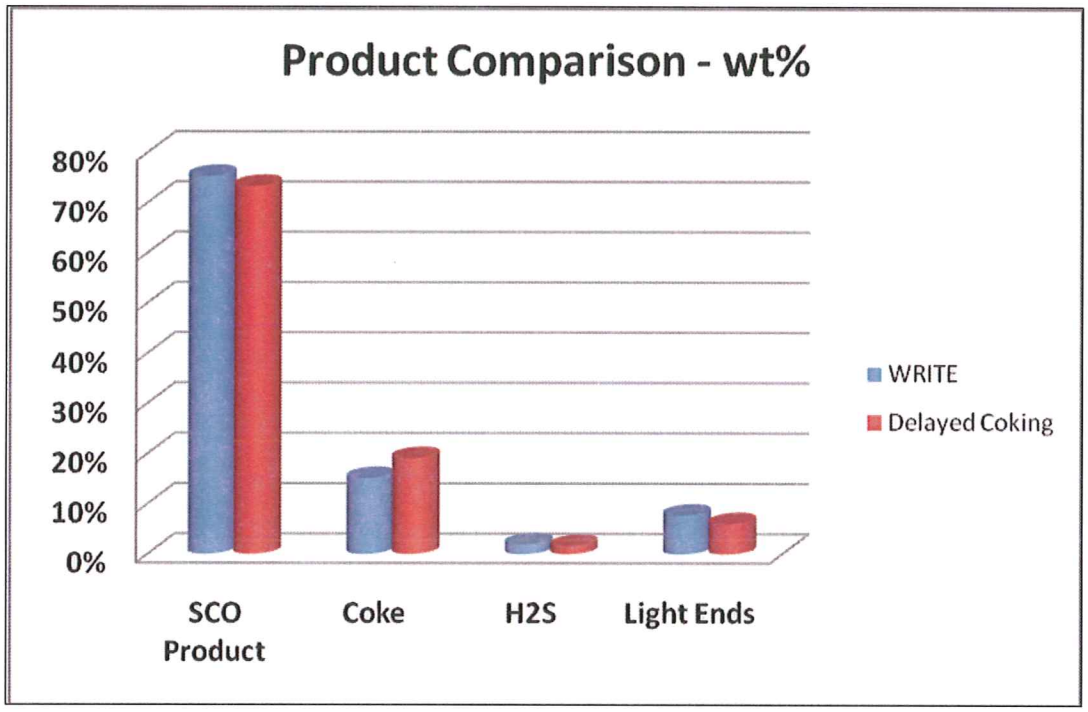


Figure 7 – Product Comparision (mass basis)

Observation: The WRITE process produces more SCO product and less coke on a mass basis.

SCO Product Comparison

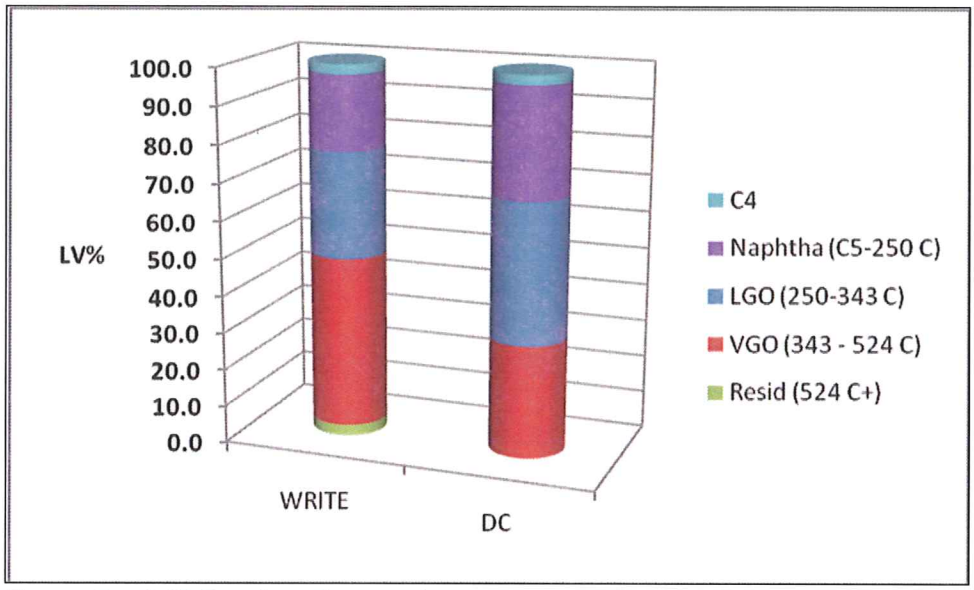


Figure 8 – SCO Product Comparison

Observation: WRITE SCO product has a different product distribution than SCO delayed coking SCO. The WRITE SCO has higher volumes of VGO and Resid.

Capital Cost and Utility Estimate Comparison

Comparison of the anticipated capital cost of a field upgrading complex has been estimated using curve type methods (Class V, +/- 50%) for a remote Canadian location, in Q3 2008 dollars.

Utility balances have been produced for both upgrading complex configurations. These utility balances provide the basis for operating cost estimates.

Basis: 100,000 BPD bitumen

	1	2	2-1	
CAPITAL COSTS	**Delayed Coking**	**WRITE**	**difference**	**Comment**
Capital Cost (*MM$*)	4,092	3,123	-969	
$C /bbl Bitumen per day	40,917	31,226	-24%	WRITE is cheaper
Utility Estimates				
Variable Cost (*Per bbl Bitumen*)				
Make-up Water (*lb*)	158	127	-31	WRITE has lower water use
Make-up Diluent (*bbl*)	0.00925	0.00925	0	
CO_2 Emission (*lb*)	56.4	43.4	-13	WRITE has lower CO2
Catalysts & Chemicals ($)	0.040	0.046	0.006	
Fixed Cost (*k$/day*)	346	264	-82	WRITE has lower fixed costs

Table 8 – Capital Cost and Utility Estimate Comparison

Observation: Lower capital and fixed costs for the WRITE process. WRITE process has lower water use and CO2 production.

Qualitative Comparison of Overall Perfromance

The following table highlights the significant differences between Delayed Coking and the WRITE process in a field upgrading application.

On a per barrel of bitumen processed basis:

Item	Delayed Coking	WRITE Process	Comment
Yield of SCO	Base	Same	Expect WRITE to be 3 lv% higher. Requires steady-state pyrolyzer operation.
Gravity	29.2	24.7	Delayed Coking SCO is lighter
Product Quality	Base	Heavier, higher sulphur	Heavier, higher sulphur SCO product from WRITE
Coke Make	Base	Base – 20 wt%	Lower coke make with WRITE
Fuel Gas Make	Base	Base – 40%	Lower intensity cracking with WRITE
CO_2 production	Base	Base – 30 wt%	Lower CO_2 production with WRITE Is an advantage
Make – Up Water	Base	Base – 20 wt%	Lower make-up water use for WRITE
Capital Cost	Base	Base – 24%	Lower capital costs for WRITE
Fixed Cost	Base	Base – 24%	Less manpower for WRITE
Technical Risk	low	medium	The pyrolysis unit of the WRITE process is unproven
other	commercial	Not commercial	Delayed coking is industry proven

Table 9 – Comparison: Delayed Coking vs WRITE

Summary

Based on the findings of this preliminary assessment, the WRITE process is able to convert bitumen into SCO product that meets pipeline gravity specification. The quality of the WRITE product produced is lower gravity with a different product distribution than that of Delayed Coking. The WRITE process applied to bitumen upgrading is technically feasible.

The performance of the WRITE process offers encouraging results with less coke make per barrel of feed processed while achieving, at a minimum, comparable liquid yields to delayed coking. The expected WRITE performance is +3 lv% SCO yield benefit compared to delayed coking.

As the WRITE process has not been commercialized, insufficient information is available to be able to do a more accurate estimate of capital or operating costs. Further development of the process design basis is necessary for better capital cost estimation.

There are some favourable process indicators suggesting economic attractiveness of the WRITE process, such as potential yield increase, lower utility consumption and lower capital costs. A full economic evaluation has yet to be completed.

Application of the WRITE process in commercial field upgrading (20,000 BPD to 60,000 BPD) will be challenged by the disposition of by-products. Effective use of coke by-products as fuel for SAGD steam production will require significant additional investment in handling, combustion and flue gas clean up. Both the WRITE and delayed coking have the same coke disposal issues.

Considerations for carbon capture from combustion or gasification processing of coke from either process will need to be addressed in light of the current environmental performance expectations.

GO/NO GO Position

There are process and economic aspects of the WRITE process which offer potential benefits in a field upgrading application compared to delayed coking technology.

The lower process severity of the WRITE process provides comparable yields of SCO to delayed coking, while producing less coke by-products. The lower capital costs for the WRITE process imply favourable economics. Steady state pyrolyzer expectation needs to be achieved to confirm liquid yields. This can now be done with the properly sized test unit which now exists.

However, the underlying assumptions of the WRITE process yields and performance require confirmation and validation using MEG SAGD bitumen feedstock. Scale-up from bench scale (1 BPD) to pilot scale (5 BPD), with a comprehensive experimental and analytical program, will confirm/refute process yield and quality estimates.

Commercialization of the WRITE process requires more work to improve the accuracy of the capital and operating cost estimates. Pilot scale operation of the WRITE process will provide better information to develop this more detailed capital cost estimate. There are additional advantages related to continuous operation of the WRITE process compared to the semi-batch operation of a delayed coker.

Within the WRITE process, there is a level of process uncertainty with the performance/cost/reliability of the pyrolyzer (continuous coker) since steady state operations could not be reached with the original equipment used in the bench scale testing programs. The thermal cracking process (Distillate Recovery unit) is well understood and poses a low process risk. There is a strong level of interest in continuing the experimental program for the Distillate Recovery Unit.

The key to the successful application of a carbon rejection process is an effective disposal route for the coke by-product. There are currently no coke disposal outlets accessible to the MEG SAGD production site. Pilot plant operation of the WRITE process would produce sufficient volumes of pyrolysis coke for alternate fuels testing in combustion or gasification which can generate "clean energy" to support SAGD steam production with CO2 sequestration potential.

MEG supports the continued development of the WRITE process to the 5 BPD pilot plant level. The goal of continued WRITE process development would be:

- Better definition of the process operating envelope
    - process yields and qualities
- Build improved confidence in the effective operation of the pyrolyzer unit
- Provide the process basis for a factored equipment based capital cost estimate with Class IV accuracy (+30/-15%)

Development Stages

The development stages for the WRITE process in a field upgrading application would follow a staged development path similar to Figure 9.

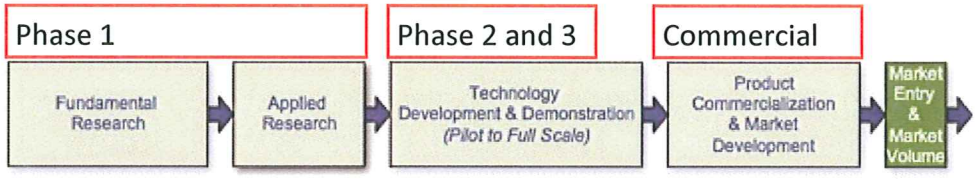


Figure 9 – Technology Development Path

The initial research of the WRITE process has been conducted at the Western Research Institute in collaboration with the Canadian National Centre for Upgrading Technology (NCUT). Phase 1 of the development program has established to "proof of concept" of the WRITE process from a bench scale (1 BPD) operation on Canadian sourced bitumen.

Information produced from Phase 1 has provided the process design basis for the 5 BPD pilot plant for the integrated thermal cracking (distillate recovery unit), and pyrolyzer (continuous coking unit) process elements of the WRITE process.

Phase 2 of the experimental development program is the operation of the 5 BPD pilot plant. Initial operation of the Phase 2 program has concentrated on the process optimization of the mild thermal cracking process (distillate recovery unit). Integrated operation of this process with the pyrolyzer is required to assess the WRITE process as an integrated unit.

Subsequent technology development will be to scale up the 5 BPD pilot plant to a suitable sized field demonstration pilot. The size of the field demonstration unit may range from 300 BPD to 1500 BPD, subject to a rigorous technical risk assessment.

A commercial scale field upgrading facility can range in size from 10,000 BPD to 60,000 BPD.

MEG has plans for a field demonstration pilot of the thermal cracking portion (Distillate Recovery Unit) of the WRITE process. Inclusion of the pyrolyzer is subject to the results of the Phase 2 program coupled with a technology risk assessment.

A detailed design of a 300 BPD Distillate Recovery Unit has been completed.

See Attachment 5 for an illustration of the development schedule for a commercial field upgrading complex.

Appendices:

1. Utility Summary Basis
2. Capital Cost Estimate Basis

Attachments:

1. WRITE process block flow diagram (BFD)
2. Delayed Coking block flow diagram
3. WRITE process utility balance
4. Delayed coking utility balance
5. Commercial Development Schedule

Appendix 1 - Utility Summary Basis

Utility Summary

The upgrader utility balances first satisfy the needs of the upgrading complex. Excess energy is available for integration with the Steam Assisted Gravity Drainage (SAGD) facility or for export.

Utility Units

Utility units could be some or all of the followings

- Sulphur Plant
- Amine Absorption and Regeneration
- Sour Water Stripper
- Cooling Towers
- Power Plant
- Deaerators (water treatment)
- Demineralization Plant (water treatment)
- Steam Letdowns
- Steam Turbine Generators
- Process Boilers

Technology

Conventional process technology estimates are based on reliable in-house sources of performance data. Non commercial processes are based on data provided by licensors or client.

Appendix 2 – Capital Cost Estimate Basis

Estimate Accuracy:
Class V (curve type) capital cost estimate +/-50%.

Capital Cost Basis:
US Gulf Coast (USGC) cost with factors below was used as the basis of cost estimates.

- Location Factor: Field (Athabasca) 2.0 and Edmonton 1.55
- Inflation Rate: 4% annually
- Exchange Rate: US$/CAD$ 1.1

Inside Battery Limit (ISBL)
ISBL cost estimates were based on in-house references, literature data, and licensor's information.

Outside Battery Limit (OSBL or U&O)
ISBL cost estimates with factors below were used to estimate OSBL costs except some items of OSBL were defined separately to allow adjustments in the future when further information becomes available.

- Storage Tanks: 7.0% of ISBL
- Steam System: 4.5% of ISBL
- Cooling Water System: 0.5% of ISBL
- IT Infrastructure not considered

And ISBL cost estimates with a single factor of 30% were used for a lump of all other facilities which may include but is not limit to the following:

- Electricity
- Fuel Oil and Fuel Gas
- Water Supply and Treatment
- Plant Air
- Inert Gas
- Fire Protection
- Flare, Drain and Waste Containment
- Plant Communication
- Product and Additives Blending
- Material Unloading and Loading
- Fences
- Railroads, Roads and Walks
- Buildings
- Vehicles

Other than ISBL and OSBL

Total Installed Cost (TIC) with factors below was used as the basis.

- Offices and Accommodation: 12.0% of TIC
- Owner Cost: 4.0% of TIC
- Logistics: 2.0% of TIC
- Start-up: 2.0% of TIC
- Initial Catalyst and Chemicals Load: 1.0% of TIC
- Spare Parts: 1.0% of TIC
- License/Royalty and Land Permit not considered

Contingency

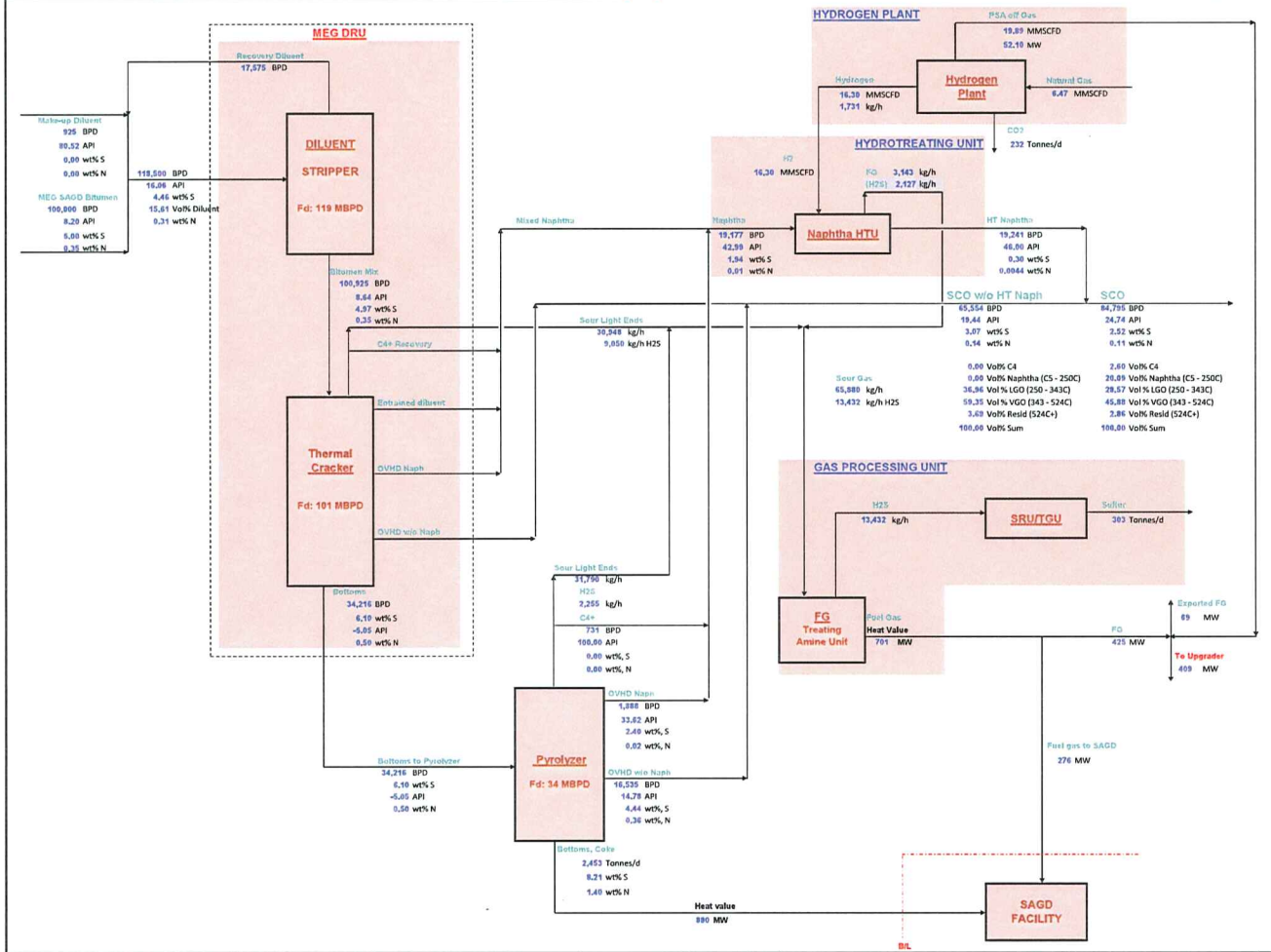
A contingency of 25% of Total Installed Cost (TIC) was used to cover expected uncertainties. Major uncertainties include inaccuracies of cost data applied for specific Cases and lack of complete definitions required.

WRITE Process Block Flow Diagram

Attachment 1.

MEG Upgrader

Case 2a: MEG DRU + Pyrolyzer, Bottoms to fuel SAGD



Stream		MEG DRU Bit	MEG DRU Overhead	MEG DRU Bottom	Pyrolyzer Overhead	Pyrolyzer Coke	Naph HTU Nap	Naph HTU HT Nap
RateV	bpsd	100,000	14,161	34,216	18,423	-	19,177	19,241
RateW	kg/h	671,046	79,298	253,655	116,682	102,223	101,029	101,617
SG		1.013	0.845	1.119	0.956	-	0.8109	0.7972
API		8.20	35.91	-5.05	16.51	-	42.99	46
Sulfur	wt%	5.00	2.20	6.10	4.25	8.21	1.94	0.30
Cetane	index							
Flash Pt	F							
CS at 122F	cs							
CS at 210F	cs							
Nitrogen	wtppm	3,500	160	5,000	3,300	14,000	144	44
Nickel	wtppm	75	0	200	0	660	0	0
Vanadium	wtppm	210	0	525	0	1110	0	0
CS Insol	wt%	16.0						
Concarbon	wt%	13.5						
Aromatics	wt%							

Upgrader Mass Balance

Into Upgrader	kg/h	Tonnes/d
• Bitumen	671,046	16,105
• Diluent	4,090	98
• Hydrogen	1,731	42
Subtotal	**676,867**	**16,245**
Leaving Upgrader	**kg/h**	**Tonnes/d**
• SCO	508,764	12,210
• Diluent		
• H2S	13,432	322
• Pyrolyzer Coke	102,223	2,453
• Light Ends	52,448	1,259
• Others (NH3, H2O)		
Subtotal	**676,867**	**16,245**
Mass Balance	**100%**	**100%**

Heat Balance
Bottom to SAGD

Required Heat	MW
• Natural gas	1,157
Subtotal	**1,157**
Provided Heat	**MW**
• Pyrolyzer Coke	880
• Fuel gas	276
Subtotal	**1,157**
Heat Balance	**100%**

Issues & Considerations

Pros:
- Middle level field upgrading, sour SCO meets pipeline transportation requirement
- Alternate fuel is balanced with coke and fuel gas from upgrader

Cons:
- Extra fuel gas to be balanced

Potential Optimization:
- Fuel gas as feedstock to hydrogen plant

NOTES
- 95% Diluent recovery
- High thermal cracking conversion

REFERENCES DWG.'S

RECORD	
DATE	2008-08-21
ENGINEER	CZ
CHK'D	CM
APP'D	DS
PROJECT#	2007083-1
SCALE	

MEG Energy Corp.

Triumph GROUP

MEG ENERGY COMMERCIAL UPGRADER STUDY

BLOCK FLOW DIAGRAM
MATERIAL AND HEAT BALANCE

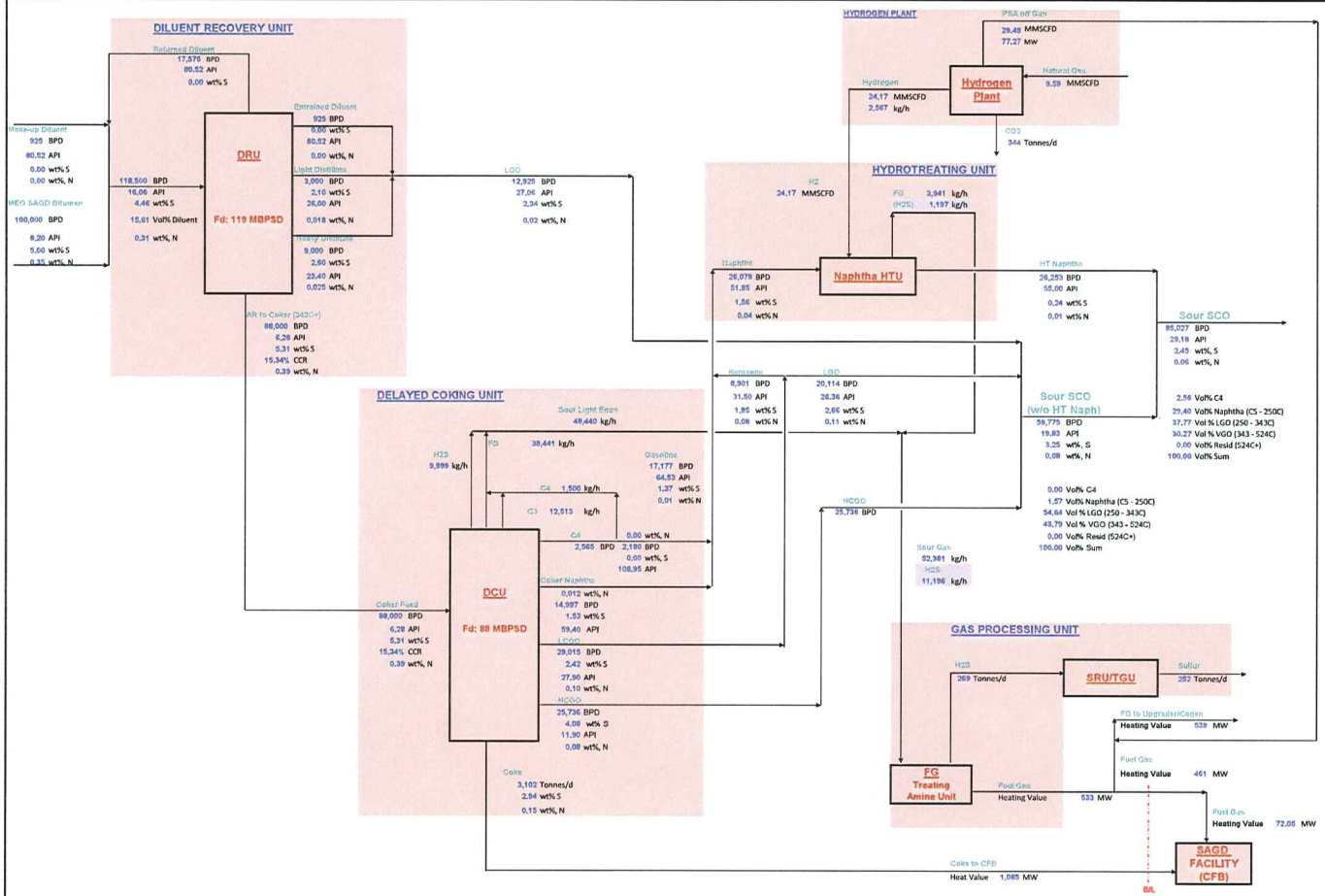
UPDATE	DW DWG NO.	REV
2008-08-21		2

Delayed Coking Block Flow Diagram

Attachment 2.

MEG Upgrader

Case 4: Delayed Coking, Coke to CFB



Stream		DRU Unit			
		Bit	L Dist	H Dist	AR
RateV	bpsd	100,000	3,000	9,000	88,000
RateW	kg/h	671,046	17,856	54,467	598,723
SG		1.013	0.898	0.914	1.027
API		8.20	26.00	23.40	6.28
Sulfur	wt%	5.00	2.10	2.60	5.31
Cetane	Index				
Flash Pt	F				
C5 at 122F	cs				
C5 at 210F	cs				
Nitrogen	wtppm	3,500	180	250	3,895
Nickel	wtppm	5			101
Vanadium	wtppm	210			308
C5 Insol	wt%	16			17.9
Concarbon	wt%	14			15.3
Aromatics	wt%				

Stream		Coker Unit				Naphtha HTU	
		Coker Nap	LCGO	HCGO	Coke	Nap	HT Nap
RateV	bpsd	14,997	29,015	25,736		26,078	26,253
RateW	kg/h	73,462	170,636	168,241	129,264	133,333	131,959
SG		0.739	0.888	0.987		0.772	0.759
API		59.4	27.9	11.9		51.85	55.00
Sulfur	wt%	1.53	2.42	4.08	2.94	1.56	0.24
Cetane	Index						
Flash Pt	F						
C5 at 122F	cs						
C5 at 210F	cs						
Nitrogen	ppm	122	1,010	800	1,500	375	0
Nickel	ppm						
Vanadium	ppm						
Concarbon	wt%						
RON							
MON							
R+M/2							
Olefins	wt%						
Naphthene	wt%						
Aromatics	wt%						

Upgrader Mass Balance

Into Upgrader	kg/h	Tonnes/d
• Bitumen	671,046	16,105
• Diluent	4,090	98
• Hydrogen	2,567	62
Subtotal	677,703	16,265
Leaving Upgrader	**kg/h**	**Tonnes/d**
• SCO	496,058	11,905
• Naphtha		
• Diluent		
• H2S	11,196	269
• Coke	129,264	3,102
• Light Ends (w/o H2S)	41,185	988
• Others (NH3, H2O)		
Subtotal	677,703	16,265
Mass Balance	100%	100%

Heat Balance
Coke to CFB

Required Heat	MW
• Natural gas	1,157
Subtotal	1,157
Provided Heat	**MW**
• Coke	1,085
• Fuel Gas	72.05
Subtotal	1,157
Heat Balance	100%

Issues & Considerations

Pros:
- Middle leve upgrading, good product quality, SCO meets pipeline transportation requirement
- Alternate fuel is sufficient to feed SAGD, use all coke produced

Cons:
- Extra fuel gas to be balanced

NOTES	REFERENCES DWG.'S	RECORD	
• 95% Diluent recovery		DATE	2008-08-21
		ENGINEER	CZ
		CHK'D	CM
		APP'D	DS
		PROJECT#	2007083-1
		SCALE	

MEG Energy Corp.

Triumph EPCM

MEG ENERGY COMMERCIAL UPGRADER STUDY

BLOCK FLOW DIAGRAM
MATERIAL AND HEAT BALANCE

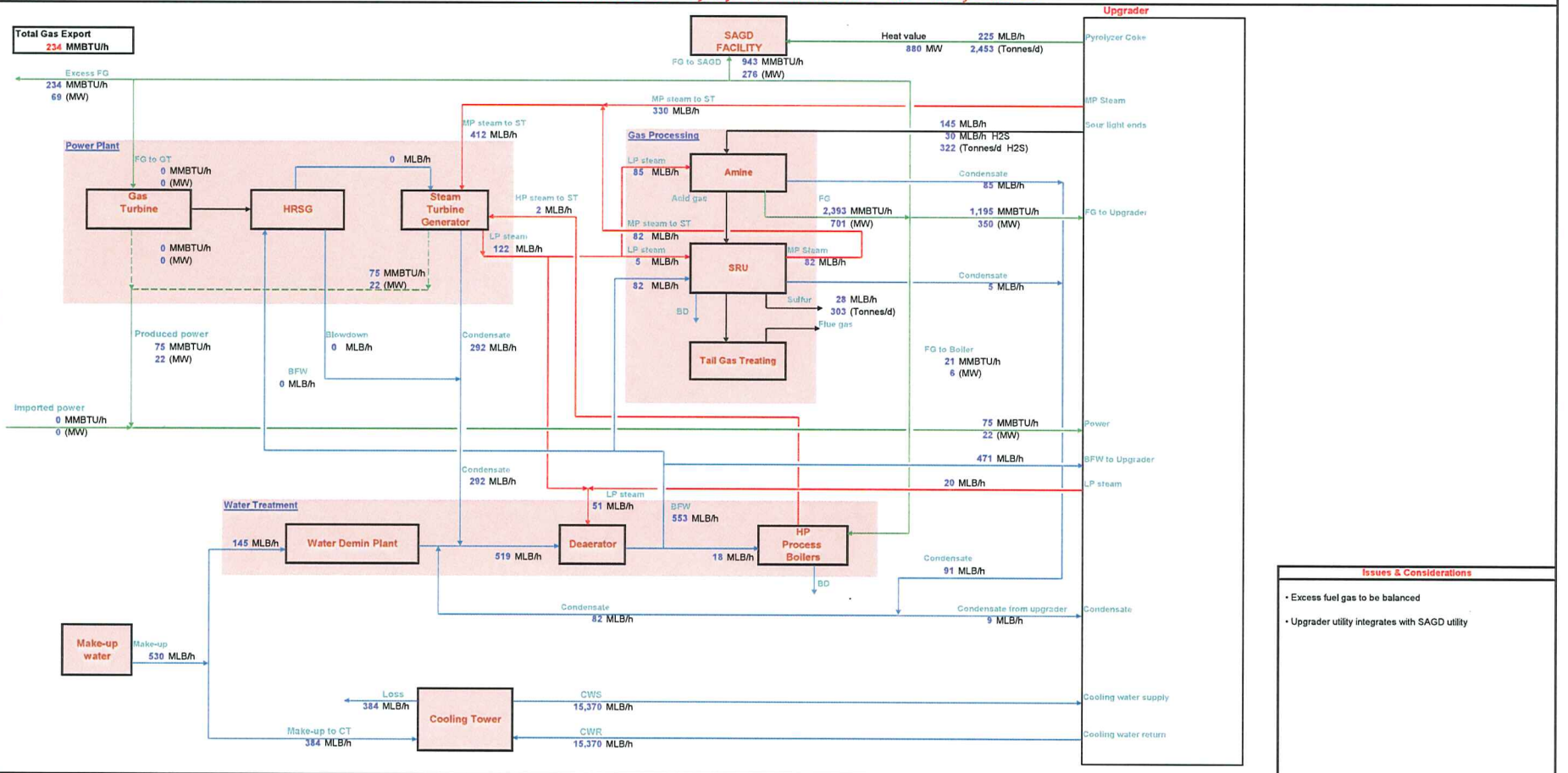
UPDATE	DWG NO.	REV
2008-08-21		2

WRITE Process - Utility Summary

Attachment 3.

MEG Upgrader

Case 2a: MEG DRU + Pyrolyzer, Pitch to Fuel SAGD - Utility Balance



Issues & Considerations

- Excess fuel gas to be balanced
- Upgrader utility integrates with SAGD utility

NOTES	REFERENCES DWG.'S	RECORDS	
• Lines instruction: Water-Blue, Steam-Red, Fuel-Green, others-Black • PSA offgas will be consumed directly in Upgrader	BFD for Case 2a	DATE	2008-08-27
		ENGINEER	CZ
		CHK'D	CM
		APP'D	DS
		PROJECT#	2007083-1
		SCALE	

MEG Energy Corp.

Triumph

MEG ENERGY COMMERCIAL UPGRADER STUDY

UTILITY BLOCK FLOW DIAGRAM

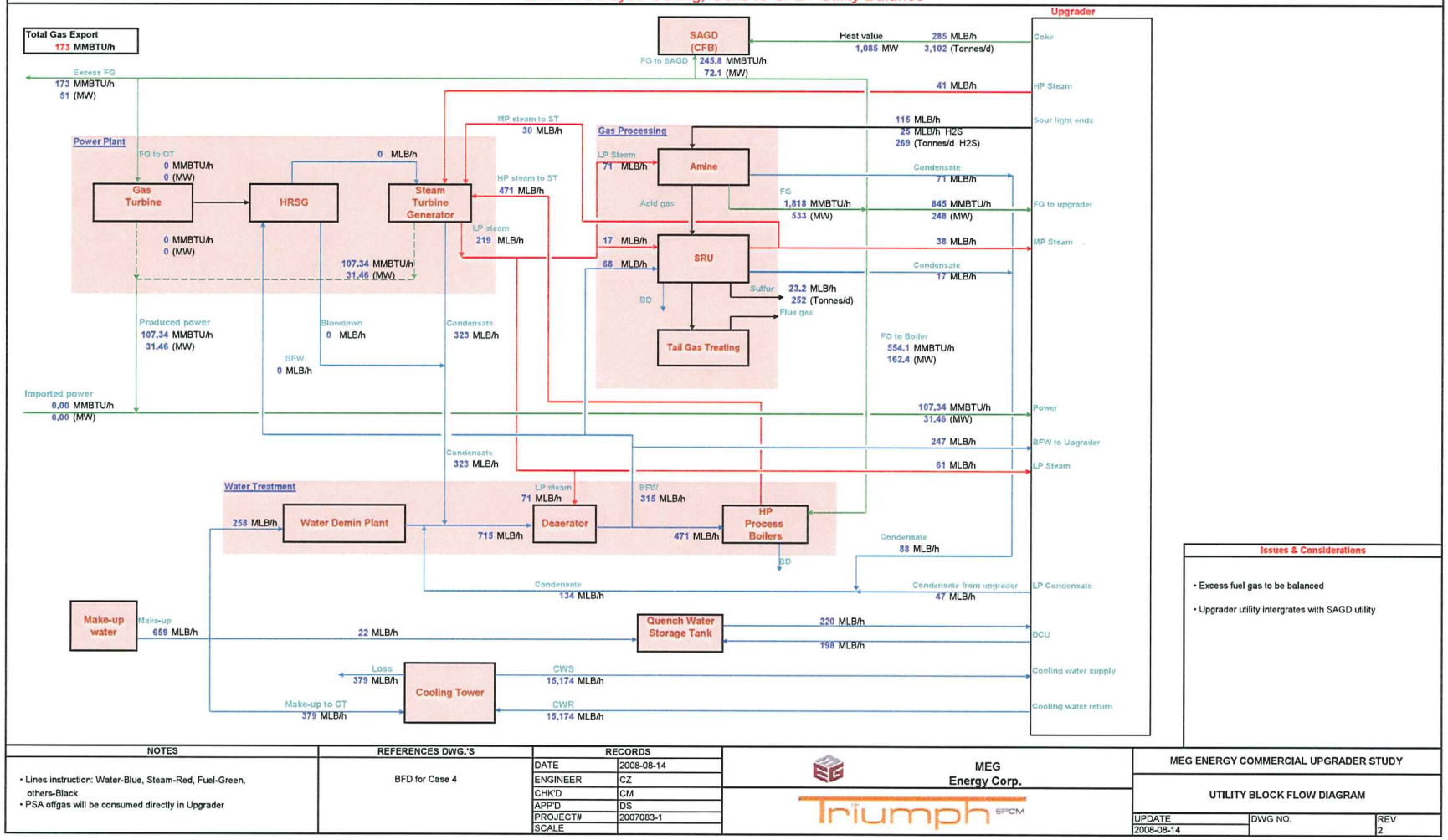
UPDATE	DWG NO.	REV
2008-08-27		2

Delayed Coking Utility Balance

Attachment 4.

MEG Upgrader

Case 4: Delayed Coking, Coke to CFB - Utility Balance



NOTES	REFERENCES DWG.'S	RECORDS			
• Lines instruction: Water-Blue, Steam-Red, Fuel-Green, others-Black • PSA offgas will be consumed directly in Upgrader	BFD for Case 4	DATE	2008-08-14	MEG Energy Corp.	MEG ENERGY COMMERCIAL UPGRADER STUDY
		ENGINEER	CZ		UTILITY BLOCK FLOW DIAGRAM
		CHK'D	CM	Triumph EPCM	
		APP'D	DS		
		PROJECT#	2007083-1		
		SCALE			

UPDATE	DWG NO.	REV
2008-08-14		2

Attachment 5 - Field Upgrading Development Schedule