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HIGH-PRESSURE SOLVENT EXTRACTION OF
METHANE FROM GEOPRESSURED BRINES:
TECHNICAL EVALUATION AND COST ANALYSIS

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Table of Contents

Abstract	1
Introduction	1
Solvent Extraction of Methane From Geopressured Brines	3
Solvent Selection	3
Solubility Measurements	4
Process Design	9
Economic Assessment	10
Comparison with Conventional Injection Disposal	12
Reinjection into the Producing Formation	15
Energy Requirements	15
Thermal Energy Contribution	16
Economics of Thermal Energy Recovery	16
Reinjection Strategy	17
Summary	18
Acknowledgment	20
References	21
Appendix A - Solvent Extraction Cost Estimate	23
Appendix B - Cost Estimate of Injection by Mechanical Pumping	26

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HIGH-PRESSURE SOLVENT EXTRACTION OF
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ABSTRACT

Solvent extraction is proposed as a means of recovering dissolved methane from geopressured-geothermal brines at high pressures. Our assessment shows that additional investment in a high pressure solvent extraction plant preceding direct injection disposal of brines into isolated aquifers can be profitable. The technical and economic issues are discussed, and compared with other injection methods such as complete depressurization for methane recovery followed by conventional mechanical pumping. The contributions of hydraulic (pressure) energy recovery and geothermal power production are also assessed. For deep injection into the producing formation, we conclude that methane extraction processes are not applicable, insofar as maintenance of high surface pressures provides no clear-cut energy benefits. As a first step in the evaluation of solvent extraction, the solubility of a promising solvent candidate, n-hexadecane, was measured in 15 wt% NaCl solutions at temperatures up to 150°C. The solubility of a potential low cost solvent, No. 2 Diesel fuel, was also measured.

INTRODUCTION

The geopressured-geothermal aquifers located along the Northern Gulf of Mexico Basin are estimated to contain from 3000 - 100,000 quads of energy in dissolved methane (CH_4) alone.⁽¹⁻⁵⁾ The wide range in the estimate stems from differences in the interpretation of the available data base regarding reservoir distribution, properties, and gas saturation. But even at the lower end of the estimates, a significant energy source is recognized to exist in the geopressured aquifers. In the past several years, this resource has been the subject of numerous studies and a federal program is now aimed at better resource definition and development feasibility.

Swanson & Osoba⁽⁶⁾, and Doscher, et al⁽⁷⁾ have assessed the technical issues and costs of energy recovery from geopressured aquifers. Even when assuming favorable reservoir and fluid parameters, they conclude that combined CH₄ and thermal energy recovery from geopressured fluids is not presently economical. Wilson, et al⁽⁸⁾ in an earlier assessment of thermal and chemical energy recovery came to similar conclusions. It is conceivable that the value of energy will rise faster than plant and field costs. If so, then unconventional energy sources such as geopressure will naturally benefit and move toward economic feasibility.

The resource factors that affect the economics are not likely to improve. Reservoir size and formation properties were already assumed to be favorable. Fluid properties such as maximum usable temperature and dissolved CH₄ concentration were also assumed to be at or very near the upper limits. Aside from artificial means, such as tax benefits, progress toward economic feasibility can only be achieved by increasing the ratio of the value of the salable commodity over total costs.

Drilling and completion costs comprise the largest share of the total development costs. Obviously cost reducing innovations in drilling would have a positive impact on development. Technical innovations which result in more efficient recovery of energy - chemical, thermal, or hydraulic would in effect increase sales and hence contribute toward economic viability. For example, plant and field operating and maintenance (O&M) costs are expected to be quite high, particularly the power costs for injection of enormous quantities of produced fluids. The pump power costs rise steeply with required injection pressures, and become a serious parasitic power loss. For this reason, maintenance of high surface pressure to drive the injection operation may be very cost effective by eliminating a power consuming operation and injection pumps altogether. At present geopressure test sites, fluids are in fact disposed of by this means into shallow aquifers. However, by not fully depressurizing the fluid, some amount of CH₄, depending on pressure and brine salinity, remains in solution and is not recovered. At 1000 psi, 5-10 scf/bbl, perhaps 25% of the total, remains dissolved and if not recovered, is unrealized revenue. Several methods have been identified for recovery of

dissolved CH_4 at temperature and elevated pressures.⁽⁹⁾ Of those techniques solvent extraction appears to be the most promising and is the focus of this paper. The suitability of other methods such as gas stripping and the use of hydraulic conversion devices has been discussed elsewhere.⁽⁹⁾

SOLVENT EXTRACTION OF METHANE FROM GEOPRESSURED BRINES

Solvent Selection

Solvent extraction is one of several mass transfer operations in chemical processing for separating components of a solution. Separation depends on the unequal distribution of the solution components between two liquid phases. To recover CH_4 from geopressured brines by solvent extraction, solvent losses owing to dissolution in the brine must be held to a minimum, particularly since CH_4 is a relatively low valued commodity and is present in the brine in low concentrations. For the process to be technically and economically viable, the solvent must have the following properties:

- 1) It must have very low solubility in brine. Solvent recovery from the spent brine stream by reprocessing is not practical because of the very large volume of brine that must be handled and maintained at high pressure for injection. Solvent losses must be held to a small fraction of the recovered CH_4 value.
- 2) There must be a highly favorable distribution coefficient between the solvent and brine for CH_4 in order to keep the solvent to brine ratio to a minimum.
- 3) Desorption of CH_4 from the solvent must be relatively simple. Methane must be readily recoverable from the solvent by simple depressurization, otherwise the need for additional separation steps will add to the cost of the process.
- 4) The solvent must have good thermal and chemical stability at the geopressured brine temperatures of 150°C .

As a group the saturated or paraffinic hydrocarbons are among the least water soluble organic compounds. With increasing molecular weight, the saturated hydrocarbons become liquid or waxy solids at room temperature.⁽¹⁰⁾ They also become less and less soluble in water. Among the higher molecular weight compounds, n-hexadecane, $C_{16}H_{34}$, appears to be the optimum choice. It is high boiling (vapor pressure at 150°C is about 10 Torr) but is a liquid at room temperature. It is available in commercial quantities. The distribution coefficient for CH_4 in the n-hexadecane - brine system is estimated to be about 100 based on CH_4 solubility in brine at 50-125°C from O'Sullivan and Smith⁽¹¹⁾, and low pressure CH_4 solubility in hexadecane at 25-200°C from Cukor and Prausnitz⁽¹²⁾. If the most optimistic CH_4 brine solubility data is used, then the distribution coefficient reduces to approximately 25.

While we would predict hexadecane to be very insoluble in brine at 150°C by extrapolating McAuliffe's⁽¹³⁾ correlation for the solubility of paraffinic hydrocarbons C_1 through C_8 in water at room temperature, there are no literature experimental data on hexadecane solubility at conditions encountered in geopressured brines. Since the economic viability of the solvent extraction process hinges on the degree of hexadecane solubility in brine, it was necessary to generate such data over the temperature range 20°C to 150°C.

Solubility Measurements

The experimental apparatus consisted of I) a temperature controlled stirred vessel for measurements up to 100°C and II) for temperature above 100°C, a temperature controlled autoclave with an exterior pumped circulating loop for dispersing hexadecane through the aqueous phase.

I) Experimental procedure for measurements up to 100°C.

Approximately 1 liter of 15 wt% brine and 0.5 liters of 95% purity hexadecane* were brought up to temperature. The batch was

* The other 5% consists of mainly C_{14} - C_{18} compounds. This quality of hexadecane was used in order to be consistent with larger scale applications.

stirred and the phases allowed to separate over a range of settling times. The vessel was covered, but not air tight. Temperature was maintained during the entire stirring and settling periods. A ~ 30 ml brine sample was drawn with a hypodermic syringe and transferred into a separatory funnel. A known amount of methylene chloride was added to extract the hexadecane dissolved in the brine. The methylene chloride fraction was isolated and analyzed for hexadecane by gas chromatography. Standards were run using 99+% purity n-hexadecane.

II) Experimental procedure for measurements at 150°C.

The autoclave was evacuated; then the brine and hexadecane were drawn in. Temperature was brought up to 150°C and maintained. The lighter phase, hexadecane, was dispersed through the heavier phase, 15 wt% NaCl, using an air-driven positive displacement pump in a closed circulating loop arrangement. An effective dispersing head was constructed with the aid of a 10- μ m pore-sized steel frit.

The two phases were allowed to separate by settling. A sample was taken at pressure and temperature through a tube extending into the brine phase. The sample is chilled to ice temperatures immediately upon entering an evacuated double-ended glass sampling bulb. The bulb is weighed before and after filling. The total pressure in the bulb at ~ 20°C never rises above ambient. A known amount of methylene chloride was drawn into the bulb to extract the hexadecane from the brine. Finally the methylene chloride fraction is analyzed for hexadecane by gas chromatography.

Measurements of hexadecane solubility in brine in the presence of CH_4 were conducted by simply introducing CH_4 from a pressurized cylinder at the pressures desired. These measurements were made to determine what effect, if any, CH_4 has on the solubility of hexadecane in brine. The gas was bubbled through

both phases using the solvent dispersing head previously described. After gas saturation was achieved a static CH_4 atmosphere at pressure was maintained over both phases throughout the hexadecane circulating, phase settling, and sample collecting periods just described. The highest CH_4 partial pressure was 450 psia and represents the upper limit of CH_4 partial pressure in the exit brine stream that is envisioned for a countercurrent solvent extraction process involving geopressured brine and n-hexadecane. The results of these measurements are summarized in Table 1.

TABLE 1. SOLUBILITY OF N-HEXADECANE IN 15 WT% NACL SOLUTIONS.

Temperature, °C	Settling Time, Min	Solubility, ppm (wt)	Remarks
20.4	10	1.4	
20.4	10	<0.6	Brine centrifuged to remove entrained hexadecane before analysis.
21.0	10	<0.4	
42.0	10	5.2	
40.0	10	<0.4	
42.0	20	0.6	
61.0	10	3.0	
79.0	10	3.3	
99.0	5	0.7	
100.0	10	<0.4	
100.0	20	<0.4	
100.0	30	<0.4	
152.0	55	1.6	
152.0	65	<0.2	
150.0	140	<0.2	450 psia pp CH ₄
150.0	60	<0.2	270 psia pp CH ₄
150.0	90	0.9	100 psia pp CH ₄
150.0	30	8.0	50 psia pp Ar
150.0	50	3.3	50 psia pp Ar

Sufficient care must be exercised during sampling in order to assure that undissolved hexadecane does not contaminate the sample. We suspected that this occurred in some of the runs as indicated by some of the relatively high values of hexadecane solubility. Results of duplicate measurements that were aimed at minimizing free hexadecane pickup produced lower solubility values and confirmed our suspicions. Also centrifuging the brine sample first to separate any free hexadecane resulted in significantly lower dissolved hexadecane values. Pickup of undissolved hexadecane in the sample can result from the presence of minute quantities in the sampling lines, or from incomplete phase separation. Measurements at 99-100°C with settling periods of 5-30 minutes indicated that a settling period of 10 minutes is sufficient insofar as longer settling periods do not result in lower solubility values.

The relative quantities of methylene chloride extractant and sampled brine were such that the concentrations of hexadecane in the extracts were below the detection limit of the chromatograph. This was true over the entire temperature range of the measurements. As a result, temperature dependence could not be determined. The results do show that n-hexadecane solubility in 15 wt % brine in the temperature range 20-150°C with and without the presence of CH_4 up to a partial pressure of 450 psia is < 1 ppm by weight. In an actual process employing n-hexadecane as a solvent to extract CH_4 from geopressured brines, the hexadecane losses owing to brine dissolution should be acceptable. For example, at 1 ppm solubility and at a cost of \$4/lb, the presumably unrecoverable losses from hexadecane dissolution in the brine would be roughly 3% of the value of methane contained in geopressured brines, assuming a CH_4 concentration of 10 scf/bbl and at a CH_4 price of \$5/Mcf.

Measurements below 100°C were also conducted with No. 2 Diesel fuel, a potential low cost organic solvent. The apparatus and procedures that were employed were the same as the hexadecane runs, except for the extraction step. Diesel fuel was extracted from the brine with trichloro-trifluorethane and the extract analyzed for Diesel by IR spectroscopy. The results are given in Table 2.

Table 2. NO. 2 DIESEL FUEL SOLUBILITY IN 15% WT NACL SOLUTIONS

Temperature, °C	Settling Time, Min	Solubility, ppm (wt)
21	10	49-59
61	10	37-46
100	10	32-40

Although Diesel fuel has a much higher solubility in brine than hexadecane, it is of course much cheaper. Using 30 ppm as the solubility of Diesel fuel in brine at 150°C and a Diesel fuel value of 14¢/lb (\$1/gallon), unrecoverable losses in an actual application would be also ~ 3% of the value of CH₄ dissolved in the brine.

Diesel fuel should be a good CH₄ extractant in that its hydrocarbon composition is a mixture of heavy paraffins, olefins, and aromatics - including a substantial fraction of hexadecane. The initial boiling point (~ 176°C) is well above expected reservoir fluid temperatures (150°C) and there are no unforeseen chemical or thermal stability problems. Equally important among these factors which determine its suitability as a solvent is its wide availability and low cost.

In summary we can say that both n-hexadecane and Diesel fuel are promising candidate solvents for recovery of dissolved CH₄ from geo-pressured brines at 150°C and elevated pressures.

Process Design

The process itself is relatively simple. Brine is contacted with the solvent in a counter flow extraction tower at a pressure that is sufficiently high to permit disposal of the brine by injection into shallow aquifers. Residual (~ 10 scf/bbl) dissolved CH₄ is extracted into the solvent which is subsequently depressurized to release the gas for

recovery. As previously discussed, because the distribution of CH_4 in hexadecane relative to brine is very high, the solvent circulation rate and hence pump energy expenditures are acceptably low. We believe that solvent losses can be kept to a minimum because as our measurements show hexadecane solubility in brine is exceedingly low. Vapor losses from the solvent depressurization step should also be tolerable owing to hexadecane's low vapor pressure. The solvent extraction operation is depicted in Figure 1.

We have conducted an economic feasibility analysis of this process. The capital costs and the rate of return on capital are computed and compared with the more conventional method of brine disposal -- viz. brine depressurization to release CH_4 , followed by pumping to ~ 1000 psi for injection.

The case in which brine is maintained at 1000 psi for injection without any attempt to recover residual dissolved CH_4 is not considered an economic option. However, this method may be operationally expedient, particularly if the other CH_4 recovery options are only marginally superior. The cost comparison and analysis simply considers that if there is 10 scf CH_4 /bbl to be recovered prior to injection, then what would be the rate of return on additional capital investment to extract this salable commodity.

Economic Assessment

Our economic analysis of the solvent extraction process is based on the following assumptions:

1. A production rate of 40,000 bbl/day with a wellhead pressure of at least 1000 psi.
2. Injection pressures of 1000 psi are required for fluid disposal into shallow aquifers.
3. Methane solubility in brine at 1000 psi and 150°C is 10 scf/bbl.
4. 90% of the CH_4 is recovered from the brine using hexadecane as the solvent.
5. A methane value of \$5/Mcf.

Figure 1. High Pressure Solvent Extraction of Residual Dissolved Methane in Geopressured Brines. Pressure is Maintained for Direct Injection.

While most solvent extraction processes are carried out at ambient pressure, this process would operate at 1000 psi. Equipment costs and instrumentation reflect high pressure operating conditions. Major components of the solvent extraction system are also shown in Figure 1.

The capital costs of an extraction plant sized to process 40,000 bbl/day of brine is estimated to be \$540,000. An inventory of hexadecane solvent is required at a cost of \$250,000. The annual operating and maintenance costs of the extraction plant is estimated to be \$81,000. This includes the cost of replenishing hexadecane which is continually lost in the process. Additional assumptions and calculations for the cost analysis can be found in Appendix A. Table 3 summarizes the steps used to determine the rate of return on capital investment. For the solvent extraction process, the rate of return on the total capital investment of \$790,000 is 41%.

Comparison with Conventional Injection Disposal

For purposes of comparison the cost of brine disposal after depressurization to 100 psi followed by injection at 1000 psi is also shown in Table 3. We assumed that the depreciable capital investment in pumps is \$150,000. Because pump power costs are high (see Appendix B), there is a net operating loss that we assume can be used to offset profits elsewhere. The 48% tax savings plus the added back depletion allowance and depreciation gives a rate of return of 23% for the \$150,000 investment.

If the hydraulic energy is utilized, then the pump power costs can be reduced by coupling hydraulic turbine drives to the injection pumps. However, additional maintenance costs attached to the hydraulic turbine drive equipment more than offset the power savings. Further, the costs of the turbine drive system almost doubles the capital investment (see Appendix B). The cost figures in column 3 of Table 3 show that the hydraulic energy recovery option is not a good investment.

Therefore, of the 3 options considered, direct injection coupled with high pressure CH_4 extraction gives the highest cash flow and return on

capital investment. This method also consumes the least amount of energy and would be the least vulnerable to equipment malfunction and breakdown. The depressurization option for CH_4 recovery followed by conventional pumping (electric motor drive) consumes the largest amount of energy - approximately 45% of the recovered CH_4 , and has relatively high maintenance costs due to the high stress environment that the injection pump must operate in. As a result there is a negative cash flow after taxes, but a small positive cash flow when depletion and depreciation allowances are added back in. As a result the rate of return on the smallest capital investment of the 3 options, is 23%.

Installing hydraulic turbine drives to utilize the hydraulic energy potential would reduce electrical energy needs for pumping by 65%. With this option, only 15% of the recovered CH_4 energy equivalent is used for injection. However, the additional maintenance costs for the hydraulic turbine drives added to the already high pump maintenance costs more than offset the energy savings. The capital investment is never returned in the lifetime of this project option. Further, the reliability of turbine drives operating at 150°C , 900 psi ΔP with continual gas evolution is not known.

Therefore, of the 3 options, pressure maintenance with high pressure methane recovery is the most economic alternative for injection based on rate of return of investment and on present worth comparison of costs.

Process evaluation tests are now in progress at the Lawrence Livermore National Laboratory. Extraction tests are being conducted with hexadecane solvent in sub-pilot scale equipment operated in a continuous, counter-current flow mode. A synthetic geopressured fluid is produced with a feed system which continuously dissolves CH_4 in sodium chloride solutions at temperatures of 150°C and pressures up to 1500 psi. Results of these measurements will be the subject of another report.

TABLE 3. ECONOMIC ANALYSIS OF SOLVENT EXTRACTION PROCESSING

Basis

Flowrate: 40,000 bbl/day

CH₄ content: 10 scf/bbl @ 1000 psi & 150°CCH₄ recovery: 90%CH₄ value: \$5./Mcf

	<u>Solvent Extraction</u>	<u>Conventional Depressurization</u>	
		<u>w/o hydraulic energy recovery</u>	<u>with 65% hydraulic energy recovery</u>
Annual Revenue	\$657,000	\$657,000	657,000
Capital Investment			
Depreciable	540,000	150,000	280,000
Non-Depreciable	250,000		
Royalty (10%)	(65,700)	(65,700)	(65,700)
Depletion Allowance (15%)	(98,600)	(98,600)	(98,600)
Depreciation (10 yr life)	(54,000)	(15,000)	(28,000)
Operation & Maintenance	(81,000)	(620,500)	(699,000)
Pretax Income	357,800	(142,800)	(236,000)
Post Tax Income (48% tax)	186,100	(74,240)	(122,700)
Add Back Depletion & Depreciation	152,600	113,600	126,600
Annual Cash Flow	338,600	39,320	3,800
Rate of Return	41%	23%	(-)
Present Worth @ 15% ROR	1,196,000	3,262,000	3,788,000

REINJECTION INTO THE PRODUCING FORMATION

Energy Requirements

Reinjection of geopressured brines into the producing aquifer will by maintaining pore pressure increase the total resource recovery over that from primary pressure depletion. Garg⁽¹⁴⁾ calculates, using a reservoir simulator, an enhancement factor ranging from 2 to 10 depending on the reservoir parameters. Maintenance of high pore pressure by reinjection may also control surface subsidence. However, there are inherent problems associated with deep injection. The biggest disadvantage is the energy consumption required by the injection pumps. According to Doscher⁽⁷⁾ the amount of methane consumed to supply the power to inject a barrel of fluid is about 5 ft³ (5000 Btu) per 10³ psi of injection pump pressure. The injection pump pressure, P_i , can be computed from the following:

$$P_i = P_r + \Delta P_{pf} + \Delta P_{dd} - P_h - P_s \dots \dots \dots (1)$$

where

P_r = actual reservoir pressure near the injection wellbore

ΔP_{pf} = injection pipe frictional losses

ΔP_{dd} = drawdown losses

P_h = hydrostatic pressure

P_s = surface pressure available at the input side of the injection pump.

To get a rough estimate of the magnitude of the injection pump pressures required, the following reservoir/well conditions were assumed:

reservoir pressure $P = 11,000$ psi
reservoir permeability $k = 20$ md
net pay $h = 500$ ft
well depth $H = 13,000$ ft
casing size $D = .4933$ ft
number of injection wells = 2
fluid flow rate, $q = 40,000$ bbl/day
produced brine temperature $T_1 = 150^\circ\text{C}$
spent brine temperature $T_2 = 150^\circ\text{C}$
production/injection well spacing = 1 mi.

We calculate that about 3000 to 3500 psi is required for full injection into the producing formation. The energy required is equivalent to 15 to 20 standard cubic feet of CH_4 per barrel of fluid, which approaches the amount of dissolved CH_4 recoverable from high salinity waters. If recovered CH_4 must be consumed to power reinjection, then there is no derived benefit!

Thermal Energy Contribution

There is, however, the thermal component of the energy that could be extracted to generate some electrical power. For a 150°C resource temperature and with a binary conversion process, Elliott ⁽¹⁵⁾ calculates a net power output of 1 kW-hr per 340# of fluid...roughly a kW-hr per barrel of fluid. The electrical power requirements for pumping fluids, computed from the usual formula found in handbooks and pump manuals, is 0.44 kW-hr per barrel of fluid per 10^3 psi, assuming a pump efficiency of 70%. Therefore, the thermal energy could supply electrical energy needs to drive injection up to about 2300 psi, about 2/3 of the energy requirements estimated above.

Economics of Thermal Energy Recovery

The cost of electrical power production is estimated using base figures from a report by Fluor, Inc. ⁽¹⁶⁾. For a 45MW (net) power plant, binary process, and a fluid temperature of 182°C , the direct capital costs of the plant alone are $\$38.7 \times 10^6$ and the operating and maintenance costs are $\$1.88 \times 10^6$ per year in 1977 dollars. While the amount of 182°C fluid required to produce a kW-hr of electricity is 159#; at 150°C , 340# of fluid will be required to generate a kW-hr of electricity, roughly twice that of 182°C fluid. ⁽¹⁵⁾ Therefore, to arrive at a cost estimate for a plant operating with at 150°C fluid we increased the Fluor cost estimates by the scale factor, $2^{0.6}$ (from Perry (17)), for a plant handling twice the fluid volume. Assuming an annual inflation rate of 10% from 1977 to 1981, the estimated capital and O&M costs for the larger binary power plant are $\$86 \times 10^6$ and $\$4.2 \times 10^6$, respectively. If we assume a 25 year

straight line depreciation, 10% royalty, 15% depletion allowance, 48% federal tax, and allow a 15% annual rate of return on the capital, the busbar power cost is 65 mills/kW-hr. This is without the inclusion of costs for producing or disposing the geopressed fluids. Thus while thermal energy is available in the produced brine, the cost of generating power from this source probably exceeds the cost of power available from other sources. Therefore, utilization of thermal energy for power production is not a cost effective option at this time.

Reinjection Strategy

A geopressed reservoir can be produced by primary pressure depletion for several years before the rate begins to decline and finally drop below an economic level of production. At this point only a small fraction of the total fluid in place will have been produced. Garg⁽¹⁴⁾ has shown that depending on the assumed reservoir permeability and compressibility, the productive period can be very short. With reinjection he shows that an economic production rate can be sustained for over 30 years. During the initial pressure depletion period, however, only a small fraction of the produced brines can be reinjected. Pressures required for higher injection rates will overpressure the formation and possibly cause fracturing.

Hubert and Willis⁽¹⁸⁾ estimate the overpressure that will initiate fracture from:

$$\Delta P = (P_{ob} - P)/3 \dots \dots \dots (2)$$

where ΔP = additional pressure to initiate a fracture

P_{ob} = overburden pressure

P = pore or fluid pressure (geopressure)

In geopressed formations, the overpressure, ΔP , can be relatively small. To protect against reservoir overpressuring, Garg's reinjection

scenario starts at a very low injection rate and increases to 100% as the reservoir pressure is drawn down at the injection well. As Garg⁽¹⁴⁾ points out for some reservoirs, the production rate will drop below its initial value before production and injection rates stabilize at equivalent rates.

The well can now produce at this rate until the spent brine ultimately reaches the production well. At this point, a several fold increase in the resource recovery will have been realized over that without reinjection.

Production rates are also affected by whether or not thermal energy is extracted from the fluids. Without thermal energy extraction the dynamic pressure losses are less because of lower fluid viscosity. This permits full reinjection sooner at greater production rates. The net result is higher resource recovery as well as reduced injection pump energy demands per barrel of fluid.

Unlike shallow injection into an overlying unconnected formation, maintenance of high wellhead pressure to supplement injection into the producing formation does not decrease net injection pumping energy. The net pump energy depends on dynamic pressure losses in the well casing and formation, which remain unchanged provided flow rate and fluid and reservoir parameters remain constant. At steady conditions the pressure field in the reservoir will rise and fall accordingly with wellhead pressure changes such that $P_r - P_s$ (see equation 1) remains constant. Therefore, in the deep reinjection scenario, the most effective mode of operation is to deplete the reservoir to the lowest wellhead pressure possible to maximize CH_4 recovery and thereby preclude the need for a high pressure CH_4 recovery process. However, it is recognized that potential changes in reservoir formation properties and/or gas evolution in the formation with steep drawdown could seriously reduce productivity.

Summary

1. Reinjection into the producing reservoir is necessary to recover a sizable fraction of the resource. Full reinjection will not be

possible at early times without exceeding the formation pressure and possibly fracturing the formation. Initial low injection rates into the producing reservoir can increase progressively as the reservoir is drawn down.⁽¹⁴⁾

2. When full injection into the producing reservoir is reached, the pressure gradient in the reservoir becomes established and fixed. This sets the conditions for computing the injection pressure that the surface pump must develop. For the reservoir previously described, about 3000 to 3500 psi is required for injection at an energy expenditure equivalent to 15-20 scf CH₄/bbl.
3. The thermal energy of the brine can be used to generate about 2/3 of the injection power requirements. However, the estimated cost of power generation (65 mills/kW-hr), even when we assign zero cost for brine production, exceeds the cost of power available from conventional sources.
4. At a fixed production rate with full reinjection, the pressure gradient between the production well and the injection well is insensitive to wellhead pressures. The most effective mode of production is to draw down the reservoir, and to maintain low wellhead pressure (without flashing) to maximize CH₄ exsolution and recovery. There is no requirement for a high pressure CH₄ recovery process.
5. During the early phases of production when the reservoir may not be able to accept full injection, part of the produced fluid must be injected elsewhere - presumably into shallow aquifers. Available wellhead pressures can be used for driving shallow injection. Methane still dissolved at these surface pressures can be recovered by solvent extraction as previously described.

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APPENDIX A

Solvent Extraction Cost Estimate

The equipment for CH_4 recovery by solvent extraction and the economic analysis for such an operation were based on the following:

The plant is sized to process 40,000 bbl/d (1167 gpm) of brine delivered at 1000 psi and 150°C. At these conditions we assumed a dissolved CH_4 concentration of 10 scf/bbl. Based on a distribution coefficient of ~ 25 for the CH_4 , hexadecane, brine system, 75 gpm of hexadecane is required at roughly equilibrium conditions at the exit of the extraction tower. The process recovers 90% of the dissolved CH_4 and delivers it at 100 psi.

Extraction Tower

To achieve a brine velocity of about 0.1 ft/s (about half the dispersed phase (bubble) rise velocity), a tower diameter of 6.1 ft is required. The height of the tower of 20 ft is a best estimate until laboratory process data is available on mass transfer characteristics. The wall thickness for the tower operating at 1000 psi would be 4.1 inches. The cylindrical shell weights 6.6×10^4 lbs. Add to this the weight of the flared bottom section, ends, and extra thickness for corrosion allowance, we obtain a total weight of 10^5 lbs. The total vessel fabrication cost at \$2/lb is \$200,000.

Solvent Circulation Pump

Pump equipment costs are estimated to be \$10,000. The pumps are sized to circulate hexadecane at a rate of 75 gpm and a ΔP of 900 psi.

The O & M costs are estimated from the relationship derived by Doscher⁽⁷⁾:

$$\$/\text{bbl} = \$0.02 + .005 (\$/\text{Mcf}) P \dots \dots \dots (\text{A1})$$

where $\$/\text{Mcf}$ is the selling price of CH_4 , and P is the pump differential pressure in 10^3 psi. The $\$0.02$ accounts for pump maintenance costs, while the second term on the right represents the power costs which are proportional to CH_4 prices and pressure. For $\$5/\text{Mcf}$ methane, the annual O & M solvent pump costs are:

$$[\$0.02 + .005 (\$5) (.9)] \frac{75}{42} \times 1440 \times 365 = \underline{\$40,000}.$$

Hexadecane Heat Exchanger

Although the vapor pressure of hexadecane at 150°C is low (about 10 torr), the hexadecane concentration in the CH_4 gas stream at 100 psi and 150°C will be about 2000 ppm. To reduce the hexadecane vapor losses to an acceptable level, the temperature must be dropped to less than 50°C . The heat load in cooling 75 gpm of hexadecane by 100°C is 2.7×10^6 Btu/hr. Assuming that the overall heat transfer coefficient is $100 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}$ and cooling water is available at 75°F and exits the heat exchanger at 100°F , the heat transfer area required is a modest 270 ft^2 . The exchanger estimated cost is $\$10,000$. The heat exchanger cooling water requirement is 216 gpm. Assuming that a water source and sink are available, we estimate $\$10,000$ for the pump cooling system and piping. O & M costs are a nominal $\$1,000$ annually.

Methane Desorption Vessel

We assume that desorption of CH_4 from the hexadecane solvent is rapid, and separation can be effected using a cyclone separator operating at 100 psi. The two-phase volumetric flow into the separator is $0.7 \text{ ft}^3/\text{sec}$, i.e., $0.16 \text{ ft}^3/\text{sec}$ of solvent and $0.54 \text{ ft}^3/\text{sec}$ of CH_4 . The CH_4 gas stream velocity exiting the separator must be kept low to minimize entrainment of the solvent. An adequate separation should be achievable with a 3 ft. diameter 10 ft tall cyclone separator with a gas outlet port 2 ft in diameter. A demister is also included. A half-inch

thick steel vessel with hemispherical ends will weight about 2500 lbs. At \$2/lb, the fabrication cost will be about \$5,000.

Total Capital Costs

The total installed equipment cost is twice the fabricated cost.

extraction tower	400,000
heat exchanger	20,000
pumps	40,000
desorption vessel	10,000
miscellaneous equipment	70,000
	<hr/>
	\$ 540,000

A hexadecane inventory is required at a cost of \$4/lb or a total cost of \$250,000. The total installed capital investment is \$790,000.

Total O & M Costs

Hexadecane losses from brine dissolution are estimated to \$20,000 annually. Pump O & M is \$41,000. Personnel costs should be minimal with a fully automated plant. We estimate \$20,000 per year for operating personnel. Total O & M is, therefore, \$81,000.

APPENDIX B

Cost Estimate of Injection by Mechanical Pumping

Conventional depressurization operation costs were based on the following:

Capital Costs

Total installed costs for an electrically driven injection pump for 1000 psi, 150°C operation, for 40,000 bbl/day disposal rate is \$150,000.

Operation and Maintenance Costs

The operation and maintenance costs are based on the previous relationship (A2).

$$\begin{aligned}\text{O\&M Costs} &= [\$0.02 + .005 (\$/\text{Mcf}) P] 40,000 \times 365 \\ &= \underline{\$620,500} / \text{year}\end{aligned}$$

Operating personnel costs are negligible compared with pump operation and maintenance costs.

For conventional depressurization with 65% hydraulic energy recovery, the following assumptions and calculations were used.

Installed Capital Costs

Pumps, 40,000 bbl/day.	\$100,000
Hydraulic turbine drive, 40,000 bbl.day, to supply 65% of required pump power. Costs assumed to be approximately equivalent to pump costs with 50% premium for direct pump coupling, speed controls, and non-standard equipment.	\$150,000

Electric motor drive to supply at least 35% of power requirements	\$ 30,000
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Total installed capital costs	<u>\$280,000</u>
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Operation & Maintenance Costs

Pump maintenance (from A2)	\$292,000
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(\$.02) (40,000 bbl/day) x 365.

Hydraulic turbine maintenance (from A2)	\$292,000
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assumed to be equivalent to pump
maintenance.

Power costs (from A2)	\$115,000
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(.005) (\$5/Mcf) (0.9) 40,000 bbl/day x 365 x .35

Total Operation & Maintenance	<u>\$699,000</u>
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