CORE ANALYSIS ISSUES IN HEAVY OIL RECOVERY USING VAPEX

Lesley James and Ioannis Chatzis
Department of Chemical Engineering, University of Waterloo, Ontario, Canada

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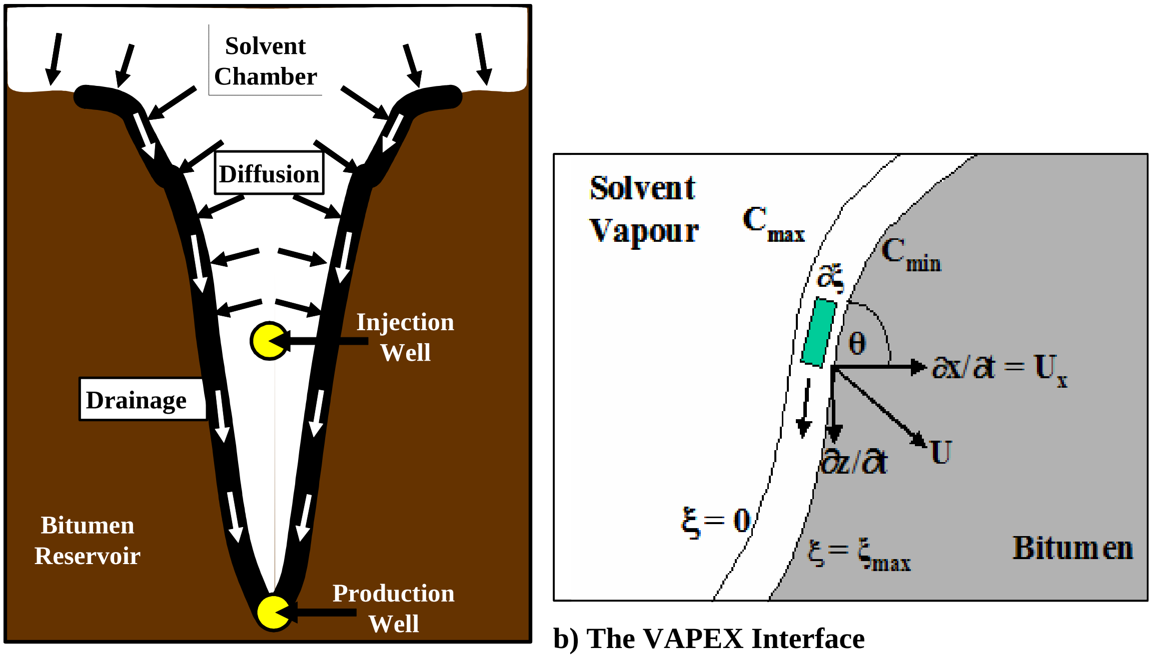
ABSTRACT

Heavy oil recovery using solvent based processes for in-situ recovery of very viscous oil and bitumen is an emerging new technology in Canada. The core analysis and testing of process conditions has not been standardized yet. This paper aims to provide a review of the various studies that aimed to collect lab-scale data and upscale the results to field-scale conditions. A summary of the research results obtained in our laboratory over the past three years are presented and discussed. Results obtained in our research and by others demonstrate that heavy oil recovery experiments in long enough systems are desirable for predicting production rates. Analysis of production history shows that the dispersion coefficient is orders of magnitude larger than the diffusion coefficient measured in stagnant heavy oil.

INTRODUCTION

VAPour **EX**traction (VAPEX) is an enhanced oil recovery technique developed primarily by Canadian researchers, for the recovery of Canada's vast heavy oil reserves. Conventional oil recovery practices can not be used for the production of heavy oil and bitumen because the viscosity of heavy oil is too high and it does not readily flow under normal reservoir conditions. The viscosity of heavy oil is generally between 1000 mPa.s and 10,000 mPa.s while the viscosity of bitumen is greater than 10,000 mPa.s. The primary driving forces in VAPEX are the mass transfer of the solvent into the heavy oil followed by the gravity drainage of the "live oil" (viscosity reduced heavy oil containing the dissolved solvent) to the production well. Normally, the VAPEX process is represented with two horizontal wells; the upper one used for solvent injection and the lower one for producing the live oil. First, the solvent is injected into the injection well where it diffuses through the heavy oil primarily in the upward direction until it encounters an impermeable barrier zone, the "cap rock". The diffusion of the solvent into the heavy oil/bitumen reduces the oil's viscosity, thus mobilizing it under the force of gravity to the production well. Upon encountering the cap rock, the solvent continues to diffuse into the heavy oil but in an outwards direction in a continuous cycle of solvent diffusion into the heavy oil, viscosity reduction and live oil drainage. Figure 1 shows the live oil being mobilised by the solvent diffusing into it due to the concentration difference of the solvent. As the live oil with significant solvent in it drains, a new layer of heavy oil is exposed to the solvent phase at the VAPEX interface. The new layer is subjected to the same mass transfer driving force that continues the cycle of essentially peeling away layers of heavy oil. In this manner, the VAPEX interface advances through the permeable zone saturated with heavy oil.

VAPEX technology competes with Steam Assisted Gravity Drainage (SAGD) technology which is currently used for heavy oil recovery. At present, there have been no VAPEX field trials in



b) The VAPEX Interface

a) The VAPEX Process

Figure 1: Schematic Diagrams of the a) Overall VAPEX Process and b) the Transport at the VAPEX Interface

Canada. VAPEX is often referred to as the "solvent analog" to SAGD as mass and heat transfer concepts can often be interchanged. While SAGD relies on heat transfer and mass transfer to lower the viscosity of the heavy oil, VAPEX relies on mass transfer. Both methods involve diffusion and convective mixing due to the development of the solvent/steam chamber above the production well. However, VAPEX can be used in thin shallow reservoirs with underlying aquifers while SAGD becomes inefficient in such situations due to excessive heat losses. VAPEX offers the additional advantages of being a closed-loop process where the solvent can be reclaimed and reused, VAPEX requires less water and energy, and if VAPEX induces asphaltene precipitation the oil can be upgraded in-situ thus reducing diluents and upgrading costs.

Since the early 1990's there has been much research in the area of VAPEX. With laboratory scale VAPEX experiments, various authors studied VAPEX in different types of physical models with different physical properties including permeability, model height, dip angle, different viscosity oils, underlying aquifers, connate water and different solvents in liquid or vapour state with various partial pressures. Some of the experimental results have been used to infer and quantify the mass transfer mechanisms in VAPEX. There has been much discussion around the role that molecular diffusion and convective mixing plays in the VAPEX process. Presently, it is widely accepted that both play a significant role but neither has been quantified correctly. Poor understanding of the mass transfer mechanisms and the limited understanding of the pore scale events have been the primary reasons that numerical modelling has failed to satisfactorily match the laboratory scale VAPEX experimental results.

The objective of this paper is to review the various work contributing to field-scale predictions, identify core analysis issues with the VAPEX process such as maintaining pressure, solvent condensation and the use of non-condensable gas and the rate of VAPEX interface advancement.

A summary of the research results obtained in our laboratory over the past three years is also presented and discussed.

VAPEX Research in Laboratory Scale Models

Experimental VAPEX research has been ongoing since the late 1980s when Butler and Mokrys (1989) first tried VAPEX using Hele-Shaw cells as a solvent analog to steam assisted gravity drainage (SAGD). The Hele-Shaw cell represented a very simplistic model where the heavy oil saturated the void space between two sealed glass plates separated by a predetermined thickness. Laboratory scale VAPEX models used to determine the effect of permeability, heterogeneity, length, dip angle, the presence of underlying aquifers on oil production rates have all been explored, some more recently than others. This section summarises and compares the different experimental work carried out using laboratory scale VAPEX models. The historical results are described briefly while the recent literature is described in more detail. Most VAPEX experiments have involved macromodels where heavy oil is recovered from various shaped geometry systems filled with saturated porous media by injecting either vapour or liquid solvent. The following summarises the majority of VAPEX experiments carried out to date and describes the VAPEX system used, the solvent used along with its state and method of injection and goals.

Butler and Mokrys (1989) using a Hele-Shaw cell [7cm x 2.6cm x 7cm wide] observed asphaltene precipitation and scale-up equations were proposed for production rates. Butler and Mokrys (1993) used 2-D porous media models [21.7cm x 3.46cm x 69.8cm] with1 mm beads ($K = 1136\ \mu m^2$ and $\phi = 0.39$) and Lloydminister (Tangleflags) heavy oil ($\mu = 10{,}000$ mPa.s at 20^oC, 16.6 wt% asphaltenes)**.** Solvents used were: 1) hot water & propane (constant injection), 2) propane, and 3) ethane. The hot water and propane results (Pressure = 1.34 – 1.38 MPa ; T = 42 – 47^oC) have shown two distinct constant production rates with 60% recovery in seven hours. A higher rate equivalent of 867 bb/day from 1h46min to 2h36min and a lower rate equivalent to 260 bb/day from 2h36min to 6h. Some asphaltene precipitation was observed. With propane (Pressure = 0.888 – 0.918 MPa; T = 26 - 30^oC), two distinct scaled-up recovery rates were noted with 66% recovery in eight hours. Production started at an equivalent rate of 270 bb/day for the first 3 hours and ended at 126 bb/day from 3 to 8 hours using a 10m pay zone and a 457m horizontal well as field scale dimensions. Oil was upgraded from 10,000 to 2,000 mPa.s. With ethane**,** two distinct scaled recovery rates were noted with 26% recovery in seven hours. Production started at 243 bb/day for the first 2 hours and ended at 66 bb/day from 2 to 8 hours using the same equivalent field dimensions; no asphaltene precipitation was seen.

Jiang and Butler (1996) used a 2-D Packed Model: [35.6 x 3.2 x 21.6 cm] with heterogeneities and having alternate layers of 20-30 mesh Ottawa Sand (coarse, $K=214\ \mu m^2$) and 30-50 mesh (fine, $K = 43\ \mu m^2$ $\phi = 0.35 – 0.37$). The Heavy oil Lloydminster (Tangleflags) $\mu = 7{,}400$ mPa.s at ambient and n-Butane at $T_W \sim 20^oC$, $T_{ambient} = 21\text{-}23^oC$. Oil production rates were lower for heterogeneous layered systems compared to homogeneous systems with same permeability. Vertical fractures and solvent injection from the top of the system both result in higher oil production rates due to better communication between the solvent and heavy oil. The solvent chamber easily extends around discontinuous lenses of low permeability. The solvent chamber advances into the higher permeability layers of the continuous layered model while growth into the low permeability layers is retarded by capillarity.

Das and Butler (1998) used Hele-Shaw cell with Peace River and Lloydminster heavy oil using propane and n-butane. The permeability varied from $K = 1344$ and $5400\ \mu m^2$. The 2-D packed model [34 cm x 3.2 cm x 22 cm]with heavy oil was extracted using Butane. Models with Ottawa sand / glass beads with permeability K = 27, 43.5, 217, 432 μm^2 and porosity $\phi = 0.35 - 0.37$. Heavy oil (μ at 20^oC)**:** Peace River, μ = 130,000 mPa.s and Lloydminster, μ = 10,000 mPa.s. Solvents used were: pure propane & butane vapour below and near their vapour pressures. In Hele-Shaw cells, higher production rates were observed at higher solvent partial pressures below the vapour pressure. When solvent at its vapour pressure was used deasphalting occurred in regular patterns in the Hele-Shaw cell. Production rates using propane and Peace River oil were increased by 35% after the onset of deasphalting. The production rates from the saturated 2-D models were found to be almost 10 times higher than expected from Hele-Shaw cell scale-up predictions. They hypothesised that production rates are higher in porous media due to increased surface area in the pores and the cross-current flow of the solvent and live oil drainage, increased imbibition of the live oil and higher spreading coefficients of the live oil over connate water compared to the heavy oil. They also suggested the use of an effective diffusivity taking into account the ratio of the diffusion area to the fluid flow area to account for differences between experimental and predicted production rates.

Oduntan (2001) used 2-D troughs [1.9 cm x 1.6 cm x 21 cm, 42 cm, 84 cm, 160 cm, and 247 cm]. Unconsolidated glass beads with: K = 25, 85, 136, 192 Darcy; Dip Angle = 45^o, 90^o and ϕ = 0.38 and heavy oil μ = 40,550 mPa.s at ambient. The n-butane was supplied at P_{vap} at T = 19-22^oC (P_{vap} = 16 – 18 psig). It was found that $Q \propto \sqrt{KH}$, where H = length of the VAPEX interface. The live oil production rate was found to be constant for most of the experimental time. Residual oil saturation was found to be 3-5% OIIP after blow down. Live oil properties were found to be essentially independent of packing length and permeability. Viscosity of the live oil ranged between 2 - 3 mPa.s. Butane in the live oil was between 0.40 - 0.42 mass fraction and density of the live oil ranged between 0.70 - 0.87 g/cm^3. Heterogeneous models exhibited higher residual oil saturation and lower live oil production rates compared to equivalent permeability homogeneous systems. Asphaltene precipitation was observed near the production end.

Ramakrishnan (2003) used 2-D Troughs [1.9 cm x 1.6 cm x 84 cm] packed with unconsolidated glass beads with permeability K = 25, 86, 156, 220 Darcy, and porosity ϕ = 0.38).The dip angle was varied from 45^o, 75^o, 80^o, 90^o**.** The heavy oil had μ = 70,000 mPa.s at ambient. Solvent was propane supplied at P_{vap} (105 – 120 psig); $T_{WB} = 20\text{-}24^oC$, $T_{ambient} = 21\text{-}26^oC$. Live oil production rates were found to be constant. Residual oil saturation (SOR) was found to depend on the dip angle. SOR for 45^o dip angles ranged from 10-13% whereas SOR for larger dip angles was found to be 5-7% pore volume after blow down. Viscosity of the live oil appeared to depend on permeability and was found between 2.5 – 5 cP.

James (2003**)** used consolidated glass beads and heavy oil with μ_O = 85,000 cP at ambient. The solvent :n-butane was supplied at its vapour pressure at water bath temperature($T_{WB} = 21.5^oC$ and 23.0^oC, P_{vap} = 16.6 psig and 19.0 psig). Live oil production rates were found to be constant for a given system and production rates varied linearly with the length of the system and were proportional to sine of the dip angle (sin θ) (see Fig. 2). The live oil density and solvent concentration was independent of length and dip angle with a density ranging from 0.81 – 0.82 g/cm^3 and the solvent concentration ranging from 0.27 – 0.32 mass fraction. The residual oil saturation decreased from 22% to 10% P.V. as the length of the system increased.

Cuthiell et al. (2003) performed Liquid Extraction in 2-D Models[25 cm x 30 cm x 2.8 cm] Silica, sand and mixture of both with K = 8, 88, 90 Darcy. They used Aberfeldy Lloydminster oil μ = 5500 cP at 25^oC and liquid toluene as extraction solvent and examined the diffusion/dispersion phenomena and viscous fingering in liquid toluene extraction of heavy oil using a CT scanner. The goal was to reproduce experimental results using STARS reservoir simulation software to predict field scale rates. X-ray images showing the solvent chamber growth and viscous fingering including some internal mixing in the dominant finger. The dispersion coefficient was found by choosing different dispersion coefficients and simulating the results in order to qualitatively match the viscous fingering observed experimentally and quantitatively matching post-breakthrough production rates using the Butler and Mokrys model. The dispersion coefficient that matched the experimental results was consistent with the transverse dispersion coefficient by Blackwell (1962).

Yazdani and Maini (2004) used rectangular channel (R) &annulus (A) type porous media. Dimensions of models were R(7.5x11.3x2.5), R(15x22.5x2.5), R(30x45x2.5), A(30x42.3x3), and A(60.1x84.6x3.2 cm). The permeability K = 220, 330, 640 Darcy. They validated the production rates between equivalent rectangular and annulus systems. Their goal was to verify the scale up predictions. The authors performed VAPEX experiments using three different permeabilities in the annulus model and results from the rectangular model to confirm that the dead oil production rate, $Q = f(\sqrt{K})$. They performed experiments using the same permeability in the different systems but could not confirm the square root dependence of production rates on system height. They observed that $Q \propto H^{1.1-1.3}$. Talbi and Maini (2004) used an annulus geometry with saturated porous media (30.48 cm high, 3.5 cm width packed with 12-16 US mesh glass beads, K = 640 Darcy $\phi = 0.35$). Two kinds of oil($\mu_1 = 4500$ mPa.s at 21°C and $\mu_2 = 18{,}600$ mPa.s at 21°C) and solvents were used. Constant injection of 1) propane (40ml/h) & CO_2 and 2) propane (40ml/h) & CH_4 at T = 21-25°C was used. Use of CO_2 results in higher production rate at mid to high operating pressures while CH_4 is more efficient at low pressures as shown in table below.

Oil	**NCG (non-cond. gas**	**P (psig)**	**Duration (hours)**	**Oil Rate (ml/h)**	**Recovery**
1	CO_2	250	9	85.69	47.5%
1	CO_2	250	9	85.42	46.9%
1	CH_4	250	9	94.85	52.4%
1	CH_4	600	9	59.56	32.1%
1	CO_2	600	9	79.11	42.5%
2	CO_2	200	12	34.33	25.1%
2	CH_4	200	12	41.51	29.8%
2	CO_2	400	12	40.76	30.0%
2	CH_4	400	12	36.03	25.8%
2	CO_2	600	12	44.26	32.7%
2	CH_4	600	12	29.58	24.3%
2	with CO_2 only	600	12	21.69	15.6%

Issues with Scale-Up

The ultimate goal of laboratory scale VAPEX experiments is to enable the prediction of field scale production rates and sweep efficiency. The Butler and Mokrys (1993) scale-up prediction takes into account the permeability (K_m) and height (H_m) of the model in order to predict the field scale production rate (Q_f). The model was derived from a one dimensional steady state balance on a unit volume of heavy oil accounting only for the molecular diffusion of solvent in the heavy oil. Later, Das and Butler (1996) added an effective diffusivity including a cementation factor (m), as shown in the following equation, to account for overall mass transfer and matched production rates observed from experiments using porous media.

$$Q = 2\sqrt{2KHg\phi^m \Delta S_o N_s} \tag{1}$$

$$N_s = \int_{c_{min}}^{c_{max}} \frac{\Delta\rho(1-c_s)D_{eff}}{\mu} d\ln c_s \qquad \text{where} \quad D_{eff} = \left(\frac{A_d}{A_f}\right) D\phi^m \tag{2}$$

Oduntan (2001) examined the effect of both permeability and system height on live oil production rates from unconsolidated porous media. He used 1.9 cm width x 1.6 cm deep x 84 cm long aluminium troughs packed with saturated unconsolidated glass beads of different diameters to examine the effect of permeability on production rates. The 1.9 cm x 84 cm face was exposed to butane at its vapour pressure. He also examined the effect of system height at 45° dip angle using the same size glass beads and varied the length. The results of laboratory scale experiments indicated that production rates are proportional to the square root of permeability and height.

$$\frac{Q_F}{Q_M} = \sqrt{\frac{(K\,H)_F}{(K\,H)_M}} \tag{3}$$

James (2003) used consolidated porous media to investigate the effect of system height on live oil production rates. Glass beads, approximately 3 mm deep, were sintered together on glass plates and cut into test strips of approximately 4.8 cm wide and various lengths. The production rates were found to be proportional to height, as shown in Fig.2. The resulting scale-up for consolidated media would be: $\frac{Q_F}{Q_M} = \frac{(HW)_F}{(HW)_M}\sqrt{\frac{(\phi K \Delta S_o)_F}{(\phi K \Delta S_o)_M}}$ (4)

Yazdani and Maini (2004) used the annulus between two concentric cylinders to pack and saturate unconsolidated glass beads to examine the effect of system height and permeability on dead oil production rates. They confirmed that production rates were proportional to the square root of permeability using systems with 220, 330 and 640 Darcy permeability respectively. However, they were unable to replicate the square root dependence on system height. Using the results from two different annulus height trials (30 cm and 60 cm) and three heights in a rectangular system (7.5 cm, 15 cm and 30 cm), the dead oil production rates were proportional to the system's height to a power n = 1.1 – 1.3: $\frac{Q_F}{Q_M} = \left(\frac{H_F}{H_M}\right)^n \sqrt{\frac{(K\phi^m)_F}{(K\phi^m)_M}}$ (5)

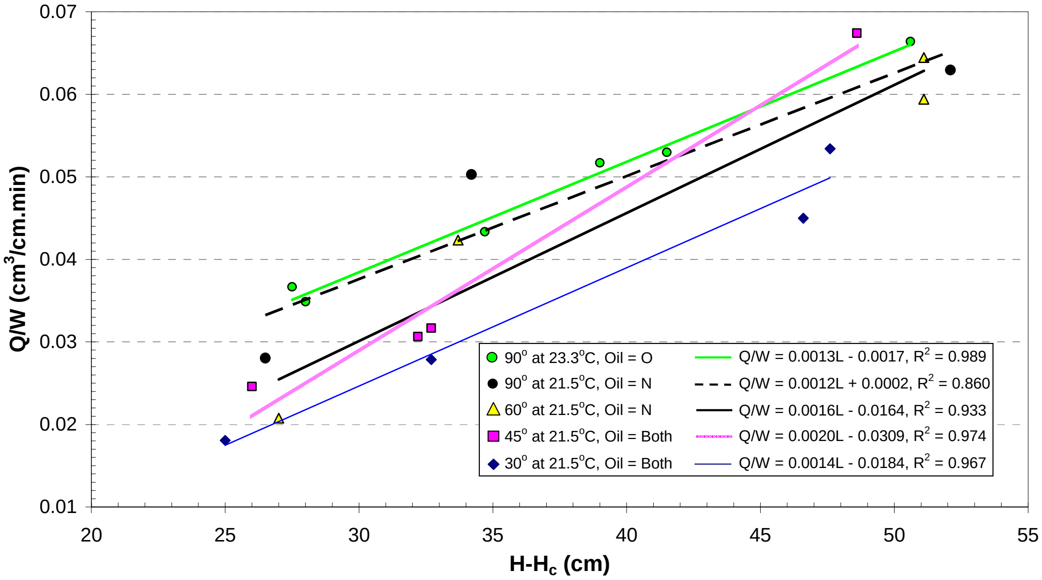


Figure 2: Effect of length on live oil production rate in consolidated media (James, 2003)

James et al. (2003) drastically changed the system permeability in two unconsolidated models of similar physical dimensions in order to observe the effect on live oil production rates and the shape of the VAPEX interface. The VAPEX interface is usually described as "S" shaped where the interface advances deeper into the bitumen phase at the top of the model (or formation) and less at the bottom. They observed that in models approximately 1 cm wide, 6 cm deep and 96 cm long that although the shape of the VAPEX interface changes with time, the equivalent length of the VAPEX interface remains fairly constant over the time that live oil production rates are constant. It is not until the majority of the oil is produced and the VAPEX interface length shortened to approximately half its original length that oil production rates slow down. From this observation, it is believed that a better understanding of the way the VAPEX interface advances throughout the model will aid in developing a more robust model to predict field scale production rates which take into account system's height, permeability and the velocity of VAPEX interface.

VAPEX Interface Advancement

The VAPEX interface velocity (V_I) may help clarify the dependence between oil production rates and height of the system. It is defined as the speed at which the moving VAPEX interface boundary moves into the bitumen in the horizontal direction at any vertical location in the system. Figure 3 shows a schematic of a typical VAPEX model indicating the solvent chamber, the direction of the VAPEX interface advancement, etc. The characteristic "S" shape indicates that the VAPEX interface advancement rate is greater at the top of the models than closer to the bottom of the system. Experimentally, it has also been observed that once the VAPEX interface reaches the depth of the system at the top (impermeable vertical wall), the solvent chamber not only grows from the horizontal advancement of the VAPEX interface but also from the vertical decline of the bitumen as well. The rate at which the VAPEX interface descends (V_z) may also be important in trying to understand and mathematically model the VAPEX process. As indicated in figure 3, we monitor the VAPEX interface as a function of time, thus making it possible to determine the velocity of VAPEX interface as a function of location, both in the horizontal (V_I) and in the vertical (V_z) direction from plots of interface position versus time.

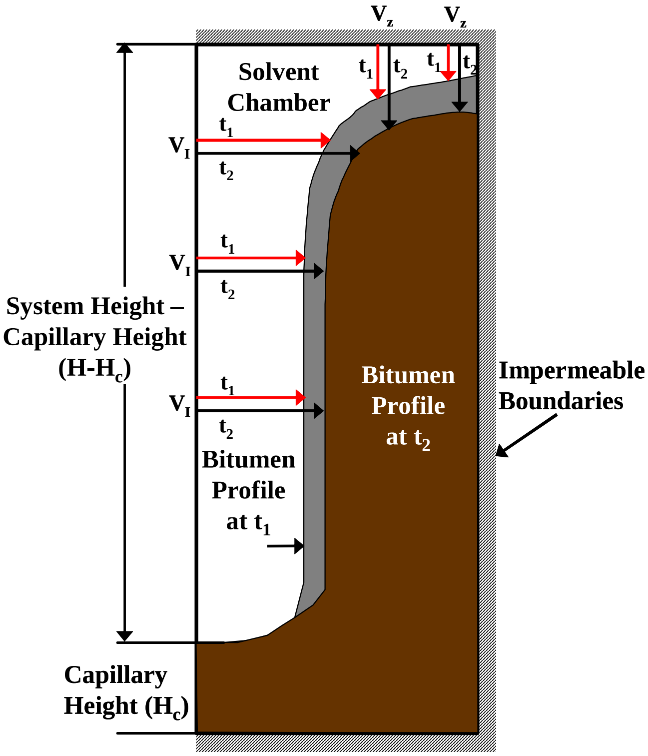


Figure 3: Schematic of Moving VAPEX Interface in Laboratory Scale Models

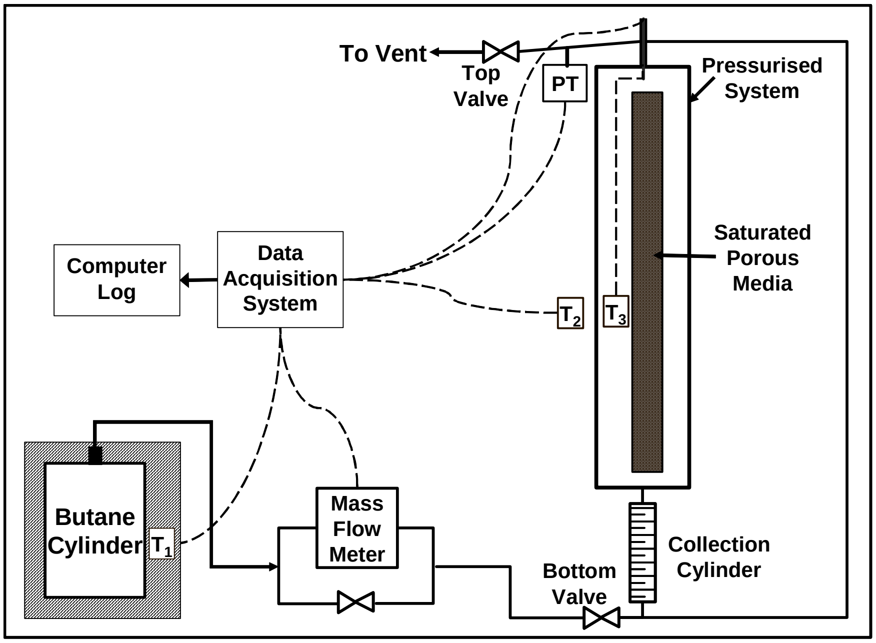


Figure 4: Experimental Set-up for Lab Scale Macromodel Experiments

The experimental set-up used to examine the VAPEX interface advancement is shown in Figure 4. As shown, the cumulative live oil production was measured volumetrically in the collection cylinder, thermocouples and a pressure transducer attached to a data acquisition system logged the water bath, ambient and system temperatures along with the system pressure. The advancing VAPEX interface was measured every 30-45 minutes at several vertical locations along the model where clear plastic rulers with 1 mm increments were affixed. Using the same time intervals, the entire VAPEX interface profile was traced on paper. Figure 5 shows the live oil production rates (Q) for the 45 cm and 92 cm models. The 92 cm model was performed a second time due to poor water bath temperature control during the first trial. In comparing the 45 cm model and the 92 cm model run 2, it is seen that the live oil production rate is proportional to the system height. When the system height is doubled the live oil production rate is also doubled. A height dependence on production rate is physically realistic. Doubling the height would double the VAPEX interfacial area exposed and the volume of heavy oil available assuming all other system parameters are constant.

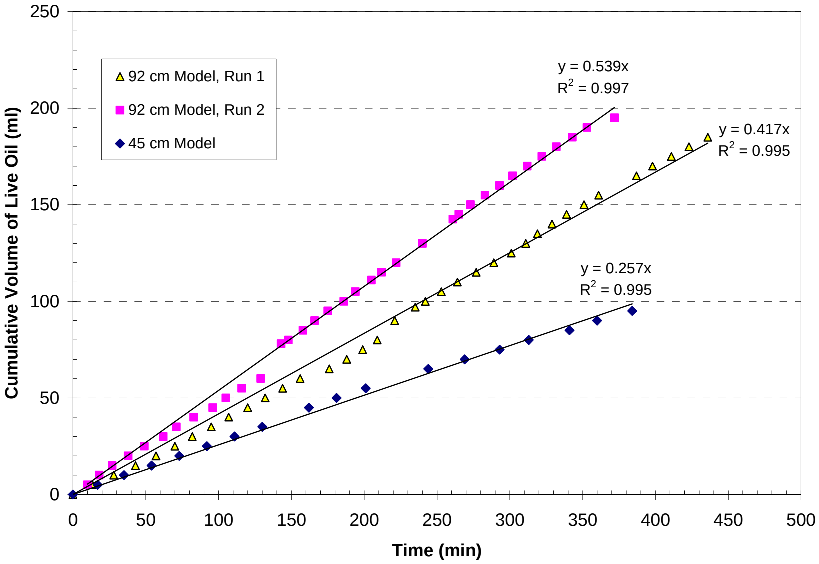


Figure 5: Live Oil Production Rates for the 45 cm and 92 cm High Models

The VAPEX interface advancement is shown in Figure 6 for several locations along the 92 cm high model and the interface velocity V_I is found by fitting the data as shoun.. The figure indicates that the rate at which the VAPEX interface advances or sweeps through the reservoir is location dependent. The rate of VAPEX interface advancement at the top of the system is greater than at the bottom, i.e. $V_I = 0.127$ mm/min at 88.5 cm from the bottom whereas $V_I = 0.063$ mm/min at 29.5 cm from the bottom. As the solvent diffuses into the heavy oil, its viscosity is reduced enabling it to drain under the influence of gravity. At the top of the system, the live oil drains freely and diffusion is not inhibited by a falling live oil film, which essentially shields the bitumen from solvent diffusion. It is the cascading live oil film which results in lower rates of VAPEX interface advancement at locations closer to the bottom of the system. In analysing the V_I rates in Figure 6, the live oil production rates and the traced VAPEX profiles, it is evident that there is a delay in producing oil further away from the top of the system. It is due to the ceiling effect that live oil is not produced instantaneously from the system as it must first drain via gravity through the porous network to the production well. Figure 6 does not depict the constant VAPEX interface advancement which occurs in approximately the mid 40% of the model. Here, the VAPEX interface was observed to advance as a vertical line. Figure 6 shows that irrespective of the height of the model, the constant VAPEX interface starts occurring approximately 20%

from the top of the model. Figure 6 also shows that the constant VAPEX interface advancement is consistent between models, at an average $V_I = 0.54$ cm/hr. From the traced VAPEX profiles, the rate at which the VAPEX interface descended (V_z) from the top of the model at the back-end of impermeable wall is given in Table 1. These results show little difference in vapex interface velocity between the 45 cm and 92 cm high models. The conclusion from the observed V_I and V_Z values is that the growth of the VAPEX solvent chamber is very similar for systems of comparable physical dimensions and permeability, assuming the same solvent is used at the same operating conditions. The solvent chamber descends from the top impermeable wall and advances horizontally from the side exposed to solvent. The interface velocities V_I and V_Z do not appear to depend strongly on the length of macromodels used. The length of the exposed interface dictates the live oil production rate for systems of same permeability.

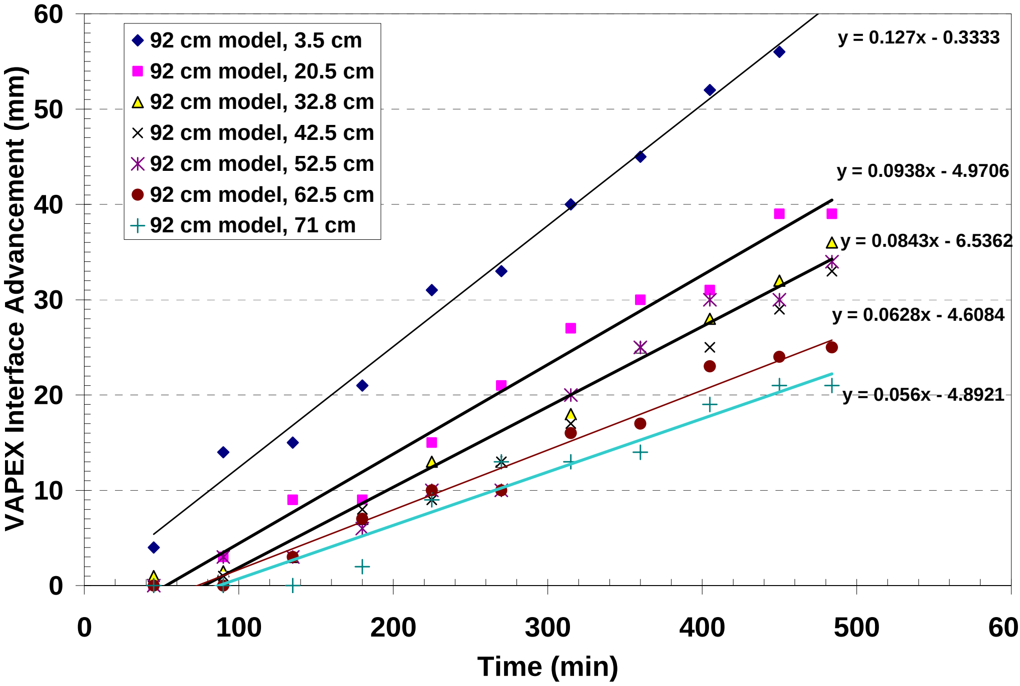


Figure 6: VAPEX Interface Advancement for the 92 cm High Model, Run 1

Table 1: Comparison of VAPEX Interface Advancement in Unconsolidated Models

Model	**% of Model Height at Top with non-Constant V_I**	**Constant V_I (cm/hr)**	**V_z (cm/hr)**
92 cm Model, run 1	22	0.54	NA
92 cm Model, run 2	18	0.50	1.1
45 cm Model	22	0.58	1.2

James and Chatzis (2004) studied the VAPEX interface velocities using different patterned glass etched micromodels. The velocity of VAPEX interface advancement observed in the glass micromodels was in the range of 0.11 to 0.19 cm/hour depending on the pattern of the micromodel. The permeability of the unconsolidated porous media using 700 micron glass beads is approximately 350-400 Darcy which is approximately three times that of micromodels OC-1 and DC-1 but approximately the same as micromodel DL-1. However, the VAPEX interface advancement in DL-1 was on average only 0.11 cm/hr. In comparing the VAPEX interface advancement rates in the micromodels and the unconsolidated porous media it is important to consider the dimensionality of the model as well as the pore to pore distance, the flow path length and the diffusion distance. The pore to pore distance ranged from 1.5 to 1.8 mm in the micromodels whereas the particle diameter (and pore to pore distance) in the unconsolidated packed glass beads was approximately 0.7 mm.

Comparing Laboratory Scale VAPEX Results

Comparisons with respect to laboratory scale VAPEX experiments are somewhat difficult to compare between research groups. Besides the authors trying different physical models due to physical constraints such as visualisation and maintaining safe operating pressures, the biggest difference is the injection of solvent. Some authors (Talbi and Maini (2004), Yazdani and Maini (2004), Das and Butler(1996)) inject the solvent at a constant rate whereas in our laboratory (James (2003), Ramakrishnan (2003), Oduntan (2001)) solvent is delivered at its vapour pressure. Constant rate injection may cause the solvent to condense if it is not mixed with a non-condensable gas and the solvent mass transfer rate into the bitumen is less than the injection rate. Solvent condensation is not desirable. Another difference is the recording of oil production rates. In our lab, the live oil (oil containing diffused solvent) is volumetrically measured with time whereas Talbi and Maini (2004) and Yazdani and Maini (2004) record the dead oil production rate once the oil has been degassed for several days. It is extremely important to consider the experimental details when trying to compare the results between different research groups.

The main criteria in delivering the solvent to the in-situ bitumen are to maintain the highest possible mass transfer driving force and to use the least amount of solvent as possible for economic reasons. Mass transfer by diffusion, is optimal when the solvent is a vapour not a liquid and the partial pressure of the solvent is maximum, i.e. one or at its vapour pressure. If pure solvent (e.g. n-butane) is used then its vapour pressure is governed by the temperature of the liquid butane. If the butane vapour contacts a surface at a temperature less than the temperature of the liquid butane, the butane condenses. Condensation of the solvent is undesirable for two reasons; 1) liquid-liquid diffusion is slower and 2) more liquid solvent than vapour is required to fill the same pore space. Condensation of the solvent at the VAPEX interface does lead to asphaltene precipitation. In qualitatively analysing the experiments performed in glass micromodels, it was evident that asphaltene precipitation only occurred in the micromodel when the butane vapour condensed at the VAPEX interface. Experimentally, this is rectified by keeping the butane at a lower temperature than the water bath. The butane vapour should not condense even if it is at the same temperature however the capillary pressure in the pore throats may be sufficient to condense it. In the field, where down hole pressures exceed the vapour pressure of the solvent, it may be necessary to mix the solvent vapour with a non-condensable gas. The non-condensable gas reduces the partial pressure of the solvent and therefore the concentration gradient (diffusion driving force) will be reduced as well.

Unresolved Issues with the VAPEX Process

While VAPEX research has made great progress in the past ten years, there are still several unresolved issues. The main point still unresolved is the role of mass transfer at the pore scale. Das and Butler (1996) showed that the molecular diffusivity alone could not account for the oil production rates from laboratory scale experiments. They hypothesized that dispersion and imbibition increase the rate of mass transfer. Chatzis (2002) and James and Chatzis (2004) observed that mixing occurs primarily due to the entrapment of solvent vapour below the draining live oil film. Explicit description of pore scale events combined with knowledge of the advancing VAPEX interface and solvent chamber growth should be incorporated in the mathematical description of the VAPEX process. All indications point out that VAPEX is an economically and environmentally good alternative to SAGD for the recovery of in-situ heavy oil and bitumen. The challenge remains to mathematically describe the advancing VAPEX interface and oil production rates in laboratory scale systems accurately. This will enable us to predict field scale response.

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