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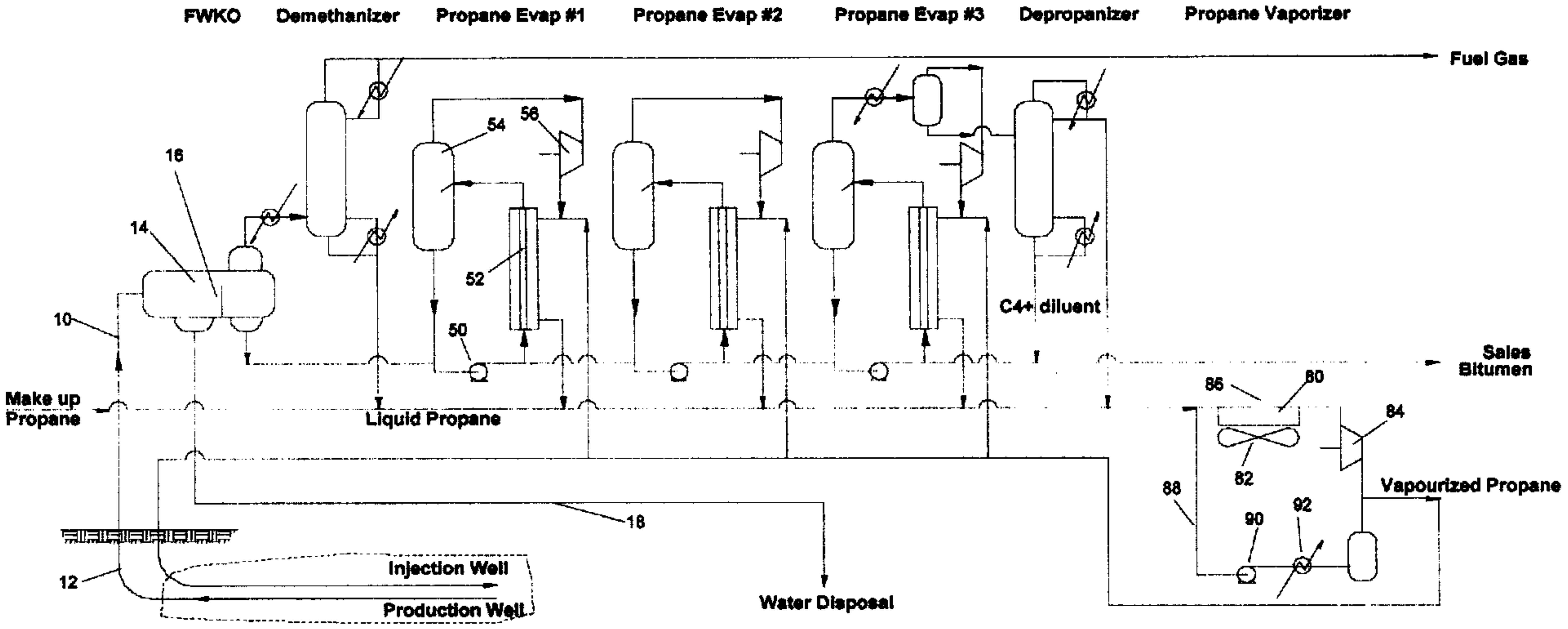
(54) Titre : METHODE ET APPAREIL A FAIBLE CONSOMMATION D'ENERGIE POUR LA STIMULATION DE LA PRODUCTION D'HUILE LOURDE
(54) Title: ENERGY EFFICIENT METHOD AND APPARATUS FOR STIMULATING HEAVY OIL PRODUCTION



(57) **Abrégé/Abstract:**

A method for extracting oil from a formation is disclosed. The method includes selecting a condensing solvent having a high vapour pressure at low temperatures and injecting the solvent into the formation. Then the solvent is condensed in the formation to deliver a latent heat of condensation to said oil and to mobilize said oil. The oil is then produced, and solvent is recovered from the produced oil. The recovered solvent is then conditioned for reinjection by being heated by means of a heat pump and compressor. Then it is reinjected into the formation. In a preferred aspect the thermal coefficient of production is greater than 1. An oil recovery system for carrying out the process is also disclosed.

ABSTRACT

A method for extracting oil from a formation is disclosed. The method includes selecting a condensing solvent having a high vapour pressure at low temperatures and injecting the solvent into the formation. Then the solvent is condensed in the formation to deliver a latent heat of condensation to said oil and to mobilize said oil. The oil is then produced, and solvent is recovered from the produced oil. The recovered solvent is then conditioned for reinjection by being heated by means of a heat pump and compressor. Then it is reinjected into the formation. In a preferred aspect the thermal coefficient of production is greater than 1. An oil recovery system for carrying out the process is also disclosed.

Title: Energy Efficient Method and Apparatus for Stimulating Heavy Oil Production

FIELD OF THE INVENTION

This invention relates generally to the field of enhanced resource recovery, and more particularly to enhanced recovery of heavy oils and bitumen. Most particularly this invention relates to energy utilization in such resource recovery.

BACKGROUND OF THE INVENTION

Currently steam is the dominant thermal fluid used for insitu recovery of bitumen and heavy oil. Steam raises the temperature of the bitumen thereby reducing its viscosity and allowing it to flow more easily. Steam extraction is subject to a number of problems including high heat losses, clay swelling problems, thief zones, water-oil emulsions, capillary surface tension effects, lack of confinement for shallower zones and disposal of large quantities of environmentally damaging salt and organic acids as a consequence of boiler feed water purity requirements. With the best available technologies, only 10% of the original bitumen resource is economic to extract.

Thermal recovery processes, using steam, require large amounts of fuel to be burned to produce the steam and can emit enormous amounts of greenhouse gases such as carbon dioxide. Estimates published by Natural Resources Canada[1] show CO2 emissions of about 70kg/bbl for bitumen production and a total of about 120 kg/bbl for synthetic crude (i.e. upgraded bitumen usually derived from surface mined bitumen).

Alternative sources of steam such as co-generation may offer some synergies with lower "unit" emissions (per kWH or per kg steam). However, co-generation also produces higher gross

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[1] Canada's Emissions Outlook: an Update, December 1999, Annex B, pg B-6, Available at www.nrcan.gc.ca/es/ceo/update.htm

emissions than direct steam generation, due to the fact that additional fuel is burnt to generate electricity. From an environmental perspective, higher total emissions means more environmental impact (even if the unit emissions are reduced).

Recent estimates released by the Alberta Energy Utilities Board[2] and the Canadian Association of Petroleum Producers[3], predict that bitumen (and synthetic crude) production rates will be 2 to 2.6 million bbl/day of bitumen by 2010. This level of bitumen production will produce at least 140 million kilograms (=70x2million) of CO2 emissions per day (i.e. 300,000,000 to 700,000,000lbs CO2 per day depending on the proportion of insitu vs synthetic crude production).

With the recent spike in natural gas prices many oil companies are looking for alternative fuels which are less costly. A recent patent application 2,332,685 proposes to deasphalt the SAGD bitumen in surface facilities and then burn the asphaltene residue to generate steam. This patent application teaches substitution of a clean burning fuel (natural gas) for a less expensive asphaltene residue which is high in carbon, sulphur (>8%) and toxic metals. 2,332,685 would double the CO2 emissions for SAGD bitumen production as compared to a process burning natural gas. While this process may have economic benefits, the benefits come at a cost of substantially higher emissions into the environment.

What is needed therefore a way to reduce the energy requirement for bitumen extraction.

BRIEF SUMMARY OF THE INVENTION

The present invention is directed to a process and apparatus to reduce energy requirements for the production of

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[2] Alberta's Reserves 2000 and Supply/Demand outlook 2001-2010, Alberta Energy Utilities Board

[3] Canada's Oil Sands Development delivered by Eric Newell, Chairman & CEO, Syncrude Canada. Available at http://www.capp.ca/

heavy oil and bitumen by solvent extraction. This process also simultaneously upgrades the heavy oil and bitumen insitu. In one aspect of the invention a heat pump is used to recover heat from a latent or free heat source to reduce the energy load required to prepare the solvent for injection and extraction. In another aspect heat pumps are used to recover the solvent from the produced fluids to again reduce direct energy consumption and to reduce emissions. In another aspect of the invention a production apparatus is provided which is capable of utilizing the energy efficient and pollution reducing processes of the present invention.

As outlined in more detail below, the present invention is predicted to reduce the amount of energy required for extraction by 90-95% percent relative to the best prior art insitu technology, i.e., steam assisted gravity drainage (SAGD). By reducing energy costs, a primary expense, the present invention also increases in the value of the production, improving both the operating and profit margins. Most importantly, the process and apparatus of the present invention is predicted to reduce the amount of emissions by 90-95% relative to the prior art SAGD.

The overall energy required by the extraction process is determined by the need to heat the oil/bitumen reservoir to the extraction temperature. The present invention is most preferably used in conjunction with an extraction process in which the bitumen extraction temperatures are provided by using a condensing solvent such as propane. By utilizing such a condensing solvent and the latent heat of condensation, the energy requirement for extraction can be reduced thereby reducing consequent emissions over SAGD methods.

For example if the original reservoir temperature is 8C and the solvent extraction process operates at a temperature of 50C (vs 230C for a steam extraction process) then the expected energy savings of the solvent extraction process would be 81%

(=(230-50)/(230-8)) relative to current steam extraction technology (SAGD). The present invention delivers additional energy and hence emissions savings over and above this level as explained in more detail below.

The heat for bitumen extraction is supplied by the latent heat of condensation of the solvent. The present invention is comprehends choosing the solvent such as propane, which has an appropriate boiling point of -42C. In this way much of the required latent heat of vaporization can be supplied via heat exchange with “free” or latent sources of heat such as ambient air or water, or even waste heat from flue gas or the like. The vaporized solvent is subsequently compressed, thereby raising its temperature and pressure to the desired extraction conditions. Thus, the present invention comprehends a hydrocarbon (i.e. heavy oil) extraction process in which the solvent absorbs most of its heat energy at low temperature and then a small amount of compression energy is used to raise the solvent condensation temperature to the preferred extraction temperature analogous to a heat pump or a Carnot refrigerator. Thus, for a solvent extraction process operating at 50C, with absorption of heat from a latent heat source at 5C according to the present invention (i.e. an arbitrary but “reasonable” heat source temperature), the thermodynamic coefficient of performance is 6.2 (=278K/(323K-278K)). This means that according to the methods and apparatuses of the present invention one joule of work in the compressor can deliver 6.2 joules of heat at 50C.

Therefore according to one aspect of the present invention there is provided a method for extracting oil from a formation, said method comprising:

Selecting a condensing solvent having a high vapour pressure at low temperatures,

Injecting said solvent as a vapour into said formation,

Condensing said solvent in said formation to deliver a latent heat of condensation to said oil and to mobilize said oil,

Producing said oil and said solvent;

Recovering said solvent from said produced oil,

Heating said solvent means of a heat pump, and

Re-injecting said solvent into said formation.

According to another aspect of the present invention there is provided a method of stripping a solvent from produced oil and preparing the solvent for re-injection, said method comprises the steps of:

De-watering said produced oil;

Separating said produced oil into a gas fraction and a liquid fraction;

Removing non-condensibles from said gas fraction;

Passing said liquids through one or more evaporation steps to recover additional solvent; and

Vaporizing said recovered solvent.

According to yet another aspect of the present invention there is provided an oil recovery system comprising:

a source of solvent for injection into a hydrocarbon bearing formation;

an injection path into an oil bearing formation;

a recovery path out of said oil bearing formation; and

a solvent recovery and reheating apparatus including at least one heat pump.

BRIEF DESCRIPTION OF THE DRAWINGS

Reference will now be made, by way of example only, to preferred embodiments of the invention as illustrated in the accompanying drawings and in which:

Fig. 1 shows a process flow diagram for one embodiment of the present invention including solvent recovery, purification and solvent vaporization.

Fig. 2 shows a simplified Molliére diagram for propane, which identifies two different paths to provide hot, high pressure propane vapour for insitu bitumen extraction, to illustrate the thermodynamic energy requirements for the present invention as compared to a direct fired vaporization process of the prior art.

Fig. 3 shows the relative energy efficiency/ reduction in energy cost/reduction in greenhouse gas emissions for the present invention compared to a steam extraction process (SAGD) as a function of bitumen extraction temperature and heat source temperature.

DETAILED DESCRIPTION OF THE PREFERRED EMBODIMENTS

In this specification the following terms shall have the following meanings. "Heavy oils" refer to crude oils, which have high specific gravity and viscosity and are therefore difficult to extract commercially because they do not readily flow. In this sense heavy oil comprehends bitumen. "Waste heat" means heat from a combustion process which is a byproduct of the process. A "non-condensable" means any gas that is volatile at a temperature and pressure at which the preferred solvent condenses to a liquid. In this sense non-condensable means any fraction that has a lower boiling point temperature than the preferred solvent used in the extraction process. For example, methane is a non-condensable gas in an extraction process utilizing propane as a solvent. A "methane injection specification" means the preferred amount of methane (or more generically a non-condensable) present in a solvent which is being re-injected and is preferably less than 5% and is most preferably about 1 to 2 % or less. The preferred amount of methane permitted depends upon the economics of purification (removing methane from produced fluids) as compared to heavy oil extraction rates (the negative effect such methane has on production rates). A “heat pump” means any device or process by which heat energy is transferred from some pre-existing heat source or reservoir to the solvent to vaporize the solvent (and typically is followed by a compression step to achieve desired injection temperatures and pressures.). Thus heat pump heating is contrasted with having to supply the heat by means of direct energy consumption such as combustion or electrical energy consumption or the like. A “latent heat source” means any free or other heat source or reservoir that is existing but otherwise dormant or undeveloped, including ambient air, geothermal, ambient water or other sources and excludes any heat sources involved in the direct consumption of energy such as combustion heating, electrical resistance heating or the like. A “secondary heat source” or “waste heat source” is any source where heat is generated as a byproduct and otherwise would be wasted if not used to vaporize solvent. “Heavies” means hydrocarbon species with higher molecular weights than the solvent. Thus if the solvent is propane, heavies comprehends C4 and higher species such as butane, pentane and the like. A solvent having a high vapour pressure at low temperatures is any solvent that has a boiling point lower that about 20 degrees C at atmospheric pressure. One preferred solvent is propane. The “thermodynamic coefficient of performance” is defined as the latent heat source temperature (in absolute terms) divided by the temperature rise of the process.

Figure 1 shows a schematic process flow sheet for an apparatus of the present invention. Figure 1 will be used as the basis for a more detailed calculation of the energy requirements of the invention. While Figure 1 shows a particular embodiment of the invention, it will be understood that the present invention comprehends such variations that may be necessary or desirable to provide a greater or lesser energy efficiency at the expense of capital cost, process stability and process control during upset conditions. Figure 1 provides one preferred arrangement of compressors, heat exchangers and pumps to provide a more detailed estimate of process conditions and consequent process energy requirements[4]. As explained in more detail below the process steps include vaporization of the solvent at relatively low pressure (temperature) and subsequent vapour compression, prior to reservoir injection, which reduces the fuel requirements as compared to a direct fired vaporizer.

By way of example consider a moderate size facility with reasonable economies of scale has a bitumen production of 16,000m3/day (100,000bbl/day). The insitu gas/bitumen ratio is assumed to be 2 m3/m3. The insitu gas is mostly methane and typically has less than one mol percent of ethane. If a 55% recovery rate of original bitumen in place (within an insitu extraction chamber) is assumed, then the effective gas bitumen ratio[5] is 3.6 (=2/.55). Where the solvent is propane a preferred reservoir extraction occurs at 50C. As will be understood the solvent has a critical temperature that is higher than the formation extraction temperature. The mean annual air temperature for the Athabasca tar sands is about -6C, but it can be assumed that by using heat storage in the summer and/or a variety of different heat sources (geological etc), an effective annual average heat source can be obtained at 5C. The produced solvent/oil ratio of 4.3 m3/m3 is determined by the heat balance at 50C, so 69,000m3 (of liquid equivalent) propane vapour is injected per day. Approximately 3% propane makeup is required, the other 97% of the injected propane comes from recycled and reconditioned solvent. Propane makeup includes about 1200m3/day is required for voidage replacement in the reservoir, 600m3/day of propane makeup is required due to propane

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[4] For the sake of clarity, the details with respect to location and actuation of process control valves have not been included in Figure 2, because isenthalpic valve operation and actuation don't consume much energy.

[5] In other words, 100% recovery of bitumen associated gas but 55% recovery of the bitumen.

residuals in the sales bitumen and another 150m3/day of makeup is required due to losses in the fuel gas[6]. Thus the total makeup propane required is estimated to be 1950 m3/day.

Since the thermodynamic properties of propane determine much of the energy requirements, the basis for the energy budget in this specification is kilojoules of fuel energy per kilogram of injected propane (kJ/kg). However the energy budget used can easily be converted to a "per bbl of bitumen" basis through the solvent/oil ratio. As will be understood by those skilled in the art, the energy budget used is conservative in that it includes items such as artificial lift for the produced heavy oil that might typically be powered with electricity from the grid. In this manner, a more thorough estimate of the true energy requirement (and the GHG emissions) can be provided.

Figure 1 shows the produced fluids 10 flowing from the production well(s) 12 into the free water knockout (FWKO) 14. The FWKO includes a baffle 16 to separate most of the produced water, which is disposed of through line 18. The well(s) 12 would typically be operated in a control mode to produce liquid and minimize production of solvent vapour.

The process parameters of the present invention depend upon the starting conditions for the process. Thus, the present invention comprehends that produced heavy oil may be either artificially lifted to the surface, such as by means of a pump or simply lifted to the surface by the pressure difference between the well bore and surface pressure. The advantage of a pump is that it is then possible to maintain the pressure in the FWKO to a pressure slightly below the bubble point to minimize the vapour overhead in the FWKO. As well the methane injection specification needs to be maintained as explained in more detail below. Alternatively, if the pressure difference is used to lift the

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[6] Reference 2 predicts 20,000m3/day of propane is available in Alberta to 2010 (i.e. excess of supply above demand)

produced fluids, a portion of the production will be vaporized. The latent heat of vaporization would chill the produced fluids and thus reduce the FWKO temperature. Relying on the pressure differential eliminates the energy costs associated with a pump as well as permitting simpler well bore configurations but is prone to slugging as the well bore will tend to randomly unload. The fuel energy requirement for artificial lift is about 7kJ/kg propane[7].

The produced fluids (heavy oil) include water, solvent, bitumen and solution gas associated with the bitumen in the reservoir. The FWKO drops out the produced water. Such a separation is relatively easy compared to SAGD since the produced fluid viscosity will be less than 100cP and the density difference between the produced water and the produced fluids will be quite large at 0.4g/cc (=1-.6).

Any gas which is present in the FWKO preferably flows to a de-methanizer distillation column or the like. Since about 90 mol% of the hydrocarbon production is propane, the gas/liquid ratio (and temperature) in the FWKO can be controlled by setting the FWKO pressure. Since methane is the most volatile species, it is preferentially concentrated in the overhead gas. Thus, by setting the FWKO to a desired pressure the hydrocarbon liquid stream will meet the methane injection specification while simultaneously restricting the demethanizer size/duty to the smaller proportion of the produced fluids in the overhead vapour. For example, for a 1.3 mol% methane in the produced fluids, the FWKO could be operated at 5 mol% vapour – 95mol% liquid. This provides an overhead vapour with >6-mol% methane while the FWKO bottoms would meet a 1-mol% methane injection specification. To obtain a lower methane specification in the liquids, then a lower pressure (temperature) can be used so that a larger proportion of the produced methane is stripped out via

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[7] This accounts for both the energy efficiency of the driver/pump and lifting the bitumen as well as the propane.

the FWKO overhead gas. For example, a 10 mol% vapour overhead would reduce the residual methane in the FWKO liquid to about 0.7mol%.

In the past it has been taught that extraction solvents preferably include non-condensibles such as methane, ethane, nitrogen or CO2 in fairly large concentrations to achieve target dewpoint specifications. In contrast, the present invention teaches that the non-condensibles such as methane should be minimized in the solvent since even small concentrations of less condensable gases such as methane in the extraction solvent dramatically reduce the energy efficiency of the separation and purification of such solvents. However, methane removal comes at a cost. For example, doubling the FWKO vapour overhead will double the size and duty for the demethanizer and yet only achieves a 30% reduction in the methane concentration in the injection solvent.

The demethanizer overheads will typically contain methane, ethane and a small amount of propane (<5%). The overhead gas can for example be used for fuel. Since propane losses in the demethanizer overhead represent an expensive source of fuel gas (about twice the cost of methane per kJ of heat), propane losses in this way are preferably minimized. Absorber columns and the like may be used to increase the propane recovery.

The demethanizer bottoms will provide near injection spec propane. The demethanizer is a fairly energy intensive separation due to the low temperature of the condenser (<-100C). Thus, restricting the demethanizer feed rate via FWKO pressure control will be helpful for controlling process energy requirements. Furthermore, since the demethanizer condenser operates at cryogenic temperatures, it will be necessary to dehydrate or otherwise prepare the demethanizer feed to avoid ice (water, CO2 or hydrates) buildup. The demethanizer energy requirement is estimated at ~3MW, which is about 8kJ/kg of injected propane.

Referring to Figure 1 again, we see that the FWKO hydrocarbon liquids flow into a series of propane evaporators. Although three are preferred, more or less could be used as desired. The evaporators use vapour recompression with condensation (in other words an internal heat pump arrangement) to supply most of the latent heat of vaporization. This arrangement with internal recycle (vapour recompression) helps to reduce the energy requirement for separation of the propane from the bitumen.

Referring to Figure 1 again, we see that the FWKO hydrocarbon liquids flow into a series of three propane evaporators. Each evaporator includes a fluid pump 50, a heat exchanger 52 and a gas liquid separation vessel 54 and a compresser 56. The gas liquid separator has a fairly large diameter and may include internal features such as baffles and demisters to facilitate gas liquid separation. The large diameter of the separator 54 provides a fairly long residence time for the fluids (i.e., slow superficial gas velocities) to help avoid liquid carryover (i.e., mist or foam) into the gas overhead stream. The overhead solvent vapour is compressed in the compressor 56 to an elevated pressure and temperature. The compressed solvent vapour is then recondensed in the shell side of the heat exchanger 52, releasing its latent heat of condensation to heat and vapourize the bitumen solvent blend. If the temperature difference across the heat exchanger 52 is kept to fairly small values (i.e. 10 to 20C) then the compression horsepower requirement is very small relative to the total latent heat required to vapourize all the propane. For example with a 10C temperature difference across the heat exchangers 52 we can vapourize/recondense 125MW of propane for about 5 MW of compressor power. The liquid draining from the bottom of the separator vessel 54 is pumped though the tube side of the heat exchanger 52, heated and flashed back into the separator vessel 54. A portion of the liquid discharged from the pump 50 is also directed towards the next evaporator stage, while fresh blend fluid is fed to the evaporator from the FWKO to offset both overhead vapour loss + liquid discharge to the next stage.

This arrangement with vapour compression recondensation effectively provides an internal heat pump. The advantage of this internal arrangement is that the process conditions can be set to optimum separation conditions without being affected by availability of external heat sources and potential seasonal variations. However, since the latent heat of condensation of propane (per kg of propane) decreases as the pressure (temperature) increases we must supply a small additional amount of heat energy to close the heat balance for each evaporator. The makeup heat requirement is minimized by keeping the shell-tube temperature difference across the heat exchangers small. Figure 1 shows the makeup heat energy supplied from condensation of process propane vapor.

The latent heat of condensation of propane (per kg of propane) decreases as the pressure (temperature) increases. Thus, a supply of an additional amount of heat energy to close the heat balance for each evaporator is required. Figure 2 shows the makeup heat energy being supplied from the condensation of process propane vapor.

As will be understood by those skilled in the art, it becomes more difficult to extract propane from the bitumen as the propane concentration in the bitumen decreases. This is because the vapor pressure of the residual propane is reduced as the propane concentration in the bitumen is reduced. Therefore, the compression energy (work) requirement (per kg of propane vaporized) increases as the propane residual becomes smaller.

Thus, according to the present invention it is preferred to evaporate the propane in a multistage process, so that a substantial portion of the propane can be vaporized at favorable conditions (i.e. higher vapour pressure). Each successive stage operates at lower pressure and/or higher temperature. It is also desirable to leave some propane residual in the bitumen. It is estimated that about 2 weight percent propane in the bitumen would allow the bitumen to meet typical pipeline viscosity and vapour pressure specifications without the need for additional diluent. This is a useful result, since reference 2 indicates that there will be insufficient diluent supply to meet pipeline bitumen blend requirements after 2005. Of course the present invention comprehends other residual weight percent of propane depending upon the circumstances. In general the most preferred weight percent is one which balances the cost of recovery, and the end properties of the produced fluids.

The energy requirement for the evaporators is largely determined by the compression energy requirement. The compression energy requirement is determined by the condenser heat transfer coefficients and exchanger surface area and the temperature difference needed to supply the latent heat to vaporize the propane in the bitumen. It is believed that an economic optimum might require about 5 MW of compression to provide the entire 125MW of latent heat required to strip the propane from the bitumen for a process sized as outlined before. This level of efficiency is achieved by using a relatively small amount of compression (i.e. temperature difference ~10C across the propane evaporation heat exchangers). Accounting for compressor efficiencies, the propane evaporators represent a fuel requirement of about 30kJ/kg of injected propane.

It can now be appreciated what the benefits of the evaporator configuration of the present invention are. The particular arrangement of Figure 1 allows the propane separation from the produced bitumen to occur at optimal process temperatures, which are effectively detached from the heat source temperature. For example, if a variable heat source (i.e., ambient air) is used to supply the latent heat for vaporization in the evaporators, then seasonal temperature variations could result in very low evaporator temperatures. This could lead to operational problems with high fluid viscosities, low vapour pressures high compression requirements, high gas velocities and potential for foaming and/or mist carryover problems. Obviously this concern would be mitigated if a warm and reliable heat source (geothermal or the like) was available throughout the winter months. Another aspect of the present invention is to deal with the impact of residual non-condensable gases in the FWKO liquids (i.e. the 1 mol% to 0.7 mol% methane discussed above). The residual methane impurities raise the bubblepoint pressure of the compressed vapour and thereby reduce heat transfer efficiency. This means that additional compressor horsepower and/or temperature drop is required to recondense the evaporator vapour overheads.

There are several ways to help mitigate this problem as comprehended by the present invention. One way is to provide a continuous bleed/purge from the condensers (to the demethanizer) to avoid accumulations of trapped methane.

Another way is to take advantage of the fact that since methane is more volatile than the propane; the methane residual in the liquids will decrease in each successive evaporator. Thus, much of the residual methane in the FWKO liquids will be concentrated in the vapour overheads from the first evaporator. If a reliable heat source is available for the first evaporator, then all of the overhead vapour from the first vaporizer could be compressed directly to injection conditions, without recondensation, thereby avoiding the difficulty of recondensing a methane/propane mixture. Whether or not to recondense the vapour from the first evaporator will depend on these considerations but is also comprehended by the present invention.

Figure 1 shows that the vapour overheads from the final evaporator go through a partial condenser and the liquids then go to a depropanizer column. The depropanizer provides a means to recover heavies (C4+) from the solvent vapour and to use this fraction as an additional diluent for the bitumen. While this separation is shown for completeness, in most cases it is expected that the C4+ fraction in the bitumen/solution gas may be too small to be worth recovering. This C4+ concentration in the reservoir gas may vary depending on the particular geographical location. For the purposes of the energy budget calculations contained herein, it is assumed that this C4+ separation is unnecessary. Thus, the propane evaporators produce two product streams, 1) liquid propane of sufficient purity to meet injection specifications goes to storage and/or the propane vaporizer and 2) produced heavy oil which goes to sales. As will now be understood, the final energy input for the process is the energy required to vaporize the propane liquid prior to reinjection downhole as shown in Figure 2.

Referring back to Figure 1, we can see that two separate parallel pathways 86 and 88 are provided to vaporize the propane prior to reinjection into the reservoir. The major proportion of the propane is vaporized via the heat pump 86, and a smaller proportion is vaporized via heat exchange 88 with flue gas etc. The heat pump as described by path B of Figure 2, consists of a two stage process whereby the solvent is first vapourized by heat exchange 80 with a latent heat source 82 typically at a temperature below the desired extraction temperature. The second stage of the heat pump process is compression of the vapour via a compresser 84 to the desired injection temperature and pressure.

The second vaporization pathway as described by path A of Figure 2 is shown at the far right hand of figure 1 is a more traditional vaporization via heat exchange. This second parallel vaporization pathway is also included to take advantage of waste heat generated within the process (i.e. from the 38% efficient engine driver for the compressor) which cannot be otherwise easily used by the heat pump. Such waste heat would typically include hot flue gases and other waste engine heat that would otherwise be rejected in a radiator. This second vaporization path includes a pump 90 and a heat exchanger 92.

Figure 2 shows a Molliere diagram for propane. The vertical axis is the pressure and the horizontal axis is the enthalpy. The diagram shows propane storage conditions (saturated liquid at 5C) and the injection conditions (saturated vapour at 50C). The figure also shows a 5C isotherm and a 50C isotherm. The thermodynamic data for propane was obtained from the NIST website of the US government.

Figure 2 shows two alternate paths labeled A and B whereby the propane is vaporized (i.e. taken from storage (saturated liquid at 5C) to injection conditions (saturated vapour at 50C).

Path A represents a conventional direct heating or energy consumption process. In this case the propane liquid is first pumped from storage at .55 MPa into a high pressure vaporizer at 1.7MPa. The line is almost vertical because it takes very little energy (<2kJ/kg) to pump the liquid propane to an elevated pressure. Heat addition initially raises the temperature of the propane to the saturation temperature of 50C. Additional heat is required to vaporize the propane. At this point, the propane vapour is at desired temperature (50C) and pressure (1.7 MPa) for injection into the reservoir. This path requires an energy input of 409 kJ/kg (the distance along the horizontal axis of path A). Accounting for the fact that a direct fired heater is only about 90-95% efficient due to flue gas heat losses etc, the total fuel requirement is ~450 kJ/kg propane. As can now be understood the inherent inefficiencies in converting any form of energy into heat and transferring the same into the propane mean that the direct heating method results in a thermal coefficient of performance of less than one.

Path B represents the more energy efficient path of the present invention that utilizes a heat pump for transferring heat into the solvent from a latent heat source. In this case, the propane is vaporized by heat exchange with a convenient (i.e. “free”) heat source. For example, the heat source might be ambient air with a temperature somewhat above 5C. The air is chilled to 5C and the propane is vaporized at a temperature of 5C to a vapour pressure of 0.55 MPa (=80 psia). The propane vapour is then compressed to the extraction chamber pressure of 1.7MPa (=250 psia). If the compression is adiabatic, then the compression path is almost parallel to the phase envelope. There is a slight amount of superheating of the vapour when it is compressed so the endpoint doesn’t lie exactly on the phase envelope and the final temperature is 54.4C. Note that the only energy input provided in path B is the work required for vapour compression (i.e.52.7 kJ/kg). Assuming typical compressor efficiencies of about 38% (by accounting for energy efficiency of the driver engine as well as compressor) then the total required energy input for path B is about 140kJ/kg of propane vapour. Thus, the heat pump offers energy savings of 69% relative to a direct energy consumption process. As set out in more detail below the present invention offers a thermal coefficient of performance of greater than one and most preferably greater than two or more.

However, Path B offers additional efficiencies. The inefficiency of the engine driving the compressor offers potential waste heat recovery from the flue gases and the radiator as a secondary heat source. It would be reasonable to recover about 50% of the initial fuel value from waste heat in the flue gas and the radiator = 70kJ/kg. The superheat of 4.4C (i.e. 54.4-50C) also offers potential recovery of 11kJ/kg via a desuperheater. These additional sources for heat recovery can be used to vaporize an additional 0.2 kg of propane (=81/409). Thus, by taking advantage of these additional efficiencies, the overall energy requirement of Path B can be further reduced to 117 kJ/kg of propane (=140/1.2). Thus, an enhanced heat pump process according to the present invention utilizing these additional heat recoveries requires only 117kJ/kg compared to the direct fired process (450kJ/kg), for an overall energy reduction of 74% relative to propane vaporization in a direct fired or energy consumption heater.

Referring back to Figure 1, two separate pathways are provided to vaporize the propane. The major proportion of the propane is vaporized via the heat pump, and, optionally, a smaller proportion is vaporized via one or more secondary heat sources such as a heat exchange with flue gas etc.

Heat for the heat pump portion of the propane vaporization is preferably supplied by heat exchange with ambient air, although the present invention comprehends all available sources of latent heat. In the case of ambient air, the heat exchangers could become fouled with frost if the surface temperature (i.e. propane pressure) for the exchanger drops too low. Strategies to mitigate frost fouling could include occasional pressure cycling of the propane in the exchangers to periodically defrost the exchanger and/or using alternate heat sources when air temperatures get to cold. For example, if there is a previously extracted well chamber that has been depleted, it may be feasible to circulate water/gas through the chamber during the cold winter months to recover geothermal heat and then switch to heat exchange with ambient air during the warm summer months. Alternatively, it may be economic to use waste heat from co-gen facilities and/or cycle water/gas through depleted SAGD chambers for heat recovery. Alternatively, geothermal heat could be supplied by circulation of water through a convenient aquifer. Alternatively, waste heat could be recovered from a gas compression station on a pipeline. In any event, it is expected that the operator might utilize one or more of these several different latent heat sources.

A detailed calculation of the expected fuel requirements of the present invention can now be understood. The fuel requirements include artificial lift (7kJ/kg), demethanizer duty (8kJ/kg), propane evaporators (30kJ/kg), and propane vaporizer (117kJ/kg) for a total of about 162kJ/kg. The total fuel energy requirement is 5600GJ/day (= 69,000 m3/day x 500kg/m3 x 162kJ / 1000, 000GJ/kJ). For the medium sized facility mentioned above, this would require 151,000m3 of methane fuel/day, with a total CO2 emission of 3kg/bbl bitumen production (as compared to 70 to 125 kg/CO2/bbl for SAGD or synthetic crude). It is worth pointing out that up to about one third or more of the fuel gas requirement could be provided by produced associated gas.

The calculations above do not fully account for smaller items such as lighting, computers and instrumentation sensors, etc. However, these are believed to be small in this context. Thus the present invention provides a method to separate, purify and vaporize the solvent that reduces the process energy requirement/energy cost/CO2 emissions by about 95% compared to the currently available steam technology.

It should be pointed out that the example presented above assumes “pure” solvent is injected and does not consider the impact of solvent purity. As, will be understood by those skilled in the art, a 1 mole% methane contamination in the injection solvent raises the extraction pressure by 150kPa (to achieve a 50C bubble point) and/or reduces the extraction temperature by about 5C (at 1.7 MPa). Similarly methane residuals in the solvent will have an adverse impact on heat transfer coefficients in the propane evaporators and thereby increase the process compression energy requirements. Then present invention comprehends that the methane specification may need to be below 1% to provide optimum economics.

Considering a carbon tax of \$200/ton and a fuel gas cost of \$3/GJ and a propane cost of \$160/m3, the present invention has about 7\$/bbl lower operating expenses than a steam process (=94%x\$3 + 94%x\$200x70x(12/44)x(2.2/2000)). Furthermore, based on Duerksen and Eloyan[8], one could anticipate an additional 10 degree API upgrade for extraction with heated propane vapor for an additional value added of \$4.00/bbl (10 x \$0.40/bbl API). An additional 2API increase could be realized from the residual propane, for an additional value added of \$.80/bbl. Additional savings of \$1 relative to SAGD will arise through the elimination of the diluent requirement and a 50% reduction in pipeline transportation costs.

An additional cost for the current process arises due to the fact that solvent that must be purchased and used for voidage replacement in the extraction chamber, and some solvent is lost to sales bitumen and fuel gas. At the present time, the fraction of the solvent vapour that can be eventually recovered via blowdown at the end of the project is unknown. Furthermore the solvent inventory costs (i.e. amount and amortization period) depends on the extraction rate. However, an upper bound on this cost can be obtained by assuming 0% solvent recovery from final blowdown. In this case, the cost of the solvent is ~\$3/bbl bitumen. If we achieve 50% solvent recovery, then the net propane solvent cost is about \$2/bbl.

The current 25% resource royalty charged by the provincial government is only applied to the operating margin after allowing the operator to fully recover all R&D, capital,

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[8] Duerksen, J.H., A. Eloyan, Evaluation of Solvent Based In-Situ Processes for Upgrading and Recovery of Heavy Oil and Bitumen, 6th Unitar Conference Proceedings, 1995, Vol 1, pg 359

operating and interest costs. Thus, the entire capital cost of a tar sands extraction facility is effectively paid by "avoided" royalties. In this financial context, the operating margin becomes the most relevant criteria to compare different extraction processes. The present invention offers operating margins $5-11/bbl higher than a SAGD process. At a royalty rate of 25%, the present invention would provide an additional $2-3/bbl of revenue for the Province Alberta. This is equivalent to 2 to 3 billion$/year of additional revenue at an anticipated production rate of 2.6 Million bbl/day.

The actual operating margins for SAGD projects are unknown. Reported supply costs vary from $8 to $14 per bbl[9]. The historical average netback price is estimated to average $13.50 (calculated from National Energy Board price data 1988-2001 using FOB costs and extrapolated light-heavy differentials). Thus, the "best" SAGD operating margin is expected to be about $5/bbl. Thus, the solvent process described in this patent application offers operating margins 2-3 times (i.e., $11-16 vs. $5) higher than the "best" SAGD with a simultaneous 95% reduction in emissions.

A further advantage of the present invention can be seen by examining the criterion of royalties paid per kg of emissions. The solvent extraction process taught herein offers an advantage 40 to 60 times higher than SAGD. The present invention is both more profitable and has a much smaller impact on the environment. Significantly, the present invention reduces the process fuel requirement so that the marginal cost of using an expensive but clean burning fuel such as gas is almost insignificant (~10-15cents/bbl).

Consider an alternative technology such as application, 2,332,685, which teaches surface deasphalting of the bitumen with combustion of the asphaltene residue as fuel for steam generation. In this case, the fuel operating expense of $3/bbl is

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[9] Canada's Oil Sands A Supply and Market Outlook to 2015, National Energy Board Canada, pg 37

saved and the API upgrade adds about $2/bbl (5API x $0.40) and some pipeline/diluent costs are avoided ($0.50/bbl) for an operating margin of $10.50/bbl. However, the CO2 emissions are effectively doubled by burning pitch instead of natural gas, so the operating margin per kg of emissions is increased by only about 5% ($10.50/($5x2) =1.05).

The discussion above assumed that a latent heat source is available at 5C for the process. However, as noted above, the actual heat source may include a variety of different sources at different temperatures and may vary seasonally. Figure 3 shows the energy efficiency of the invention relative to SAGD, as a function of heat source temperature and extraction temperature.

Interestingly, the more detailed energy calculation, as summarized in Figure 3, using a heat source at 5C and extraction at 50C, actually predicts a better (i.e. 96%) savings compared to the earlier estimate of 94-95% savings above. This difference is probably due to additional process inefficiencies and/or adverse fuel characteristics implicit in the NRCan SAGD CO2 emission factor of 70 kg/bbl.

Figure 3 shows that the energy requirement for a 70C extraction temperature is approximately twice as high as a 50C extraction. Thus, the first 42C rise (from an initial reservoir temperature of 8C to 50C) “costs” as much as the next 20C rise from 50C to 70C. This non-linear behavior reflects the fact that heat pumps have very high coefficients of performance for a small temperature rise, but heat pump advantage decreases rapidly as the required temperature rise becomes larger.

Figure 3 shows that as the temperature of the heat source increases, the process energy requirement decreases. This reflects the fact that less compression horsepower is required if the propane is vaporized at a higher temperature (pressure).

Figure 3 also shows that the energy efficiency does not achieve 100%, even if the heat source is at the extraction temperature (i.e. 50C heat source with a 50C extraction temperature). This reflects the fact that there are other additional process energy requirements beyond vaporization which include for artificial lift, solvent separation, solvent purification, pumps and an appreciable temperature driving force is required for heat exchange to occur.

Figure 3 shows a range of extraction temperatures from 40C to 70C. The optimum extraction pressure (temperature) may vary during the extraction cycle (high initially to quickly grow the solvent chamber, low during the final stages of extraction to recover the propane inventory in the chamber). The energy cost of raising the extraction temperatures from 50 to 70C is fairly small (15cents vs. 30cents per bbl), but the propane inventory cost (assuming 0% recovery) increases by $1.30/bbl, and solvent bitumen ratio and the size (cost) of the surface facilities would be increased at higher operating temperatures. Thus, the strategy of high extraction temperatures/pressures makes particularly good sense when the chamber is small, so that high initial production rates are achieved while keeping the total solvent inventory (costs) relatively small.

It can now be appreciated that the present invention is most economical if the cost per kg of solvent for compression and heat exchange is less than the cost of a direct fired heater. Thus, for solvents where the vaporization temperature (pressure) is lower, the compression energy requirement is higher so there is a practical limit to the usefulness of a heat pump. For example, a heat pump would not be useful for a steam extraction process (i.e. SAGD) because the compression energy requirement for a 220C temperature rise would greatly exceed the latent heat of vaporization savings.

However, for a solvent extraction process operating at 50C, with absorption of heat from a latent heat source at 5C according to the present invention (i.e. an arbitrary but "reasonable" heat source temperature), the thermodynamic coefficient of performance is 6.2 (=278K/(323K-278K)). This means that according to the methods and apparatuses of the present invention one joule of work in the compressor can deliver 6.2 joules of heat at 50C. In reality, compressors are inefficient, so only about 38% of the fuel energy consumed by the driver engine is actually available for compression. However, for a low temperature extraction process (i.e. at 50C) substantial process heat can also be recovered from both the flue gas and the radiator (about 50% of the initial heating value of the fuel). Thus, for 100 Joules of compressor fuel energy, it is expected that about 38Jx6.2 +50J = 285J of energy can be delivered as vaporized solvent at 50C. So a practical coefficient of performance would be about 2.8. The present invention comprehends a coefficient of performance of more than one and most preferably more than 2 or more.

By comparison, a direct fired boiler is only expected to deliver 90-95% of the heating value of the fuel as heat due to inefficiencies such as flue gas heat losses and control of the fuel air ratio, etc. Thus, the present invention is expected to provide a 68% (=1-(1/2.8))/.95 reduction in energy cost relative to a direct fired heater for solvent vaporization. Furthermore, this invention also provides a 68% reduction of GHG (green house gas) emissions relative to solvent vaporization via a direct fired heater (burning gas).

If one considers that the heated solvent extraction process offers potential energy reductions of 81% (as noted above) compared to SAGD, and that the present invention teaches a further energy reduction of 68% (i.e. of the remaining 19%) through the use of a heat pump and various heat recovery techniques, it is demonstrated that the present invention offers a potential energy reduction of about 94% relative to SAGD[10].

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[10] The reduction depends on both the extraction temperature and the heat source temperature

Additional energy efficiencies arise from desuperheating of compressed solvent vapour, and the ability to utilize convenient low temperature sources of process heat. Additional energy requirements arise from the energy required for solvent recovery from the produced fluids and the energy required for purification of the solvent to meet injection specifications as well as lift supplied to pump the produced fluids from the reservoir.

By way of further example, if the present invention was used to deliver the entire 2.6 million bbls/day of bitumen production by 2010, then the total emissions from bitumen extraction would be about 3-5 megatons/year vs 73 megatons/year for the best current SAGD bitumen extraction technology or 125megations/yr for synthetic crude.

To put these numbers in context, the elimination of all automobiles in Canada would reduce emissions by only 53 megatons/year (Annex C, pg. C-26 of reference 1). Current estimates of Canada's excess emissions above the Kyoto Agreement are about 200 megatons/year. Thus, the elimination of 70 -120 megatons/year of CO2 emissions, as taught by this invention, provides a means to achieve about 50% of the reduction mandated by the Kyoto accord and simultaneously increase the profit margins of production.

The present invention provides strong commercial incentives for implementation, as the expected operating margins (revenue, net of operating expenses) and profits are expected to be 2-3 times higher than the best current SAGD technology. Higher profit margins will also enable a larger percentage of the tar sands to be extracted profitably. Industry, government and the environment should all benefit from this technology.

While the foregoing description has concentrated on preferred embodiments of the invention it will be understood that various alterations and modifications are possible without departing from the broad scope of the attached claims. Some of these have been discussed above, and others will be apparent to those skilled in the art. For example, although the method and apparatus of the present invention is most useful in association with heavy oil, they also have application in enhancing recovery of less viscous oils.

I CLAIM

1. A method for extracting oil from a formation, said method comprising:

    selecting a condensing solvent having a high vapour pressure at low temperatures wherein said solvent has a boiling point lower than about 20C at atmospheric pressure,

    heating said solvent by means of a heat pump;

    injecting said solvent as a vapour into said formation,

    condensing said solvent in said formation to deliver a latent heat of condensation to said oil and to mobilize said oil,

    producing said oil and said solvent;

    recovering said solvent from said produced oil,

    re-heating said solvent by means of said heat pump, and

    re-injecting said solvent into said formation.

2. A method for extracting oil from a formation as claimed in claim 1 wherein the solvent evaporates at a temperature below an average ambient temperature of a location where the extraction is occurring, and heat from ambient air is used in said heat pump.

3. A method for extracting oil from a formation as claimed in claim 1 wherein the solvent evaporates at a temperature below an average ambient temperature of a location where the extraction is occurring, and geothermal energy is used in said heat pump.

4. A method for extracting oil from a formation as claimed in

claim 3 wherein said geothermal energy is recovered from a previously extracted underground formation.

5. A method for extracting oil from a formation as claimed in claim 1 further including the step of heating said solvent by means of a secondary heat source.

6. A method for extracting oil from a formation as claimed in claim 5 wherein said secondary heat source is a source of waste heat.

7. A method for extracting oil from a formation as claimed in claim 1 wherein said step of heating said solvent by means of a heat pump includes compressing said solvent vapour to a predetermined formation injection pressure and temperature.

8. A method for extracting oil from a formation as claimed in claim 1 wherein said step of recovering said solvent from said produced oil further comprises a dewatering step.

9. A method for extracting oil from a formation as claimed in claims 1 or 8 wherein said step of recovering solvent from said produced oil further comprises stripping at least some noncondensables from said oil.

10. A method for extracting oil from a formation as claimed in claim 9 wherein said step of stripping at least some noncondensables occurs prior to said solvent being recovered from said oil.

11. A method for extracting oil from a formation as claimed in claim 9 wherein said step of stripping at least some

noncondensables from said oil occurs prior to said solvent being reinjected into said formation.

12. A method for extracting oil from a formation as claimed in claims 7 to 11 wherein said noncondensables include at least methane.

13. A method for extracting oil from a formation as claimed in claim 1 wherein said step of recovering said solvent from said oil further includes heating said oil by means of a heat pump to evaporate said solvent.

14. A method for extracting oil from a formation as claimed in claim 13 wherein said step of heating said oil to evaporate said solvent further includes a vapour recompression step to supply a portion of the latent heat of evaporation of the solvent.

15. A method for extracting oil from a formation as claimed in claim 1 wherein said step of recovering said solvent from said oil further includes stripping heavies out from said recovered solvent.

16. A method for extracting oil from a formation as claimed in claims 1 to 15 further including the step of adding make up solvent prior to re-injection of the solvent into the formation.

17. A method for extracting oil from a formation as claimed in claims 1 to 16 wherein said solvent has a critical temperature that is higher than a formation extraction temperature.

18. A method for extracting oil from a formation as claimed in claims 1 to 17 wherein said solvent is propane.

19. A method for extracting heavy oil from a formation, said method comprising:

supplying a condensing solvent comprising a solvent having a boiling point lower than about 20C at atmospheric pressure,

supplying thermal energy to said solvent by evaporating said solvent at ambient conditions and

compressing said vapour solvent to a predetermined temperature and pressure to cause said solvent to condense in situ extraction conditions;

injecting said solvent into said formation,

condensing said solvent in said formation to deliver a latent heat of condensation to heat said heavy oil, and

recovering said solvent upon extraction of said heavy oil.

20. A method for extracting oil from a formation as claimed in claim 14 wherein said step of heating said solvent has a thermal coefficient of performance greater than 2.

Figure 1

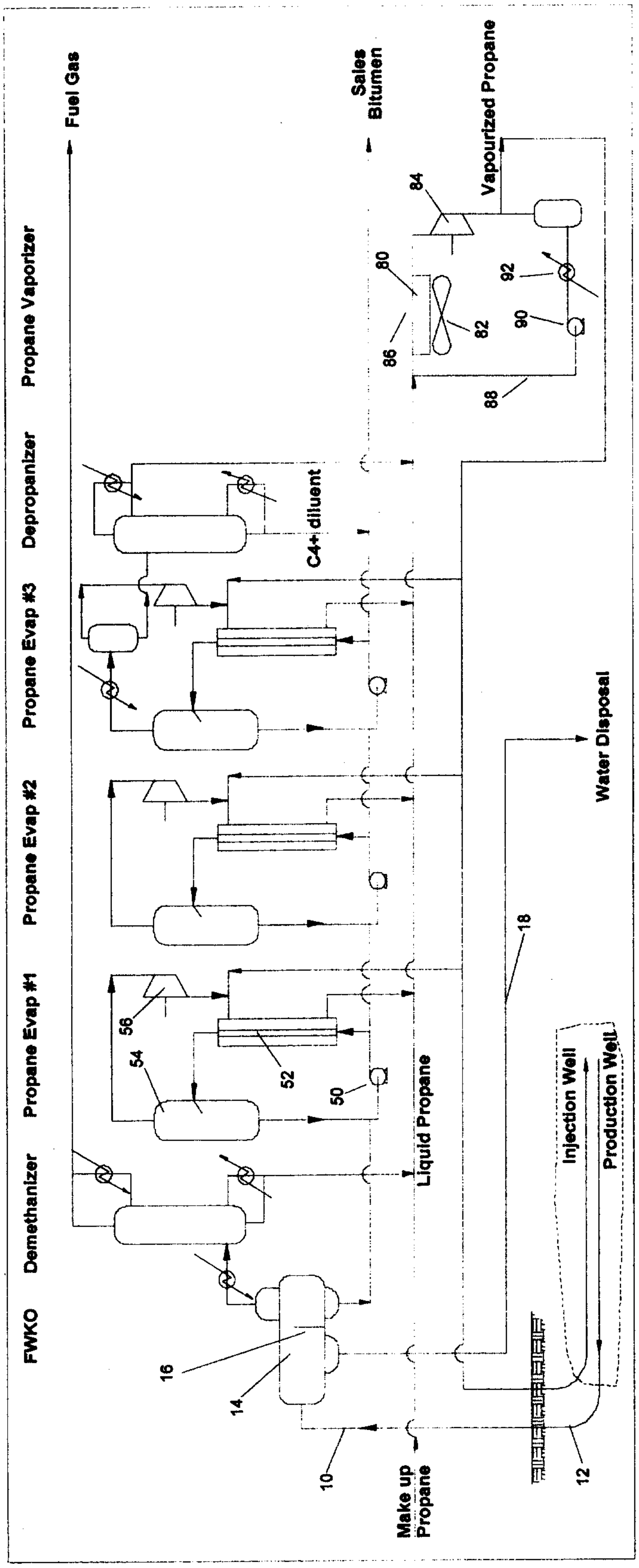

FWKO
Demethanizer
Propane Evap #1
Propane Evap #2
Propane Evap #3
Depropanizer
Propane Vaporizer
Fuel Gas
Sales Bitumen
Vapourized Propane
C4+ diluent
Liquid Propane
Water Disposal
Make up Propane
Injection Well
Production Well
10
12
14
16
18
50
52
54
56
80
82
84
86
88
90
92

Figure 2

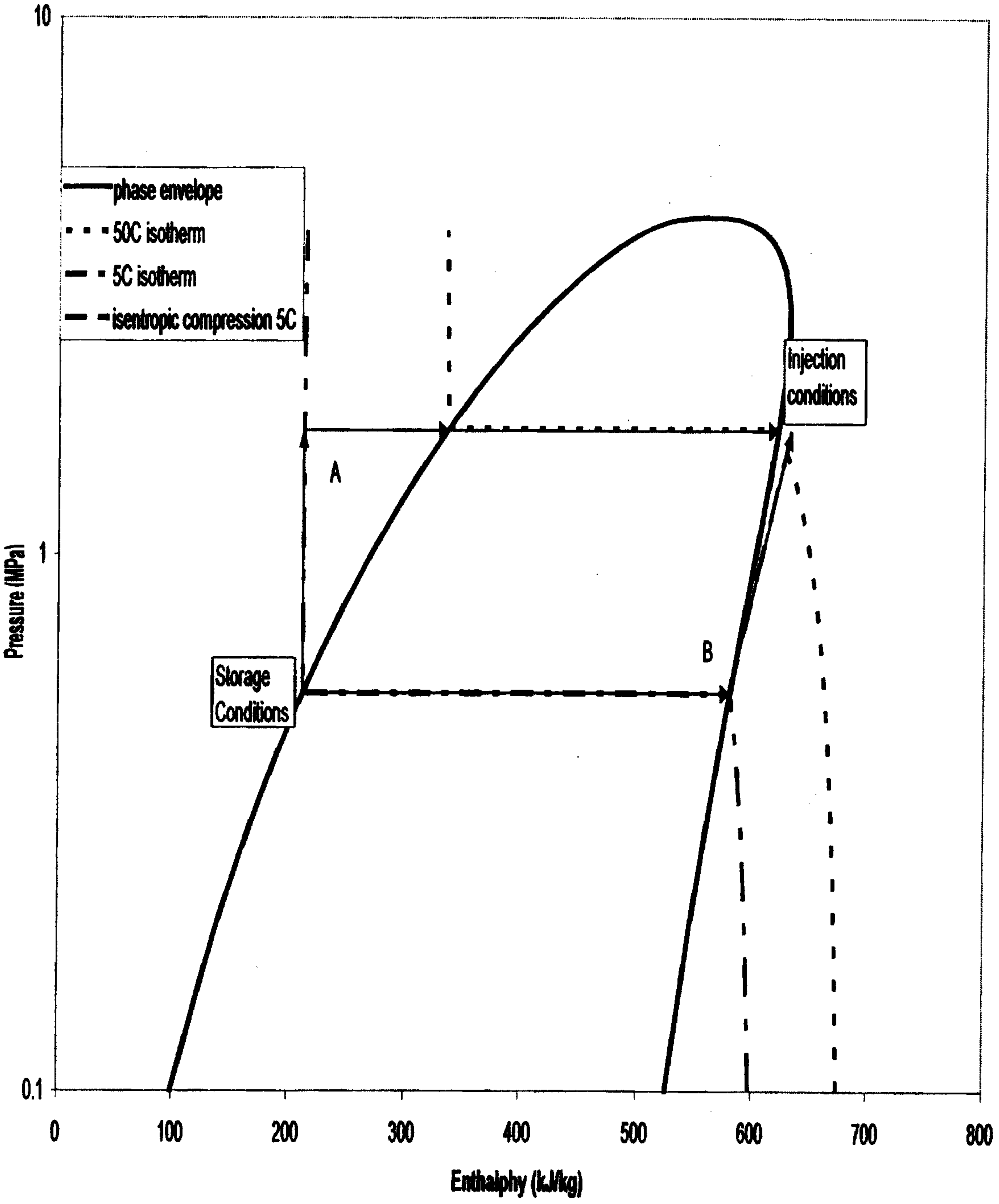
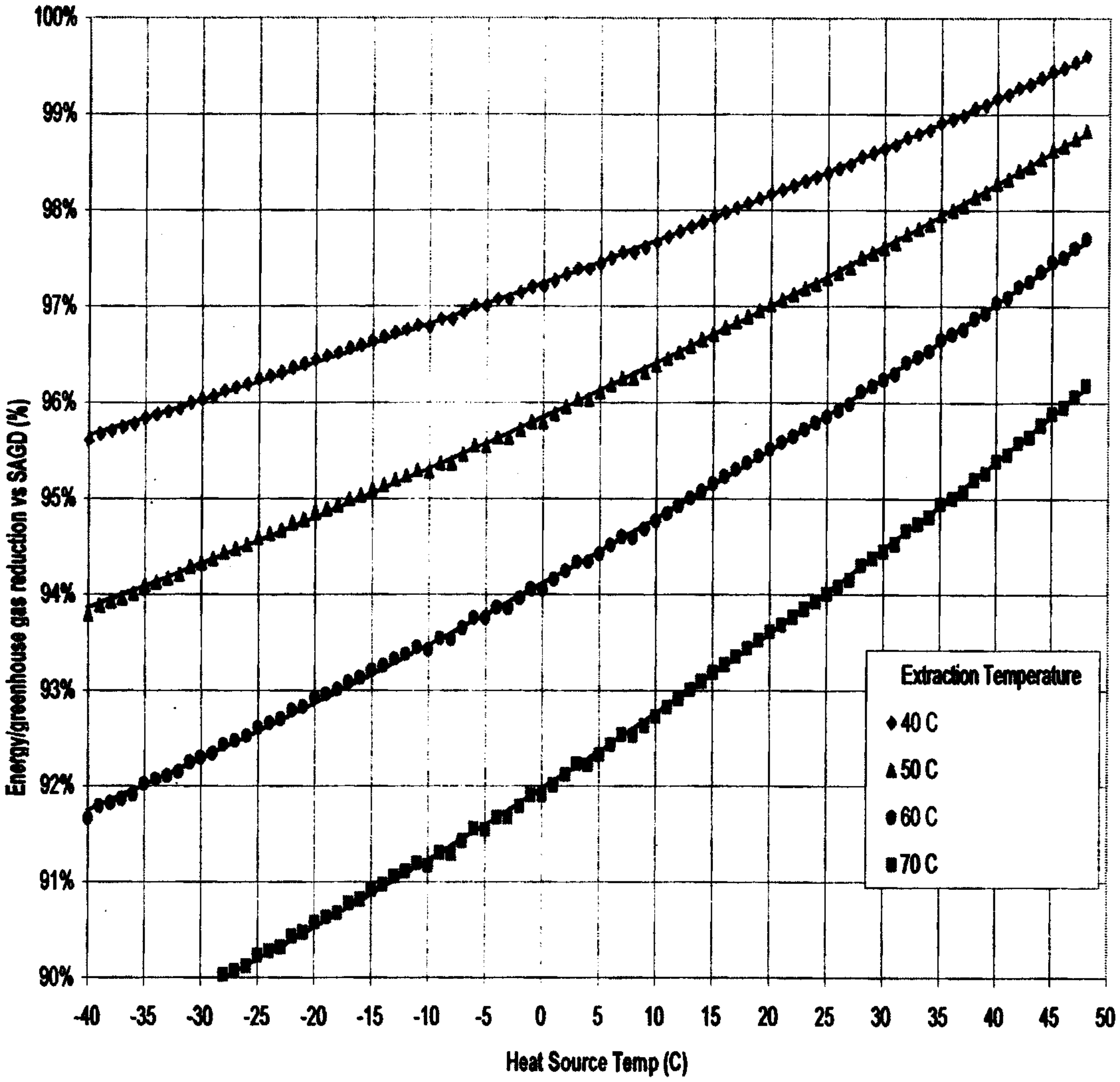

10
1
0.1
Pressure (MPa)
0
100
200
300
400
500
600
700
800
Enthalphy (kJ/kg)
phase envelope
50C isotherm
5C isotherm
isentropic compression 5C
Injection conditions
Storage Conditions
A
B

Figure 3


100%
99%
98%
97%
96%
95%
94%
93%
92%
91%
90%
Energy/greenhouse gas reduction vs SAGD (%)
-40
-35
-30
-25
-20
-15
-10
-5
0
5
10
15
20
25
30
35
40
45
50
Heat Source Temp (C)
Extraction Temperature
◆ 40 C
▲ 50 C
● 60 C
■ 70 C


FWKO
Demethanizer
Propane Evap #1
Propane Evap #2
Propane Evap #3
Depropanizer
Propane Vaporizer
Fuel Gas
16
56
54
14
52
50
10
C4+ diluent
Sales Bitumen
Make up Propane
Liquid Propane
86
80
84
82
Vapourized Propane
88
90
92
18
12
Injection Well
Production Well
Water Disposal