



**TECHNICAL
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Heavy and Extra Heavy Oil: Midstream processes and transportation

Master of Science in Petroleum Engineering

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*To my parents
and those who know how to love, believe and offer.*

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ABSTRACT

The gradually growing energy demands have created a large interest towards the vast unconventional oil reserves that include heavy oils, extra heavy oils and bitumens, the exploitation of which is particularly challenging. These hydrocarbons exhibit high Specific Gravity, high viscosity, low mobility and increased content in asphaltenes, resins and N, S, O heteroatoms.

This thesis is focused on the exploitation of heavy and extra heavy oils that have °API gravity coarsely from 20 to below of 10 and a viscosity between 100 and 10.000 cP as well as on the implemented or emerging techniques for their recovery, surface processing and transportation. The main characteristics of heavy and extra heavy oils are presented and analyzed with respect to the oil properties that affect their flowability during production. The detrimental effect of asphaltene, wax and scale precipitation, the role of naphthenates and naphthenic acids, the formation of stable emulsions and the foamy and slug flow tendencies during the production of heavy/viscous oils are also covered.

The recovery methods used for the extraction of heavy and extra heavy oils are presented. Special reference is made in enhanced oil recovery methods including both thermal and non-thermal EOR techniques. The oil treatment is described in detail with the available options which include mechanical, thermal, chemical and electrostatic treatment or, more often, a combination of these. Electrostatic treatment is especially highlighted due to the variety of applications in dehydration and desalting, the energy efficiency and the ability to reduce the dosage on chemical demulsifiers. Additionally, the advances attained in the past fifteen years have allowed the use of the electrostatic separation systems in first-stage separators or pipeline segments.

Finally, a review of the available technologies for pipeline transportation of heavy and extra heavy crude oils is conducted. The review is based on viscosity and friction reduction options as well as on the in situ and partial upgrading on site. The fundamentals of mechanical separation, as well as a number of case studies are cited in the appendices.

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INTRODUCTION

When the petroleum exploitation started all crude oils that were extracted were of conventional type. The term “conventional” describes the oils that have low viscosity, less than 100cP, flow easily and have a density coarsely higher than 20 °API. API density is commonly used to classify the oil and it is related to specific gravity by the expression:

$$^{\circ}API = \frac{141.5}{SG} - 131.5$$

The SG is measured at 60°F (15.6°C). For a long period of time the energy demand was satisfied by conventional resources with established procedures. As the demand in hydrocarbons continued to grow, so did the need to increase the amount of oil that could be extracted from each reservoir. That was accomplished with the implementation of various novel techniques described with the term “Enhanced Oil Recovery” (abbreviated EOR).

Gradually, with the energy demands growing even more, the interest towards the unconventional oil reserves increased respectively. The term “unconventional reserves” includes heavy oils, extra-heavy oils, bitumens, oils sands and shale oils. Potential exploitation of these reserves became of great significance but, in the same time, rendered conventional methods of production and processing insufficient. For this reason, the development of new technologies that will shift unconventional reserves to commercial resources that would be economically produced became vital.

This work is focused on the heavy and extra-heavy oils that have °API gravity coarsely from 20 to below of 10 and a viscosity below 10.000 cP, and it leaves out the solid hydrocarbon sources such as oil sands and shale oil. Heavy and extra heavy oils are characterized by high specific gravity, high viscosity, low H/C ratio as well as high contents of asphaltenes, resins, heteroatoms (N, S, O) and trace metals (Ni, V, Fe). The high content in asphaltenes and resins directly impacts recovery, transport and refining processes and a prediction of the probable conditions for the precipitation of these heavy components is needed.

The scope of the thesis is to present the up-to-date methods for heavy and extra-heavy oil separation and transportation. As the heavy and extra-heavy oils exhibit high viscosities makes both their recovery and processing a challenging task. Consequently recovery methods used widely in the case of conventional oils are not adequate in this case and new technologies must be established. Gravitational separation of gas, oil and brine is also enhanced by new advanced methods, while transportation via pipelines is facilitated with effective techniques.

The extraction methods used today for enhanced oil recovery (EOR) are based on three primary techniques: introduction of gas, heat or chemicals into the reservoir. The selection of the appropriate method depends on the properties of the reservoir fluids and the characteristics of the reservoir [1]. For Heavy and Extra-heavy oils, thermal methods are the most commonly employed ones in order to increase oil recovery. These methods are based on supplying heat to the reservoir through steam or water injection or by means of petroleum combustion inside the reservoir thus the improvement in oil recovery is due to the reduction of the oil viscosity and the improvement of its mobility. Examples of these methods, which are described in detail in the following sections, are Steam Assisted Gravity Drainage (SAGD), Cyclic Steam Stimulation (CSS), Vapor Assisted Petroleum Extraction (VAPEX) and Toe-to-heel Air Injection (THAI).

Separation of oil from water, gas and solids is what follows its extraction from the reservoir to the surface. For viscous oils separation of water is not an easy task: Heavy oil viscosity is generally high and the density difference between oil and water is small which, on its turn, renders gravity separation difficult. One additional difficulty is the formation of stable water-in-oil emulsions during the separation process. Conventional oil-water gravity separation depends on the density difference, the oil viscosity and the water droplet diameter. Oil-water separation through gravitation requires retention time for phase equilibration and while the retention time increases so does the required separator size. In the case of viscous oils the time needed for efficient oil-water separation is substantially increased compared to the one for lighter crudes. Common methods to assist gravity separation are by forcing water droplets to coalesce and increase their diameter, usually through electrostatic fields, by implementing centrifugal forces or by heating the mixture.

Transportation of heavy and extra heavy oils is also a challenging issue. Pipelines are considered to be the most convenient means for the transportation of crude oil. In the case of heavy and extra heavy oils, however, it is much more demanding due to their low mobility, risk of asphaltene, salt and/or wax precipitation and deposition on the pipeline wall surfaces, and corrosion issues. Consequently, problems related to multiphase flows, clogging of pipes, high pressure drops and unscheduled shutdowns are likely to occur. Methods for viscosity reduction, drag minimization, in situ upgrading and partial upgrading on site are currently implemented to mitigate such problems and to comply with pipeline specifications [2].

HEAVY AND EXTRA HEAVY OIL

DEFINITIONS

Petroleum is a naturally occurring mixture of hydrocarbons, generally in a liquid state, which may also include compounds of sulfur nitrogen oxygen metals and other elements [3], [4]. Different petroleum fluids may vary widely in specific gravity (API gravity), viscosity, the amount of residuum, as well as the content in metals such as Nickel and Vanadium. Metal containing constituents are measured in parts per million and in high concentrations occur in the more viscous crude oils.

Typically petroleum is found in the pores of sedimentary permeable rocks, the reservoirs. The most common reservoir rocks are sandstone, siltstone, limestone and dolomite and they have the capability to store and transmit fluids. The contained fluids are not only petroleum but also brine saturated with minerals and gas. The ability of the oil to flow in the reservoir depends on its viscosity and the permeability of the reservoir rock.

Petroleum components are separated into categories.

The hydrocarbon constituents are compounds of hydrogen and carbon and display great variation in their molecular structure. The simplest hydrocarbons belong to a large group of straight or branched chain-shaped molecules, the paraffins. Paraffins extend from methane, which is the main component of the natural gas, through liquids (C₅ to C₂₀) that are refined into gasoline and up to paraffin waxes (>C₂₀). The naphthenes is a series of ring-shaped hydrocarbons which extends from volatile liquids, to high molecular weight substances. Another group of ring-shaped hydrocarbons is the aromatics whose lighter compound is benzene.

The nonhydrocarbon constituents of petroleum include organic compounds of nitrogen, oxygen and sulfur as well as compounds containing metallic elements including iron, nickel and vanadium. Most of these impurities are removed during refining [5].

Another useful classification of petroleum compounds is a scheme specifically designed for phase behavior and precipitation/deposition issues. Generally there are two major groups in which petroleum compounds are separated, the volatile fraction C₆- and the relatively non volatile fraction C₆+

The C₆- fraction consists of all hydrocarbon and non hydrocarbon components with carbon numbers up to C₅. These also include all the isomers. All components of this fraction are recorded in literature.

C₆+ fraction on the other hand is far more complex due to the multiple isomer combinations of hydrocarbons with increasing chain length. This group of components is further classified as paraffins (P), naphthenes (N), aromatics (A), resins (R) and asphaltenes (A). The fraction of paraffins and naphthenes is also called the saturates (S) fraction [6].

API GRAVITY SCALE

In 1921, the American Petroleum Institute created the API Gravity Scale, initially to measure the specific gravity of liquids less dense than water, especially petroleum. Today the use of the API Gravity Scale is

extended to the whole range of specific gravity, is recognized by the petroleum industry and is widely used.

Specific gravity and API gravity evolve in opposite directions. That means the smaller the API the heavier the fluid.

According to the API scale and the American Petroleum Institute petroleum types are separated in:

- Light crude, which has API gravity greater than 31.1 °API, and specific gravity less than 0.87.
- Medium crude, which has API gravity between 22.3 and 31.1 °API, and specific gravity between 0.87 and 0.92 and
- Heavy crude, which has API gravity less than 22.3 °API, and specific gravity greater than 0.92.

Another definition for the heavy crude oils, which is used by the World Petroleum Congress and the US Geological Survey and is adopted in this work, considers as heavy crude oil the ones with API between 20 and 10.

A further classification between heavy crude oil, extra heavy crude oil and bitumen is given by the Canadian Center of Energy. This classification considers both API Gravity Scale and viscosity in reservoir conditions and it is the following:

- Heavy oil, the API degree of which is greater than 10 and less than 20, its viscosity is less than 10,000 cP (10 Pa.s) and it flows at reservoir conditions.
- Extra-heavy oil, the API degree of which is less than 10, the level of viscosity in reservoir conditions is less than 10,000 cP (10 Pa.s), which means that it has some mobility at reservoir conditions.
- Natural bitumen, the API degree of which is less than 10, the level of viscosity in reservoir conditions is greater than 10,000 cP (10 Pa.s) and it does not flow at reservoir conditions.

Extra-heavy oil and bitumen which have an API less than 10, i.e. a specific gravity greater than 1, which means that they are heavier than pure water.

The above classification is illustrated in the following figure:

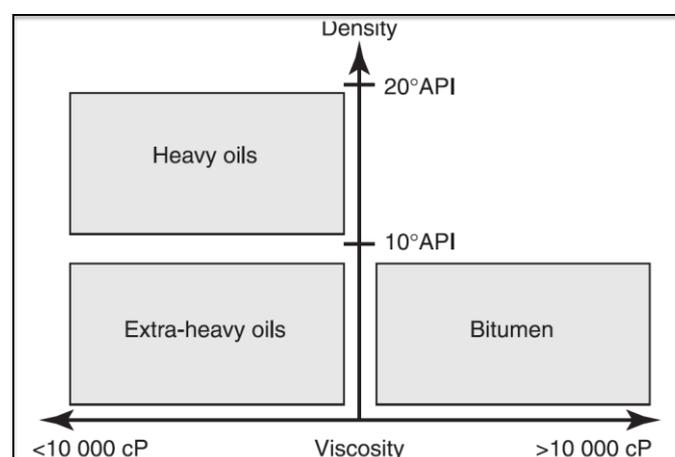


Figure 1: Heavy, Extra-Heavy oil and Bitumen, Saniere, 2004 [7]

There is a positive but very loose correlation between gravity and viscosity and any quantitative transform from API Gravity to Viscosity is a rough approximation because there are no transformation rules.

In summary, the types of the oil are categorized in:

- Conventional crude oil: it is mobile in the reservoir. API gravity > 25°, high permeability reservoir (>100-1000 mD).
- Tight oil: it is immobile in the reservoir. API gravity > 25°, low permeability reservoir (<100mD).
- Medium crude oil: it is mobile in the reservoir. API gravity 20 - 25°, high permeability reservoir (>100-1000 mD).
- Heavy crude oil: it is more viscous than conventional crude oil. API gravity 10 - 20°, it is mobile in the reservoir. Viscosity greater than 100 cP. High permeability reservoir (>100mD).
- Extra-heavy oil: it is mobile in the reservoir. API gravity < 10°, viscosity is commonly lower than 10,000 cP, high permeability reservoir (>100mD).
- Tar sand bitumen: it is immobile in the reservoir, API gravity < 10°, viscosity is commonly greater than 10,000 cP high permeability reservoir (>1000mD).

However, we should not distinguish natural bitumen from extra heavy oil based only on viscosity parameter. Natural bitumen, defined on the basis of specific gravity, is frequently less viscous than 10.000cP [8].

Some of the standard crude oils are given in the following table together with their API Gravities:

	API Gravity
• West Texas Intermediate	40
• Canadian Syn-crude	33
• Arab Light	32
• Alaska NS Crude	29
• Arab Heavy	27
• Alaska Viscous	16 to 24
• Alaska Heavy	8 to 14
• Water	10
• Venezuela (Orinoco)	10
• Canadian Lloydminster	9 to 18
• Canadian Athabasca	6 to 10

Table 1: Standard crude oils and their API Gravities, BP-Heavy-Oil_Milne Point S-Pad Heavy Oil Facility (Provided by the supervisor).

Heavy and Extra-heavy oils in particular are characterized by their high viscosity (10³ to 10⁶ cP) and their low API Gravity (Heavy oil less than 20°API and Extra-heavy oil less than 10°API). This is because of the high concentration of asphaltenes and the relatively low proportion of light weighted components.

HEAVY OIL ORIGIN

The largest accumulations of heavy, extra-heavy oil and bitumen occur in similar geological settings. These are huge, shallow deposits trapped on the flanks of foreland basins. Foreland basins are huge depressions formed by downwarping of the Earth's crust during mountain building that takes place in convergent plate margins. The marine sediments in the basin become the hydrocarbon source rock. Gradually, hydrocarbons migrate upwards into sediments eroded from the newly built mountains and which often lack sealing caprocks. These shallow, cool sediments are creating conditions that are suitable for degradation of hydrocarbons (Figure 2).

Biodegradation is the main cause of the formation of heavy oil. Microorganisms degrade light and medium hydrocarbons, producing methane and enriched heavy hydrocarbons. These microorganisms can grow in relatively cool reservoirs. Dissolved oxygen and nutrients such as nitrate and phosphate must be present and the salinity of the water must be less than 100-150‰. Also, the amount of H₂S in the oil must be very low as it is toxic to aerobic bacteria. Consequently, cool, shallow reservoirs that are flushed by oxygenated, nutrient-rich fresh water can be expected to contain oil that is being actively biodegraded [9].

The effect of biodegradation is to cause oxidation of oil. Oxidation of oil during biodegradation leads to a decrease in saturated hydrocarbon content, to a smaller decrease in aromatic hydrocarbon content and decrease in API gravity. On the other hand, oil oxidation increases oil density, sulphur content, acidity, viscosity and metal content which affect negatively oil production by reducing well flow rates, and refining operations [10], [11], [5]. Through biodegradation, oils also lose a significant fraction of their original mass [12]. As Figure 2 illustrates, the optimal conditions for microbial degradation of hydrocarbons occur in petroleum reservoirs at temperatures less than 80°C [176°F]. As a result, because of the thermal gradient, the process is mainly restricted to shallow reservoirs 4000-6000ft (1220-1830m) deep, or less. However, biodegraded oils are also present in deeper reservoirs. In areas where tectonic activity causes subsidence of reservoirs, biodegraded oils are found preserved below the arbitrary 6000 ft cutoff of bacterial activity. For that reason, cases where biodegraded oils are found in reservoirs 8000-10000ft (2440-3050m) deep do exist. In addition, lower thermal gradients can permit the occurrence of biodegraded oils in bigger depths than expected.

Microbial degradation can also be found at the oil-water interface [10] [13]. The lower part of an oil accumulation will be degraded instead of the upper parts unless more than one oil-water contact is present. Some reservoirs have multiple oil-water contacts and different hydrological regimes. This can lead to a complex interchange of degraded and undegraded oils.

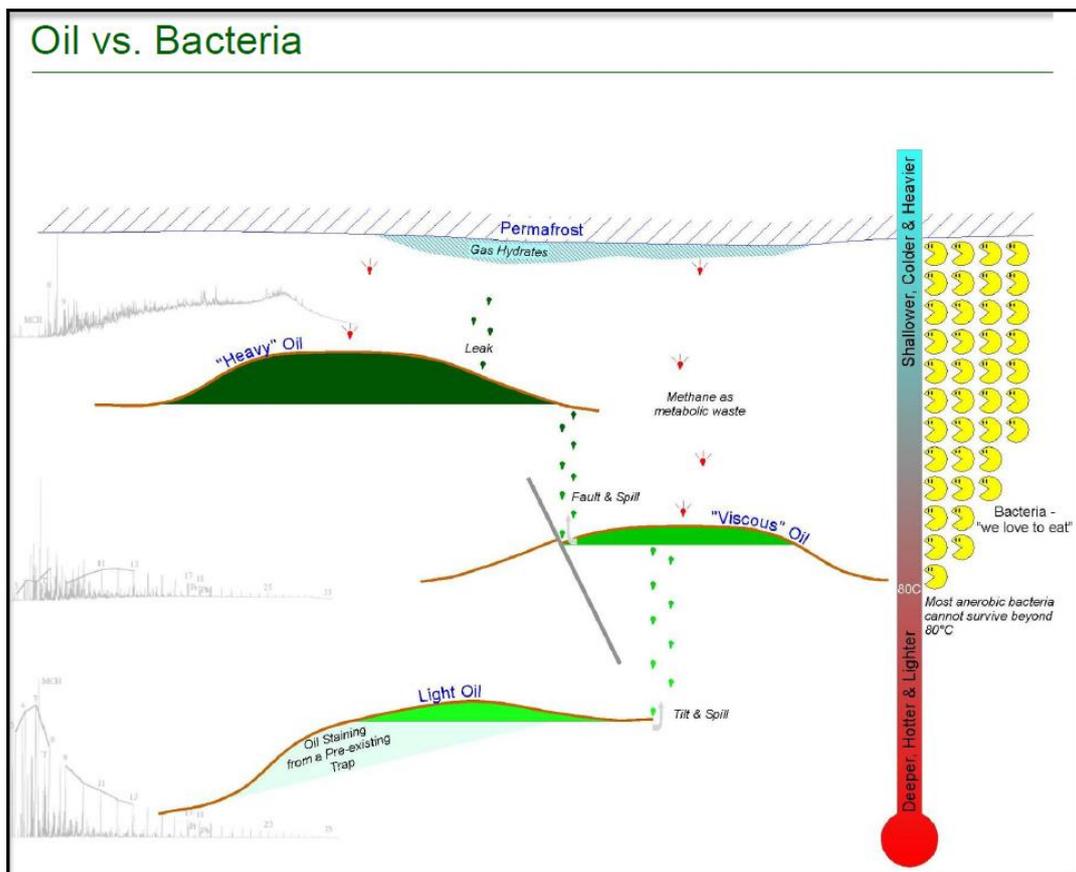


Figure 2: Optimal conditions for microbial degradation of hydrocarbons in temperatures less than 80°C. Biodegradation as the main cause for heavy oil formation, Oil vs Bacteria_BP Heavy Oil Milne Point Presentation (Provided by the supervisor).

Water washing is another process that usually accompanies biodegradation. During water washing the more water-soluble hydrocarbons especially the lower molecular weight aromatics such as benzene and toluene are removed from the oil. Since water is present during biodegradation, water washing and biodegradation can occur simultaneously. In that case it is difficult to determine if biodegradation and/or water washing has changed the oil. Water washing can also happen in deep reservoirs where the lack of oxygen and high temperatures do not allow aerobic bacteria to grow.

Other mechanisms that contribute to the formation of heavy oil are phase fractionation and evaporative loss. In phase fractionation there is light and heavy ends fractionation during migration and maturation in the reservoir. Evaporative loss is the removal of the more volatile components of the oil through vaporization which leads to the alteration of the composition [10], [14].

QUANTITY AND DISTRIBUTION OF HEAVY OILS

It is useful to determine the commonly used terms of oil *reserves* and oil *resources* before moving further. An oil *resource* is the entire quantity of petroleum that exists in the sediments and strata whereas oil *reserves* represent the fraction of petroleum that can be economically recovered. More specifically reserves are those quantities of petroleum which are anticipated to be commercially recovered from known accumulations from a given date forward. Contingent resources are those quantities that are

estimated on a given date to be potentially recoverable from known accumulations but are not currently considered as commercially recoverable [5].

Reserves can be further categorized in proved, probable and possible. Proved reserves include those quantities of oil that have at least 90% confidence of being recoverable under current economic conditions and existing technology. Probable and possible reserves represent known accumulations that have 50% and 10% confidence of recovery respectively based on future economic conditions.

One can easily conclude that the estimated figure of global oil reserves may change if a significant petroleum accumulation, formerly characterized as contingent resource, is characterized as proven reserve. This happens for instance due to the development of new technologies or price changes. An example of such a shift in oil classification that caused a rapid change in reserve's estimates was the reclassification of Venezuela's Orinoco Heavy Oil to proved reserves. In this case the National reserves increased from 99.4 billion barrels in 2007 to 172 billion barrels in 2008 and to 211.17 billion barrels in 2009 as it is shown in the following figure.

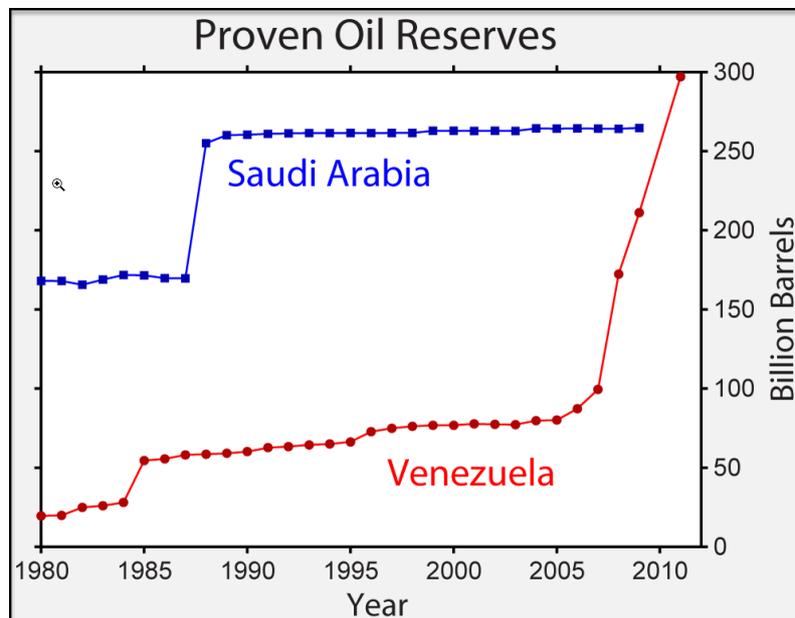


Figure 3: History of Venezuela's claimed oil reserves in comparison to those of Saudi Arabia. (Source Wikipedia/Oil reserves in Venezuela)

Up to 2016 Venezuela's proved oil reserves have reached 300.9 billion barrels according to British Petroleum's statistical review of world energy (2016).

Until recently it has been proved very difficult to acquire data that compare conventional versus unconventional reserves. According to Kovarik [15], there is a tendency to leave out unconventional oil reserves from published data in order to show that the Middle East holds approximately two thirds of the world's oil reserves. This tendency is attributed mainly to political reasons. Only in the year 2000 a US Geological Survey report [16] was published stating that there is much more oil outside the Middle East than inside the region. Based on this report BP published the following chart that shows the dramatic difference in world oil reserves estimation when the unconventional oils are included.

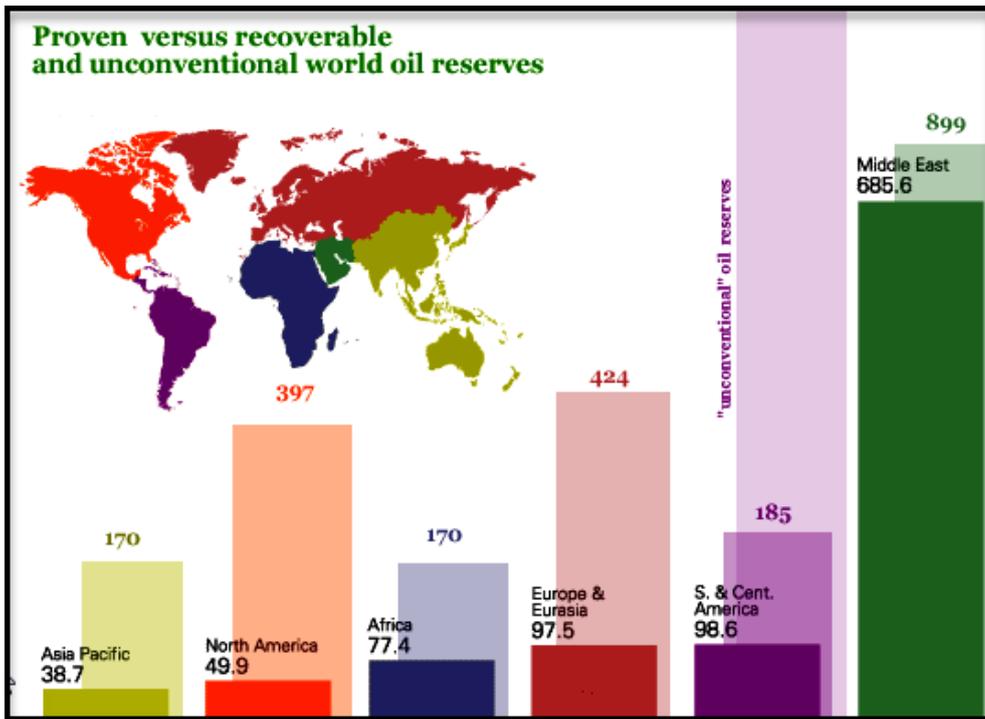


Figure 4: BP world oil reserve chart edited using USGS 2000 World Survey

The chart (Figure 4) combines three bar plots. The most front bar plot represents the proven reserves, the middle bar plot the recoverable reserves and the further back plot - there is only one bar actually - represents the unconventional oil reserves. According to the chart the Middle East holds almost 54% of identified reserves. When it comes to ultimately recoverable reserves the Middle East holds possibly 40% and maybe less than the 30% when the unconventional oil reserves are included. When extra-heavy oil and bitumen are included in the world reserves estimation the pie chart (Figure 5) published in 2006 is closer to what may be actually happening.

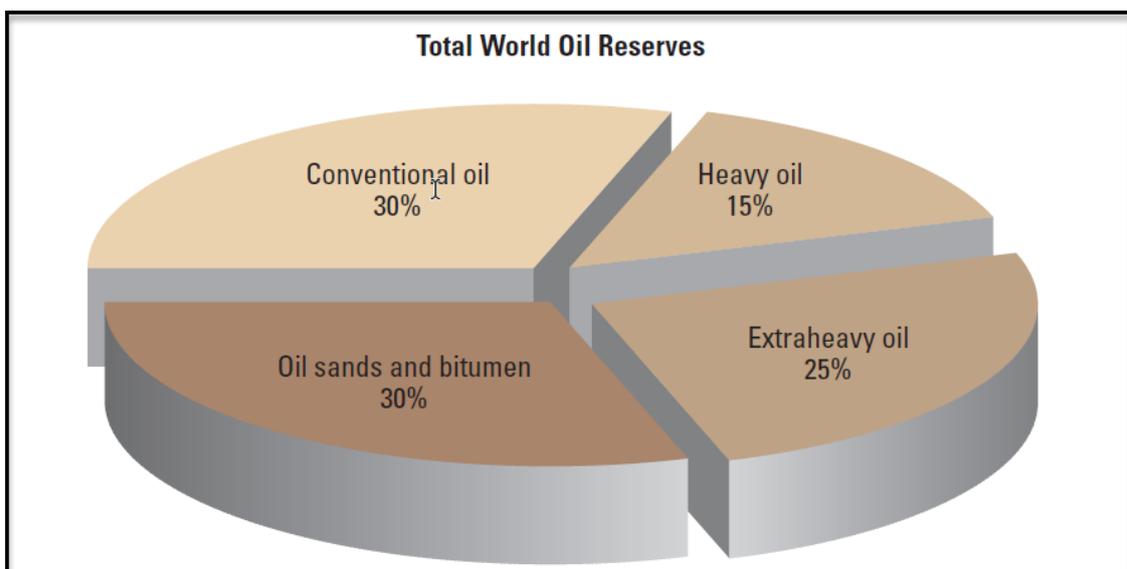


Figure 5: Distribution of Total World Oil reserves by classification, Oil Field Review, Highlighting Heavy Oil 2006

Heavy oil, extra heavy oil and bitumen in figure 5 are expected to represent 70% of the world's total oil reserves. In 2004 Saniere et al. [7] presented data underlining the increasing importance of heavy oil and

bitumen production. As Figure 6 shows heavy oil, extra heavy oil and bitumen reserves sum up to almost the same amount as the conventional reserves although very little from the former reserves has already been produced.

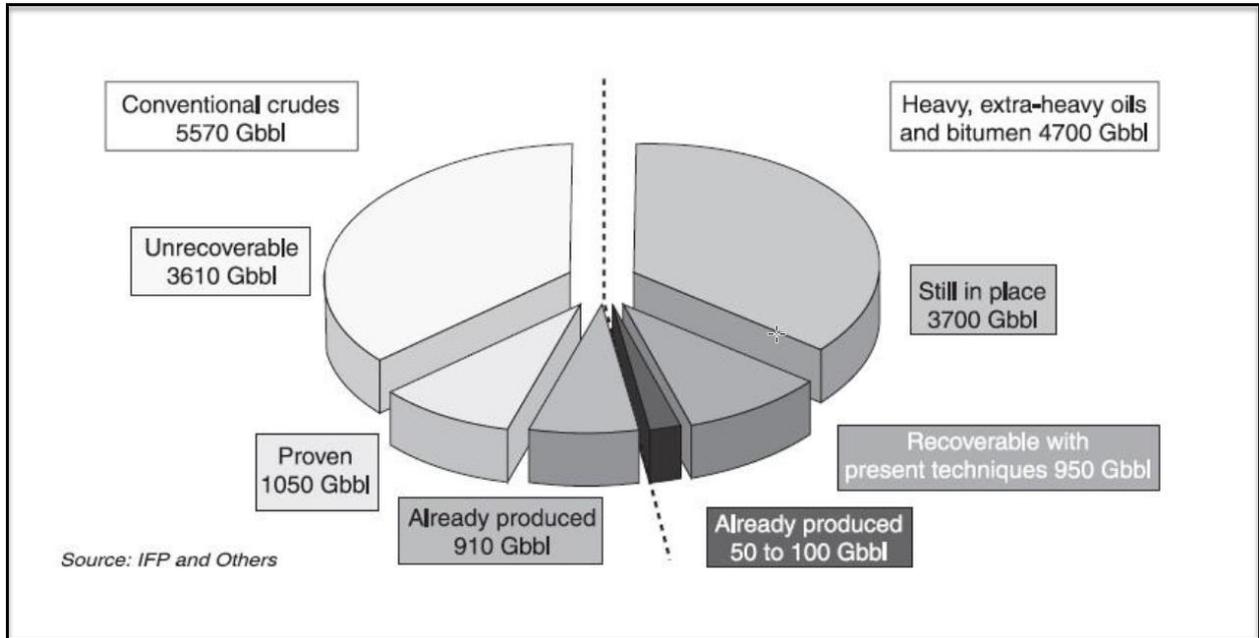


Figure 6: Total world oil resources produced and in place, Saniere et al., 2004

The figures presented by Speight (2016) are similar to the ones of figure 6. The conventional oil reserves are estimated to be approximately 836 billion bbl worldwide, while the estimated recoverable heavy oil is 430 billion bbl and tar sand bitumen 650 billion bbl. These coarse figures show that the estimated volume of recoverable heavy oil and bitumen is at least equal to the remaining conventional oil reserves, which is also shown in the following pie chart (Figure 7):

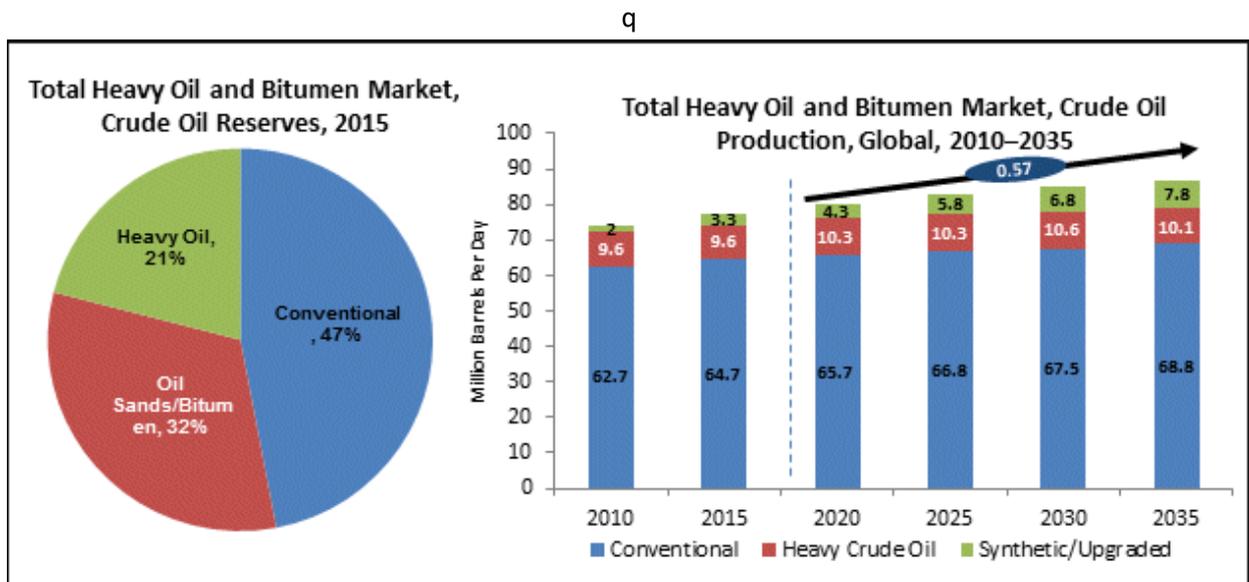


Figure 7: Source World Oil Outlook 2014 - OPEC - BP Statistical Review of World Energy 2015 - USGS [17]

In the following table (Table 2), British Petroleum presents the most recent figures of the global proved reserves by region without defining the fraction attributed in conventional and unconventional oils. Only in the case of the Canadian oil sands and the Venezuela Orinoco belt the quantities are given separately.

Total proved reserves							
	At end 1995	At end 2005	At end 2014	At end 2015			
	Thousand million barrels	Thousand million barrels	Thousand million barrels	Thousand million barrels	Thousand million tonnes	Share of total	R/P ratio
US	29.8	29.9	55.0	55.0	6.6	3.2%	11.9
Canada	48.4	180.0	172.2	172.2	27.8	10.1%	107.6
Mexico	48.8	13.7	10.8	10.8	1.5	0.6%	11.5
Total North America	126.9	223.6	238.0	238.0	35.9	14.0%	33.1
Argentina	2.4	2.2	2.4	2.4	0.3	0.1%	10.2
Brazil	6.2	11.8	16.2	13.0	1.9	0.8%	14.1
Colombia	3.0	1.5	2.4	2.3	0.3	0.1%	6.3
Ecuador	3.4	4.9	8.0	8.0	1.2	0.5%	40.4
Peru	0.8	1.1	1.4	1.4	0.2	0.1%	34.3
Trinidad & Tobago	0.7	0.8	0.8	0.7	0.1	*	18.1
Venezuela	66.3	80.0	300.0	300.9	47.0	17.7%	313.9
Other S. & Cent. America	1.0	1.5	0.5	0.5	0.1	*	9.9
Total S. & Cent. America	83.7	103.6	331.7	329.2	51.0	19.4%	117.0
Azerbaijan	1.2	7.0	7.0	7.0	1.0	0.4%	22.8
Denmark	0.9	1.3	0.6	0.6	0.1	*	9.6
Italy	0.8	0.5	0.6	0.6	0.1	*	14.7
Kazakhstan	5.3	9.0	30.0	30.0	3.9	1.8%	49.3
Norway	10.8	9.7	6.5	8.0	1.0	0.5%	11.3
Romania	1.0	0.5	0.6	0.6	0.1	*	19.5
Russian Federation	113.6	104.4	103.2	102.4	14.0	6.0%	25.5
Turkmenistan	0.5	0.5	0.6	0.6	0.1	*	6.3
United Kingdom	4.5	3.9	2.8	2.8	0.4	0.2%	8.0
Uzbekistan	0.3	0.6	0.6	0.6	0.1	*	25.3
Other Europe & Eurasia	2.2	2.2	2.1	2.1	0.3	0.1%	15.0
Total Europe & Eurasia	141.2	139.5	154.6	155.2	21.0	9.1%	24.4
Iran	93.7	137.5	157.8	157.8	21.7	9.3%	110.3
Iraq	100.0	115.0	143.1	143.1	19.3	8.4%	97.2
Kuwait	96.5	101.5	101.5	101.5	14.0	6.0%	89.8
Oman	5.2	5.6	5.2	5.3	0.7	0.3%	15.3
Qatar	3.7	27.9	25.7	25.7	2.7	1.5%	37.1
Saudi Arabia	261.5	264.2	267.0	266.6	36.6	15.7%	60.8
Syria	2.6	3.0	2.5	2.5	0.3	0.1%	253.7
United Arab Emirates	98.1	97.8	97.8	97.8	13.0	5.8%	68.7
Yemen	2.0	2.9	3.0	3.0	0.4	0.2%	176.5
Other Middle East	0.1	0.1	0.2	0.2	†	*	2.8
Total Middle East	663.3	755.5	803.8	803.5	108.7	47.3%	73.1
Algeria	10.0	12.3	12.2	12.2	1.5	0.7%	21.1
Angola	3.1	9.0	12.7	12.7	1.7	0.7%	19.0
Chad	-	1.5	1.5	1.5	0.2	0.1%	52.4
Republic of Congo	1.3	1.5	1.6	1.6	0.2	0.1%	15.8
Egypt	3.8	3.7	3.7	3.5	0.5	0.2%	13.2
Equatorial Guinea	0.6	1.8	1.1	1.1	0.1	0.1%	10.4
Gabon	1.5	2.1	2.0	2.0	0.3	0.1%	23.5
Libya	29.5	41.5	48.4	48.4	6.3	2.8%	306.8
Nigeria	20.8	36.2	37.1	37.1	5.0	2.2%	43.2
South Sudan	n/a	n/a	3.5	3.5	0.5	0.2%	64.9
Sudan	0.3	0.6	1.5	1.5	0.2	0.1%	39.2
Tunisia	0.4	0.6	0.4	0.4	0.1	*	18.6
Other Africa	0.7	0.5	3.7	3.7	0.5	0.2%	38.3
Total Africa	72.0	111.3	129.3	129.1	17.1	7.6%	42.2
Australia	3.8	3.7	4.0	4.0	0.4	0.2%	28.3
Brunei	1.1	1.1	1.1	1.1	0.1	0.1%	23.8
China	16.4	15.6	18.5	18.5	2.5	1.1%	11.7
India	5.5	5.9	5.7	5.7	0.8	0.3%	18.0
Indonesia	5.0	4.2	3.6	3.6	0.5	0.2%	12.0
Malaysia	5.2	5.3	3.6	3.6	0.5	0.2%	14.2
Thailand	0.3	0.5	0.4	0.4	†	*	2.3
Vietnam	0.8	3.1	4.4	4.4	0.6	0.3%	33.3
Other Asia Pacific	1.1	1.4	1.3	1.3	0.2	0.1%	12.0
Total Asia Pacific	39.1	40.8	42.6	42.6	5.7	2.5%	14.0
Total World	1126.2	1374.4	1700.0	1697.6	239.4	100.0%	50.7
of which: OECD	149.2	244.0	253.9	255.3	38.0	15.0%	29.7
Non-OECD	976.9	1130.4	1446.1	1442.3	201.3	85.0%	58.0
OPEC	786.6	927.8	1211.1	1211.6	169.9	71.4%	86.8
Non-OPEC	339.6	446.6	488.9	486.0	69.4	28.6%	24.9
European Union#	8.3	7.0	5.6	5.6	0.7	0.3%	10.1
CIS	121.5	122.2	141.9	141.1	19.1	8.3%	27.8
Canadian oil sands: Total	41.5	173.6	166.2	166.2	27.0		
of which: Under active development	3.6	10.2	24.4	24.4	4.0		
Venezuela: Orinoco Belt	-	-	221.7	222.3	35.7		

Table 2: Total proved reserves, BP statistical review 2016 [18]

Most of the world's heavy oil, extra heavy oil and natural bitumen reserves are located in Canada, Venezuela, Middle East and Russia. This distribution is depicted in the following figure (Figure 8).

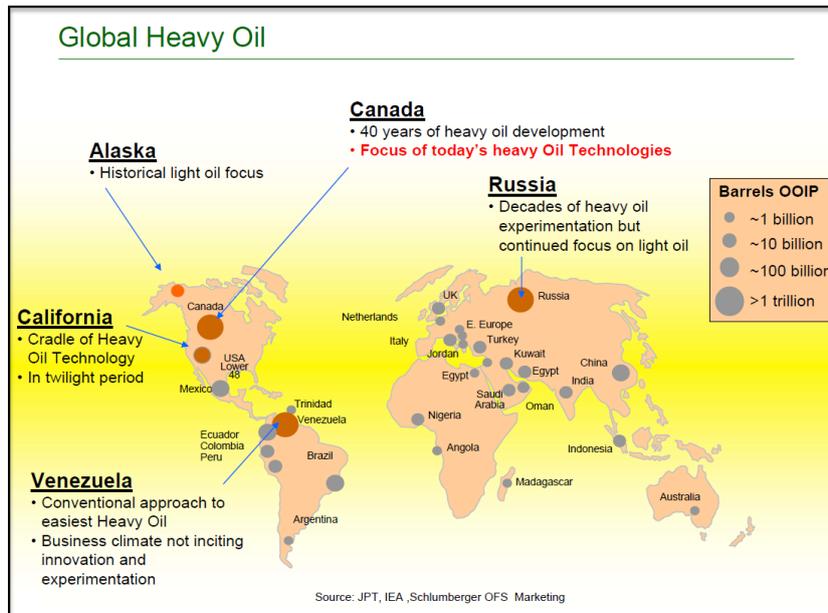


Figure 8: Geographical distribution of heavy oil resources, BP Heavy Oil Milne Point Presentation (Provided by the supervisor)

According to the USGS open file report of 2007 [8], North America, followed by South America has by far the largest accumulations of natural bitumen in the world. Accordingly, most of the world's heavy and extra heavy oil is located in South America and the Middle East, followed by North America. Very large deposits of natural bitumen are located in Eastern Siberia, Russia. Western Siberia mostly contains conventional and medium oil resources and a moderate amount of heavy oil.

In Canada, according to BP's review of the world's energy (2016) [18], oil reserves were estimated at 172 billion barrels. This figure includes gas condensate, natural gas liquids (NGLs) and crude oil reserves but the amount attributed to heavy and extra heavy crude oils is not stated. Bituminous sands reserves are estimated to be 166.2 billion barrels of which 24.4 billion barrels are under active production. Canada's largest producing province is Alberta (Figure 9) that provided 79.2% of Canadian oil production in 2015. Most of Alberta's production comes from the enormous bituminous sands deposits.

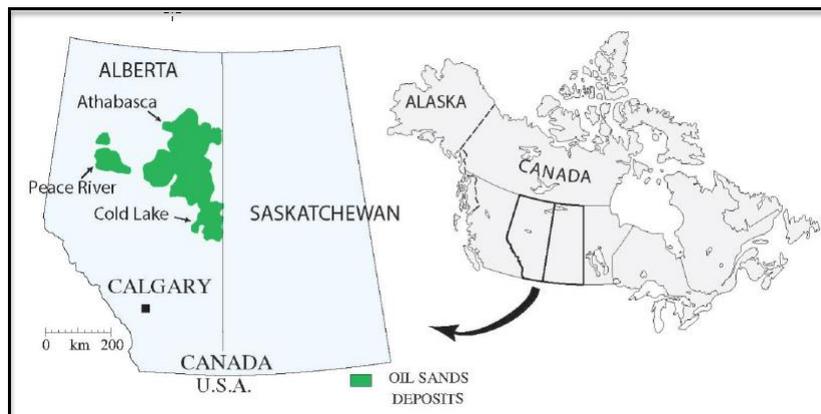


Figure 9: Location of the oil sands deposits of Canada, Survey of Energy Resources World Energy Council Natural Bitumen and Extra-Heavy Oil, 2010 [19]

Alberta's reserves are concentrated in Athabasca, where there is surface and shallow subsurface extraction and Cold Lake and Peace River from which the extraction is from the subsurface [8]. Adjacent

to the oil sands, heavy oil is found in significant accumulations. (source: Wikipedia, petroleum production in Canada). Saskatchewan is the second oil producing province in Canada after Alberta and produces mainly heavy oil. It is noteworthy that none of Saskatchewan's heavy oil deposits is classified as bituminous sands. In this province very large reserves of heavy crude oil are located in Lloydminster area where the oil field crosses the Alberta/Saskatchewan borders.

Venezuela holds the largest amount of oil reserves in the world. Eastern Venezuela basin known as the Orinoco Oil Belt is of particular interest because it includes large accumulation of conventional and medium oil but it also possesses an immense resource of both heavy oil and natural bitumen [8].



Figure 10: Map of the Orinoco Oil Belt Assessment Unit, USGS 2009 [20]

US Geological Survey has conducted an assessment of the Orinoco Oil Belt in 2009 [20]. According to this assessment, the heavy oil in the Orinoco Oil Belt has a range of SG from 4° to 16° API and the viscosity is rather low ranging from below 8000cP to 2000cP. The Orinoco oil in place was estimated with a medium recovery factor of 45% at 1300 billion barrels and with a minimum recovery factor of 15% at 900 billion barrels.

[%, percent; BBO, billion barrels of oil; ft, feet; scf/bbl, standard cubic foot per barrel]

	Minimum	Median	Maximum
Orinoco oil-in-place (BBO)	900	1,300	1,400
Recovery factor (%)	15	45	70
Net oil-saturated sandstone thickness (ft)	1	150	350
Porosity (%)	20	25	38
Water saturation (%)	10	20	25
Formation volume factor	1.05	1.06	1.08
Gas/oil ratio (scf/bbl)	80	110	600

Table 3: Key input data for the assessment of Orinoco Oil Belt Assessment Unit, USGS 2009

The assessment of technically recoverable heavy oil concluded in a mean estimation of 513 billion barrels with a possible range from 380 billion barrels to 652 billion barrels. In the following table (Table 4) there is a regional distribution of discovered heavy oil and natural bitumen in place.

Region ¹	Discovered original oil in place	Prospective additional	Total original oil in place
Heavy oil			
North America.....	650	2	651
South America.....	1099	28	1127
Europe.....	75	0	75
Africa.....	83	0	83
Transcaucasia.....	52	0	52
Middle East.....	971	0	971
Russia.....	182	0	182
South Asia.....	18	0	18
East Asia.....	168	0	168
Southeast Asia and Oceania.....	<u>68</u>	<u>0</u>	<u>68</u>
Total.....	3366	29	3396
Natural bitumen			
North America.....	1671	720	2391
South America.....	2070	190	2260
Europe.....	17	0	17
Africa.....	13	33	46
Transcaucasia.....	430	0	430
Middle East.....	0	0	0
Russia.....	296	51	347
South Asia.....	0	0	0
East Asia.....	10	0	10
Southeast Asia and Oceania.....	<u>4</u>	<u>0</u>	<u>4</u>
Total.....	4512	993	5505

Table 4: Regional distribution of heavy oil and natural bitumen resources in billion barrels, USGS Heavy Oil and Natural Bitumen Resources in Geological Basins of the World, Open File-Report, 2007 [8].

Europe is not a major player in the field of heavy and extra heavy crude oil reserves as it can be seen in table 4. However, interest is drawn in the Mediterranean due to its geological characteristics. Basin types and tectonic activity such as folds, thrusts and nappes make it an area prone to oil exploration.

More specifically, the presence of oil and gas seeps that have a NNW-SSE orientation from Croatia to Greece has attracted the interest of exploration oil companies immediately after the First World War [21]. A significant hydrocarbon province in this area is Albania. Heavy oils and asphalts from Selenica mine were used even from the Roman times. In 1927 the Kukova and Patos oil fields were discovered related to Messinian clastic reservoirs and in 1957 Marinza oil field, the biggest in Albania, was discovered and it is related to Messinian – Tortonian clastic reservoirs [22]. The Patos Marinza oil field is located 10 km east of the city of Fier in south-central Albania. Estimates of the original oil-in-place volumes range from 2.0 billion barrels to 7.5 billion barrels. Patos Marinza has only heavy oil and in the fourth quarter of 2013, it was producing 19,303 bopd. Enhanced oil recovery technologies such as polymer flooding and thermal methods are extensively applied at this field [22]. In table 5 some of the major oilfields in Albania are presented.

Field	Discover year	Reservoir type	Reservoir depth(m)	O/G gravity API	Sulfur content(°o)
Patos	1927	Mess-clastic	Sur. to 1.2 km	Oil(12–24 API)	2.5–6
Marinza	1957	Mess-clastic	1200–1800m	Oil(12–35 API)	4–6
Visoka	1963	Cret/Eoc.Carb.	800–1000 m	Oil(5–16 API)	5–6
Gorisht-Kocul	1965	Cret/Eoc-Carb.	1000–2500 m	Oil(17 API)	6
Ballsh-Hekal	1966	Cret/Eoc-Carb.	1000–3000 m	Oil (12–24 API)	5.7–8.4
Cakran-Molaj	1977	Cret/Eoc-Carb.	3000–4500 m	Oil (14–37 API)	no
Amonica	1980	Cret/Eoc-Carb.	2200–2600 m	N/A	N/A
Delvina	1989	Cret/Eoc-Carb.	2800–3499	Cond., 53 API	0.7

Table 5: Oil fields in Kurveleshi anticline, New ideas on the tectonic of the Kurveleshi anticlinal belt in Albania, and the perspective for exploration in its subthrust, Telo Velaj, 2015 [23]

Most of the oil fields were discovered in the Ionian zone which is a major oil and gas province in Albania with recoverable reserves in Tertiary clastic and carbonate reservoirs [24]. The Ionian Zone crops out in the central southwestern parts of the country and extends south onshore towards the western part of Greece. The Greek Ionian zone, should be active oil or gas province since there are oil seeps distributed in various locations and is a continuation of the proven Albanian fields (Figure 11).

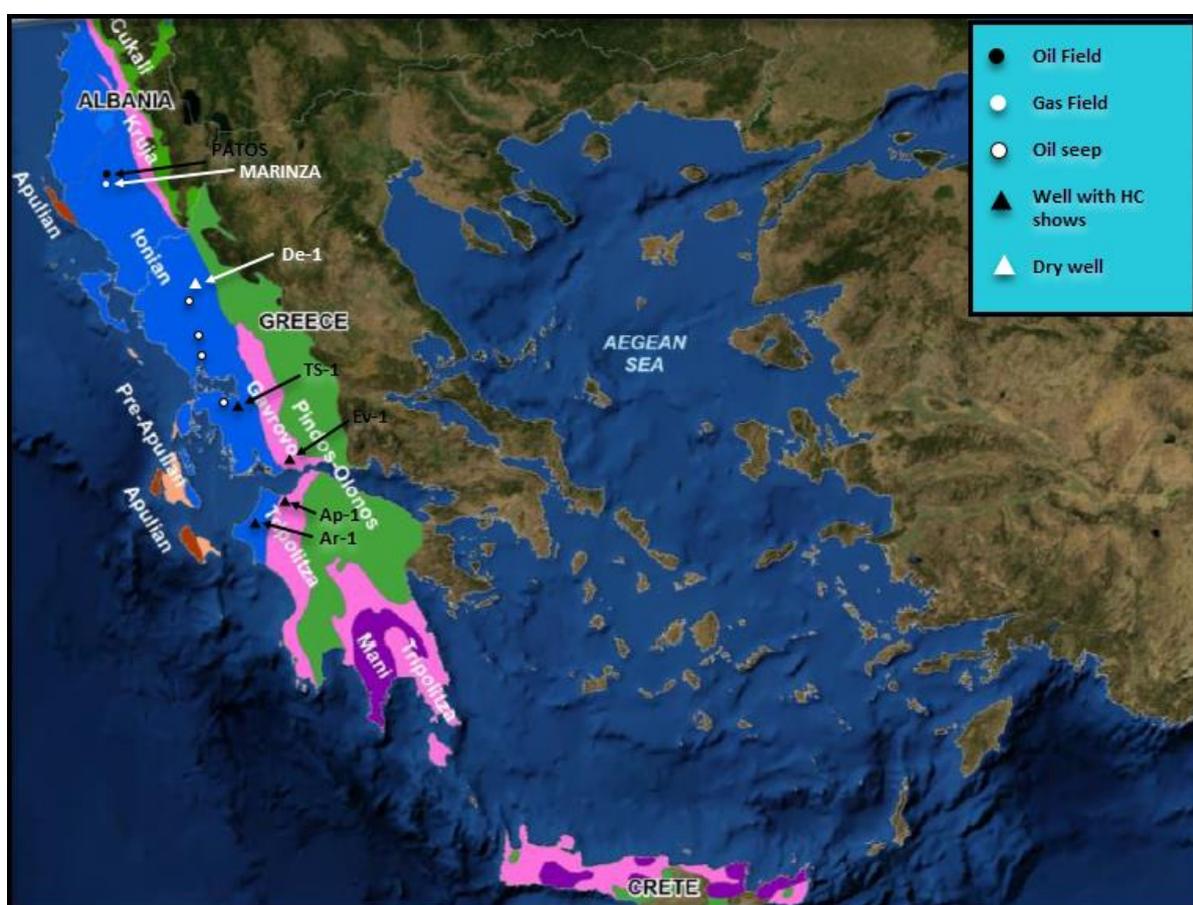


Figure 11: Major tectonic zones, oil and gas fields, major oil seeps and drilled wells in Greece (Ap-1 = Apollo-1, Ar-1 = Artemis-1, De-1 = Demetra-1, Ev-1 = Evinos-1, TS-1 = Trifos South-1), A review of recent hydrocarbon exploration in Greece and its potential, Mavromatidis, A., Kelessidis, V. C., Monopolis D. G., 2004. (Image produced by GEPlan Consulting, modified by C. Argyropoulou)

In 1995, the Greek state oil company Hellenic Petroleum (HP) announced an international tender for six regions in western Greece. In 1997 the government signed four contracts for exploration and exploitation, in Ioannina and northwest Peloponnesos between Enterprise Oil and HP and in Aitolokarnania and Gulf of Patraikos between Triton Ltd. And HP. The wells that appear in figure 11 were drilled during this period

but since the results were discouraging they were plugged and abandoned. However, Western Greece is geologically the southern prolongation of the oil producing Albania. For this reason there is a great possibility for a commercial production to be established in western Greece, an area of active oil seeps, asphalt saturated strata, shows in wells and thick, dark colored bituminous carbonate rocks [24].

The similar geological traits between Albania and western Greece have not led so far to a significant oil discovery in Western Greece. On the other hand, geological analogues between Albania and Italy have led to significant discoveries for both sides as it can be seen in the figures 12 and 13.

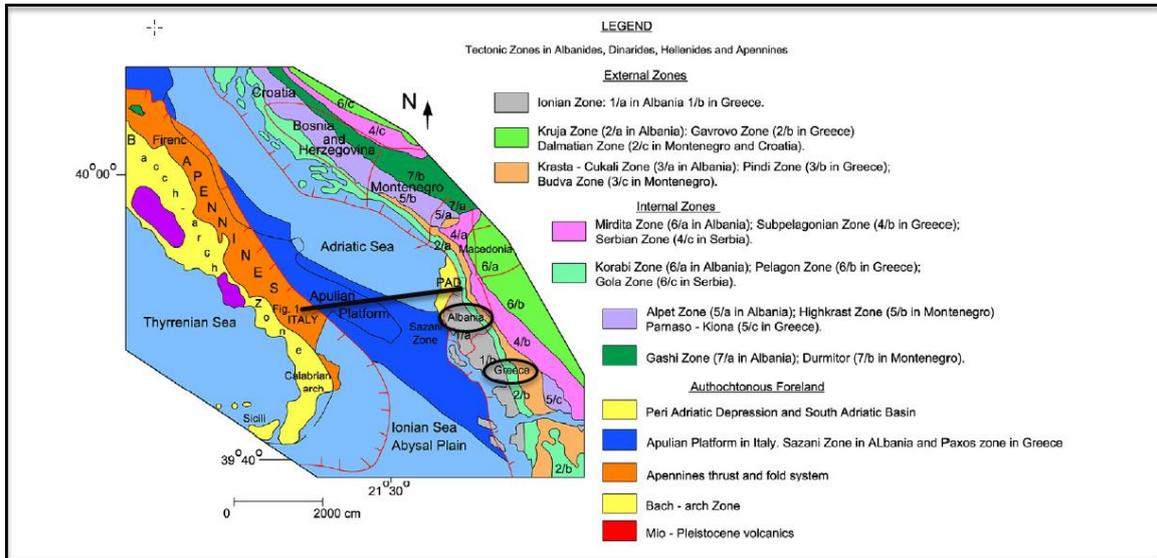


Figure 12: Schematic map of tectonic units in peri Adriatic area between Italy and Albania, Velaj, 2015, modified by C. Argyropoulou

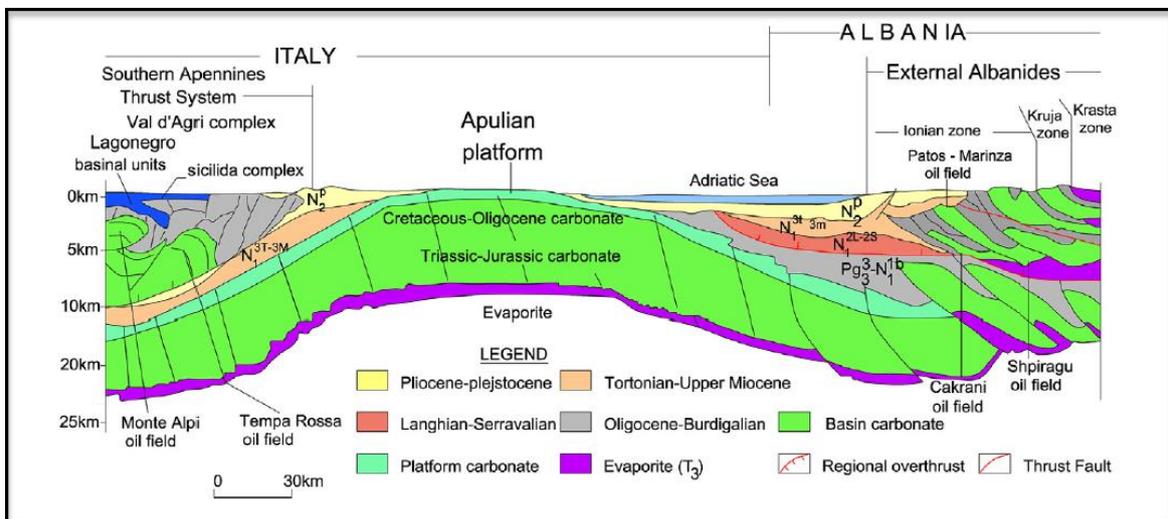


Figure 13: Geological profile in Adriatic area between Albania and Italy. The location of the crosssection is shown in Figure 12, Velaj, 2015

Petromanas Energy (TSXV:PMI) that was operating in Albania believed that Southern Albania region hosts multiple deep oil formations analogous to the Val d’Agri oil field in Italy on the other side of the Adriatic sea (Figure 13). Val d’ Agri has been a producer for a consortium which includes Royal Dutch Shell since 1996 and because of that Petromanas was able to attract Shell as a joint venture partner on Albania’s Blocks 2 – 3 in February, 2012 (Figure 14).

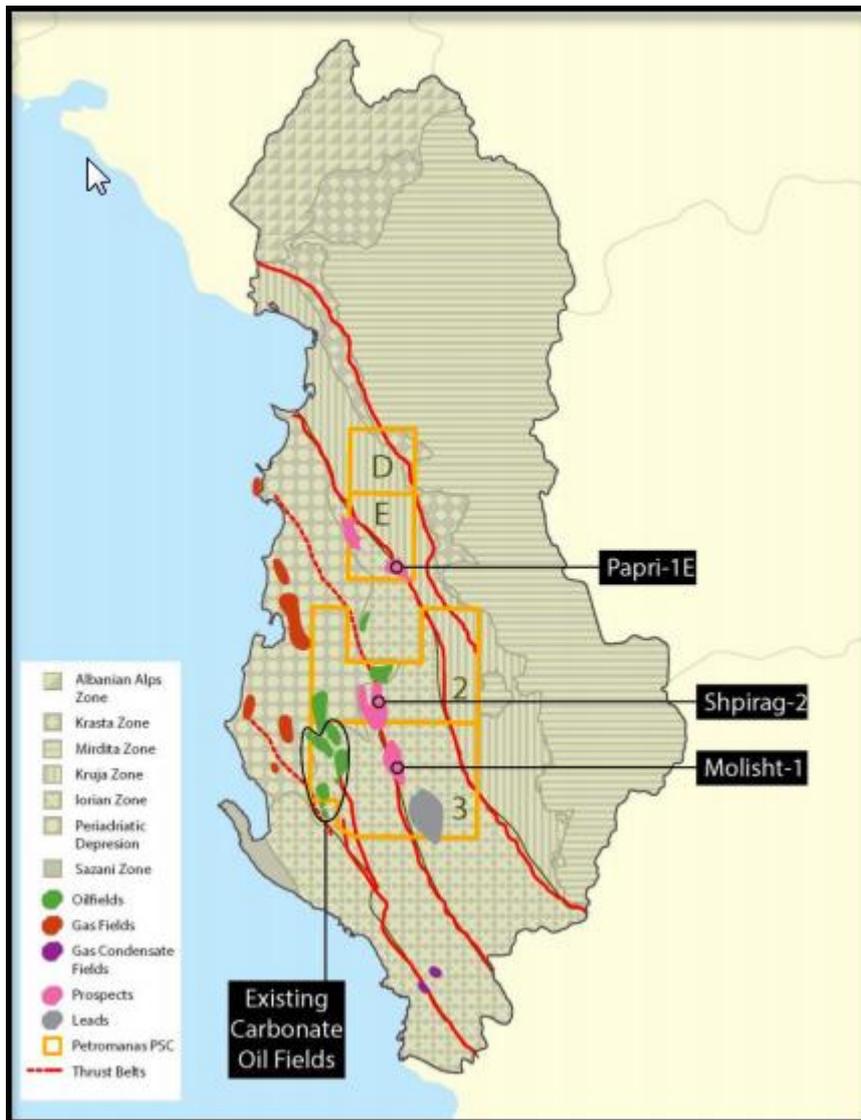


Figure 14: Blocks 2 and 3 in Southern Albania region (source: blog.ceo.ca)

In November 4, 2013, PMI and Shell announced a discovery at Shpirag-2. Producing rates from the well were similar to most wells at Val d' Agry.

In 2016 agrees to acquire Shpiragu block in Southern Albania just above the Greek Albanian borders. This agreement coincides with the announcement of the results of a tender (almost 20 years after the last one) for exploration and exploitation licenses in 3 onshore blocks in Western Greece. These are the Arta-Preveza block, the Etoloakarnania block and the Southwestern Peloponnese block. It is very interesting that the Ioannina region in Northwestern Greece shares similar geological characteristics with the neighboring Albanian regions possessing oil deposits. As a result this raises the oil prospectivity in Greece (source: <http://miningsee.eu>).

An analytical list of the countries reporting deposits of heavy oil and/or natural bitumen, grouped by region is also given in the following table (Table 6).

North America	South America	Europe	Africa	Transcaucasia	Middle East	Russia	South Asia	East Asia	Southeast Asia and Oceania
Canada	Argentina	Albania	Algeria	Azerbaijan	Bahrain	Russia	Bangladesh	China	Australia
Mexico	Barbados	Austria	Angola	Georgia	Iran		India	Japan	Brunei
United States	Bolivia	Belarus	Cameroon	Kazakhstan	Iraq		Pakistan	Taiwan	Indonesia
	Brazil	Bosnia	Chad	Kyrgyzstan	Israel				Malaysia
	Colombia	Bulgaria	Congo (Brazzaville)	Tajikistan	Jordan				Myanmar
	Cuba	Croatia	Democratic Republic of Congo (Kinshasa)	Turkmenistan	Kuwait				Philippines
	Ecuador	Czech Republic	Egypt	Uzbekistan	Neutral Zone				Thailand
	Guatemala	France	Equatorial Guinea		Oman				Tonga
	Peru	Germany	Gabon		Qatar				Vietnam
	Suriname	Greece	Ghana		Saudi Arabia				
	Trinidad & Tobago	Hungary	Libya		Syria				
	Venezuela	Ireland	Madagascar		Turkey				
		Italy	Morocco		Yemen				
		Malta	Nigeria						
		Moldova	Senegal						
		Netherlands	South Africa						
		Norway	Sudan						
		Poland	Tunisia						
		Romania							
		Serbia							
		Slovakia							
		Spain							
		Sweden							
		Switzerland							
		Ukraine							
		United Kingdom							

Table 6: Heavy Oil and Natural Bitumen Resources in Geological Basins of the World, open file report 2007, USGS

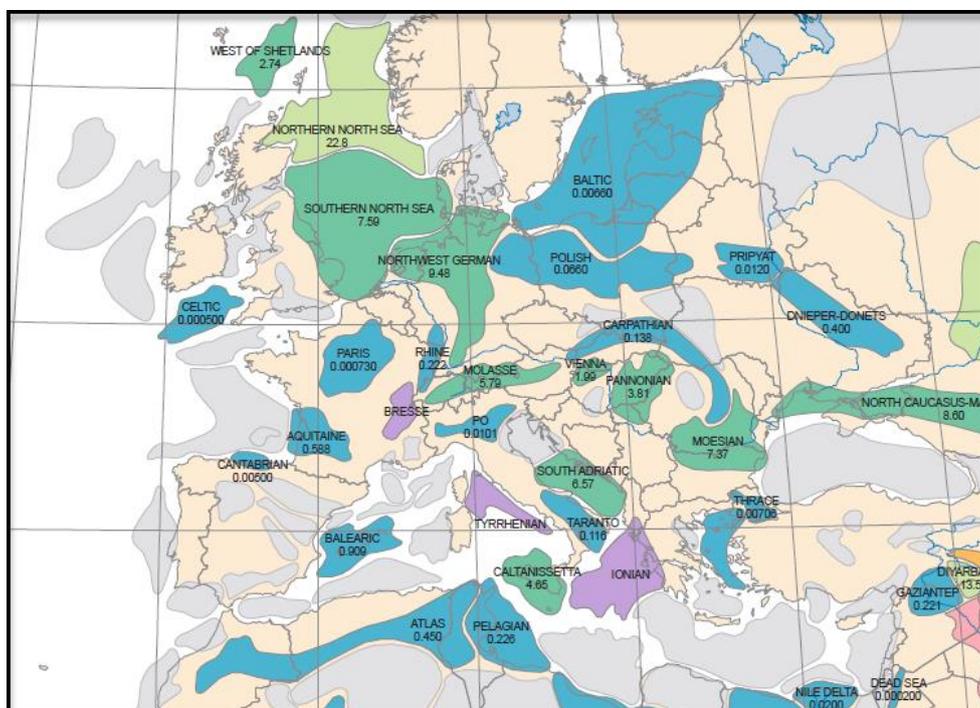


Figure 15: Detail from the map of sedimentary provinces with heavy oil. The European Continent. Meyer, R.F., Attanasi, E.D., and Freeman, P.A., 2007

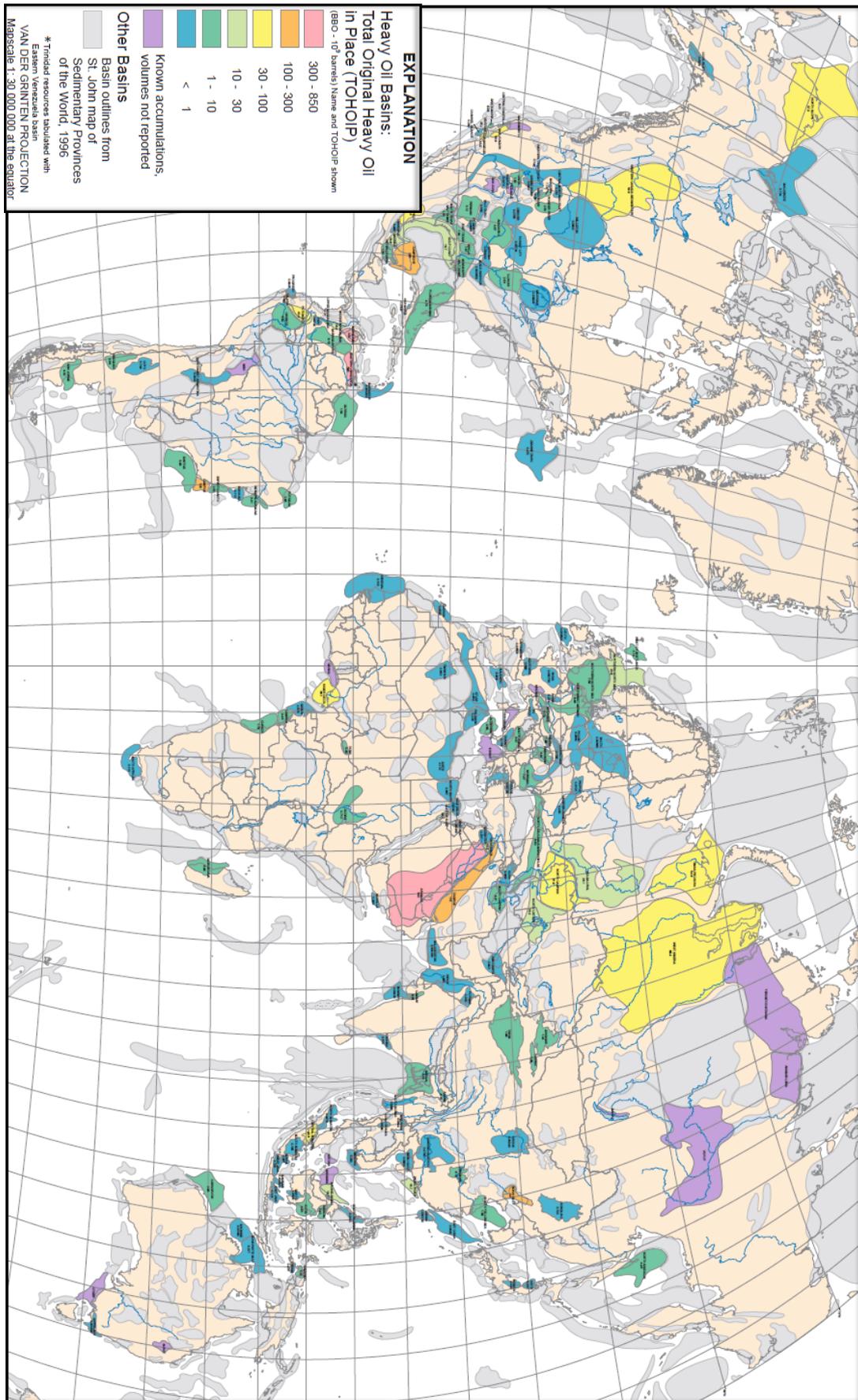


Figure 16: Map of Sedimentary Provinces Reporting Heavy Oil, Meyer, R.F., Attanasi, E.D., and Freeman, P.A., 2007, Heavy oil and natural bitumen resources in geological basins of the world: U.S. Geological Survey Open-File Report 2007-1084

PROCESSING AND FLOW ASSURANCE PROBLEMS OF HEAVY OIL PRODUCTION

INTRODUCTION

Flow assurance and processing challenges related to heavy and extra heavy oils are the result of their high viscosity, low API gravity, and composition. These parameters should be taken into account when surface facilities and operational practices are decided.

Problems encountered during heavy and extra heavy oil production are the following:

- High dynamic viscosities of these oils render their lifting from production wells to the surface very difficult especially in offshore locations.
- High restart pressures after lengthy shutdowns especially in subsea offshore systems
- High tendency to form precipitated solids during depressurization towards the surface such as asphaltenes, waxes, hydrates, naphthenates and inorganic scales.
- Production of fine solids and sand from the formation.
- Multiphase flow and slugging in production pipelines.
- Tendency to form stable emulsions because of their high content in emulsifying agents. These agents act as surfactants that stabilize oil-water interfaces and include asphaltenes, waxes, naphthenates, naphthenic acids and inorganic solids (scales, sands, clays)
- High emulsion viscosities, reaching a peak close to the inversion point, and the risk of the formation of ultra-viscous “chocolate mousse” emulsions at high water cuts.
- Tendency of formation of stable foam in separators and slug catchers that cause liquid to entrain to the gas treatment system (liquid carryover)
- Flashed gas in the oil continuous phase may cause swelling of the oil and gas carryunder
- Very slow oil-water gravitational separation rates linked to the small density differences between oil and water and the high viscosity of the oil phase. Heating up to 120°C or more lowers the fluid viscosities and helps degassing and stabilizing these heavy oils [25].

Consequently, heavy and extra heavy oil processing and flow assurance requires:

- The use of artificial lifting techniques both offshore and onshore otherwise these crudes cannot be produced because of their high viscosity and density
- Efficient handling of the suspended solids often associated with the production and processing of heavy oils
- Handling complex issues concerning flow regimes, emulsion stability, foaming tendencies and transportation.
- High energy supply for process heating and large equipment for gravitational separation due to high retention times.

Artificial lifting is not in the scope of this thesis but is briefly discussed in the next chapter along with enhanced oil recovery. Retention times and heating are discussed in gravitational separation chapter. Solid production, emulsion stability, foaming and flow regimes are dealt with in this chapter.

Asphaltenes, paraffin waxes, gas hydrates, inorganic scales deposition and naphthenate precipitates are issues that should be addressed prior to detailed infrastructure design. Some additional problems are the high stability of heavy oil emulsions as well as the formation of foams. Emulsions and foams are very often the result of high shear rate conditions.

Shear rate is the rate of change of velocity during which one layer of fluid passes over an adjacent layer [26]. This situation is always present during fluid flows but it is especially intense when the fluid passes through restrictions such as safety valves, chokes, pumps, tubing and other process equipment during production [27].

ASPHALTENES

Asphaltenes are the heaviest and the most polar fraction of crude oil. Their concentration is higher in heavy and extra heavy crude oils than in medium and light ones. Asphaltenes can precipitate during field operations with changes in pressure, temperature and composition of the crude oil. They also contribute to the high viscosity and coking tendency of heavy and extra heavy oils [28]

	Type of crude oil			
	Extra light	Light crude	Heavy crude	Extra heavy
API Gravity	> 50	22 - 32	10 - 22	< 10
Hydrocarbons				
Asphaltenes, wt.%	0 - <2	<0,1 - 12	11 - 25	15 - 40
Resins, wt.%	0,05 - 3	3 - 22	14 - 39	--
Oils, wt.%	--	67 - 97	24 - 64	--
Impurities				
S, wt.%	0,02 - 0,2	0,05 - 4,0	0,1 - 5,0	0,8 - 6,0
N, wt.%	0,0 - 0,01	0,02 - 0,5	0,2 - 0,8	0,1 - 1,3
Ni+V, ppm	< 10	10 - 200	50 - 500	200 - 600

Table 7: Main properties (average) of different types of petroleum, Future technology in heavy oil processing, Jorge Ancheyta, Mohan S. [29]

Asphaltenes are defined as the fraction of crude oil which is soluble in aromatic solvents such as toluene, benzene or xylenes and insoluble in light paraffinic solvents like n-pentane or n-heptane. Or, in other words, asphaltenes are the crude oil fraction that precipitates upon the addition of an alkane, usually pentane or heptane, but remains soluble in toluene, benzene or xylenes.

Asphaltenes typically contain carbon and hydrogen at a ratio of approximately 1 to 1.2, oxygen, sulfur, nitrogen and sometimes metals like nickel and vanadium. Their density is between 1.1 to 1.2 g/cm³. They are complex molecules with loosely defined composition. That means that even though their elemental composition is known they do not have an exact molecular weight. Consequently, because of their complex structure they are not defined according to their chemical formula but according to their properties.

The molecular structure of asphaltenes has been a subject of debate for a long time. Starting from the Yen model back in 1961 [30] up to the recently modified Yen model proposed by Mullins in 2007 [6] [31], science on asphaltenes has progressed. The proposed molecular structure includes one single aromatic fused ring system in the molecule with peripheral alkane substituents of varying chain lengths and small amounts of heteroatoms (such as oxygen, nitrogen, sulfur) and heavy metals (such as vanadium and

nickel). By applying time-resolved fluorescence depolarization (TRFD) Henning Groenzin and Oliver C. Mullins [6] estimated the fused ring system containing 4 – 10 rings per asphaltene molecule including a small number of aliphatic chains.

The molecular weight of asphaltenes has been estimated with several techniques. With TRFD [6] asphaltenes were found to have a molecular weight distribution of ~ 750 g/mol (± 200 g/mol). This molecular weight involves asphaltene molecules with an average number of seven fused rings [31].

Asphaltenes have tendency to aggregate making the oil viscous. They are dispersed in crude oils as nanoaggregates with approximately 4 – 10 monomers in each aggregate. Therefore the molecular weight of the nanoaggregates can range between 500 – 7500 g/mol [32]. Asphaltene nanoaggregates form at low asphaltene concentrations in the petroleum fluid. When several molecules are combined in a nanoaggregate, the produced effect of steric hindrance excludes further molecules from incorporating to the same aggregate. At high asphaltene concentrations, the nanoaggregates form clusters but with weaker binding than the aggregates [31]. These clusters have approximately 8 - 12 aggregates each.

Asphaltene aggregates remain in suspension in the crude until they reach a critical concentration (saturation). After the saturation point the excess asphaltenes start to precipitate. This is called the asphaltene precipitation onset. Precipitation of asphaltenes can also be triggered by lowering their saturation point i.e. by increasing the concentration of n - alkanes. As mentioned before, asphaltenes are insoluble in paraffines so increased concentration of n-alkanes can cause asphaltenes to precipitate.

Addition of low molecular weight alkanes (like n-pentane or n-heptane) induces asphaltene precipitation because asphaltene solubility is proportional to the molecular weight of the solvent components. That means that heavy and extra heavy oils which have high asphaltene content are less prone to asphaltene precipitation than the lighter crudes. However, one very usual technique of viscosity reduction in order to facilitate transportation of heavy oils is dilution with lighter hydrocarbons. This apparently changes the composition of the fluid, increases the concentration of low molecular weight alkanes and may cause asphaltene precipitation.

Pressure and temperature changes for a given composition of heavy crudes can also cause asphaltene precipitation. That explains why asphaltenes precipitate in single phase fluids when pressure approaches bubble point pressure during production. This phenomenon is clearly depicted in the asphaltene precipitation envelope which is illustrated in the following figure.

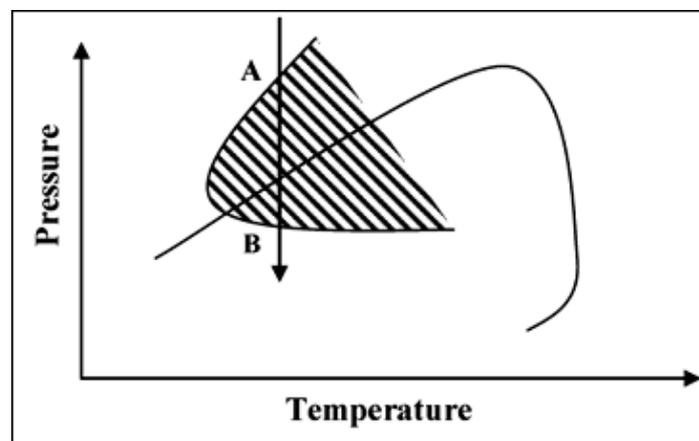


Figure 17: Asphaltene precipitation envelope. The shaded area shows the P, T region in which asphaltene precipitation occurs and includes both single and two phase fluids. source: <http://www.scielo.br>

Asphaltene Stabilization in crude oils

A crucial question to explain asphaltene stability is whether asphaltenes are soluble or they exist only in a colloidal state in crude oil [33]. Two different thermodynamic models were developed for this reason, the solubility model and the colloidal model.

The solubility model describes an asphaltene containing fluid as a mixture of solute (asphaltene) and solvent (bulk oil) in a homogeneous liquid state. Therefore, solubility models assume that asphaltenes are dissolved in crude oil [34]. The asphaltene precipitation can be treated as solid–liquid or liquid–liquid equilibrium, and is reversible. In the solid – liquid equilibrium the precipitated asphaltene is treated as a single component solid phase while the liquid (oil) and gas phases can be modeled with a cubic equation of state. The problem here is that pure asphaltenes do not precipitate alone, they do not have defined melting point and since they have a high aromatic content they form true solution in aromatic solvents. Liquid – liquid equilibrium on the other hand suggests that the high molecular weight of asphaltenes results in insufficient mixing entropy. If the interaction with the rest of the solution becomes unfavorable, for example with the addition of a light aliphatic (non aromatic) fluid, this results in a liquid – liquid phase separation.

The liquid – liquid type was initially modeled with Soave equation of state but a lot of corrections in pressure and temperature should be made to correctly simulate asphaltene solubility at reservoir conditions. More recently the perturbed chain equation of state form of the Statistical Associating Fluid Theory (PC – SAFT) has demonstrated accurate predictions for the bubble point, density and asphaltene precipitation onset of the oils [6], [34].

The colloidal model describes asphaltene phase behavior assuming that the crude oil can be divided into polar and non polar subfractions. In that case, resins stabilize asphaltenes in a micelle formation. The asphaltenes exist in the oil as solid particles, they self–associate into aggregates to form the core and the resins adsorb onto the core based on polar-polar interactions to form a steric shell. This model was based on the observation that increase of resins concentration improves stability of asphaltenes. That means that more and more resins adsorb onto the asphaltene core. Additionally, asphaltenes precipitate when an amount of diluent (aliphatic) is added in the mixture. That is because resins become increasingly soluble and they are separated from the asphaltene micelles. On the other hand, addition of an aromatic solvent (ex, toluene) brings the asphaltenes back in solution.

Even though the colloidal model is widely used to predict asphaltene phase behavior it is also in question. In 2010 Goual [35] demonstrated that the diffusion coefficient of asphaltenes (ie. Function of concentration dependent size of the asphaltene aggregates) is the same in presence and in absence of resins. That latter means that it is not possible for resins to coat asphaltene nanoaggregates and provide them stability. In the same manner, the modified Yen model also questions the colloidal model by not taking into account any resin interaction in the formation of the asphaltene nanoaggregates and nanoaggregate clusters [31].

However, the colloidal model is the one used in practice to predict asphaltene precipitation. The resins help maintain asphaltene stability if they are present in the fluid at a ratio of 3:1, resin:asphaltene. The ratio of saturates to aromatics also plays a role in the asphaltene stability. If the ratio of saturates to aromatics is greater than 1, then destabilization may occur. Analysis of many oils has shown a high asphaltene destabilization probability if the percentage of saturates in the oil exceeds 65% and the ratio of resin to asphaltene is less than 3:1.

This association can be disturbed in various ways. Composition changes such as addition of n-alkanes or mixing different crude oils and/or condition changes such as pressure drops can cause destabilization of asphaltenes. In addition, miscible floods of non-polar gases such as nitrogen and carbon dioxide can also move the stabilizing agents away from the asphaltene molecules [36].

Asphaltene precipitation and deposition problems

Asphaltene precipitation and deposition can occur during production of the reservoir fluid, transportation of the produced fluid and processing the fluid in downstream operations. During heavy oil operations, the dilution of heavy oil with paraffinic solvent or lighter oils for viscosity reduction can cause asphaltene precipitation in pipelines, tubing and surface facilities. Asphaltenes also precipitate during vapor extraction process for heavy oil recovery.

Deposition of asphaltenes in oil wells, pumps, pipelines and production facilities can reduce well productivity, damage pumps, restrict or plug pipelines and foul production facilities [7].



Figure 18: Asphaltenes clogging a pipe. Courtesy of A. Pomerantz, Schlumberger (chapter 2 Petroleum Asphaltenes Lamia Goual University of Wyoming USA) [37].

Precipitated asphaltenes may also accumulate near the wellbore, reservoir rock and clog the porous matrix of the reservoir during drilling and chemical treatment. Besides formation damage, asphaltene deposits can also cause change in reservoir wettability from water wet to oil wet

The current methods used to deal with asphaltene deposition problems include injecting aromatic solvents with or without dispersants to dissolve the deposit, combining solvents and heating, pigging the pipe section of the deposit, or hydraulic fracturing near the wellbore to overcome restricted flow from the damaged formation.

The refining of heavy oils with asphaltene content in downstream operations is also very challenging. This can lead to coking, fouling, and catalyst deactivation during processing or upgrading, as high temperatures and vacuum conditions are required.

OTHER PETROLEUM PRECIPITATES AND DEPOSITS

Petroleum waxes are complex mixtures of n-alkanes, i-alkanes and cycloalkanes with carbon numbers ranging approximately from 18 to 65 [38]. Waxes may form or precipitate as a solid or gel material when

the temperature of the reservoir fluid falls below a certain temperature, known as the wax appearance temperature (WAT). Wax precipitation does not necessarily lead to deposition because wax crystals tend to disperse in the fluid instead of forming deposits on a surface. However, if the number of wax crystals is large enough or other nucleating materials such as asphaltenes, formation fines, clay or corrosion products are present, wax crystals may agglomerate into larger particles that may separate from the fluid and form solid deposits.

In field operations, paraffin waxes may coprecipitate with asphaltenes and residual oil components leading to precipitated solids of varying colors and texture. Waxes from condensates or wet gases tend to be more clear and pure compared to those of heavier crudes. In general, small amounts of aromatic hydrocarbons precipitate with paraffin waxes and a usual remedial technique is to apply heat in order to melt the waxy solid deposits [6].

The most common environment for wax formation and deposition is the transportation of highly paraffinic fluid from off-shore reservoirs by using subsea pipelines. In that case, the low temperatures of the subsea environment progressively cool down the transported reservoir fluid. If the temperature of the fluid within the pipeline falls below the wax appearance temperature, wax precipitates and deposits as a solid layer inside the pipeline (Figure 19).

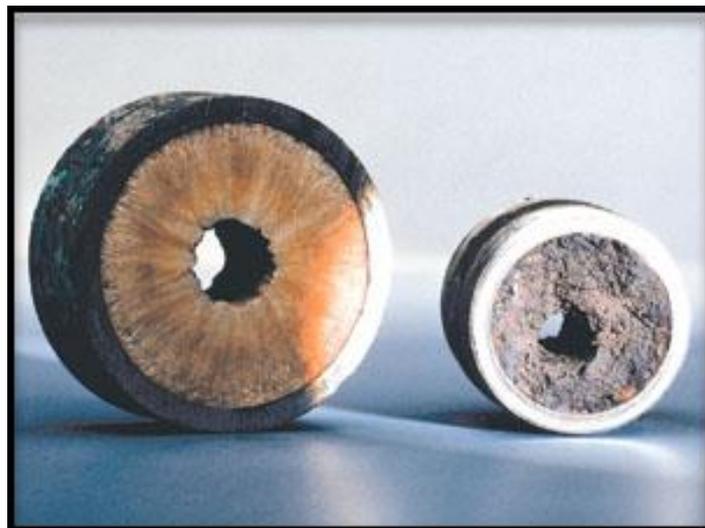


Figure 19: A cross-section of paraffin and scale blocked flowlines. (Photo courtesy Halliburton), source: www.offshore-mag.com/articles/print/volume-57/issue-3/departments/technology-focus/deepwater-challenges-paraffins.html [39].

Precipitated wax, as mentioned before, may not form deposits but be transported in suspended form. That leads to an increased viscosity of the transported fluid and associated high pressure drops. Wax precipitation can also lead to the formation of wax-gel during shut-down period of a well or pipeline. That creates difficulties during start-up because greater pumping power is required and it can even lead to the abandonment of the well or pipeline [28].

Gas hydrates are formed in high pressure and low temperature conditions when hydrogen bonded water molecules, referred to as “hosts” encage “guest” gas molecules from the hydrocarbon fluid. When the lattices formed by the water molecules are filled with these compounds a semi-stable solid may form, similar to ice. The most common hydrocarbon components that form gas hydrates are the ones with C6-carbon numbers. There are three different hydrate structures namely structure I, II and H. Their difference

is the size of the guest molecules. Structure I hydrates can contain the smallest molecules such as methane, ethane and CO₂ whereas structures II and H contain mixtures of both small and large guest molecules. However structure H encages relatively larger molecules than structure II. Structure II hydrates contain molecules smaller than pentane [40].

The main enclathrated hydrocarbon component for the most hydrates in nature is methane, while the respectives for non-hydrocarbon are, CO₂ and H₂S.

The formation of gas hydrates, plugging, and deposition is the most common flow assurance problem in the upstream petroleum operations.

Naphthenic acids are present in many crude oils which exhibit high Total Acid Number and for this reason they are referred to as High TAN crudes [41]. Their chemical formula is R-CO₂H and they derive from the oxidation of the naphtha fraction. Naphthenic acids are responsible for naphthenic acid corrosion phenomena in refineries and pipelines. In addition, naphthenic acids are the main toxicants in wastewaters from oil refineries and oil sands extraction [42]. They are toxic to animals and to aquatic creatures and they remain in the environment for a long time due to their weak biodegradation tendency.

Naphthenates are salts that derive from interaction of naphthenic acids with ions such as sodium and calcium contained in produced formation water with high pH. During production, oil moves from the reservoir to the production tubing and then to the separation equipment. When depressurization takes place, carbon dioxide evaporates from the solution to the gas phase and the pH of the brine increases. Consequently, naphthenic acids lose one hydrogen ion as the following chemical reaction demonstrates $\text{RCO}_2\text{H} \rightarrow \text{RCO}_2^- + \text{H}^+$ [43]. This way, naphthenates either act as natural surfactants and stabilize emulsions (Figure 20) or form solid deposits with the calcium cations contained in brine (Figure 21).

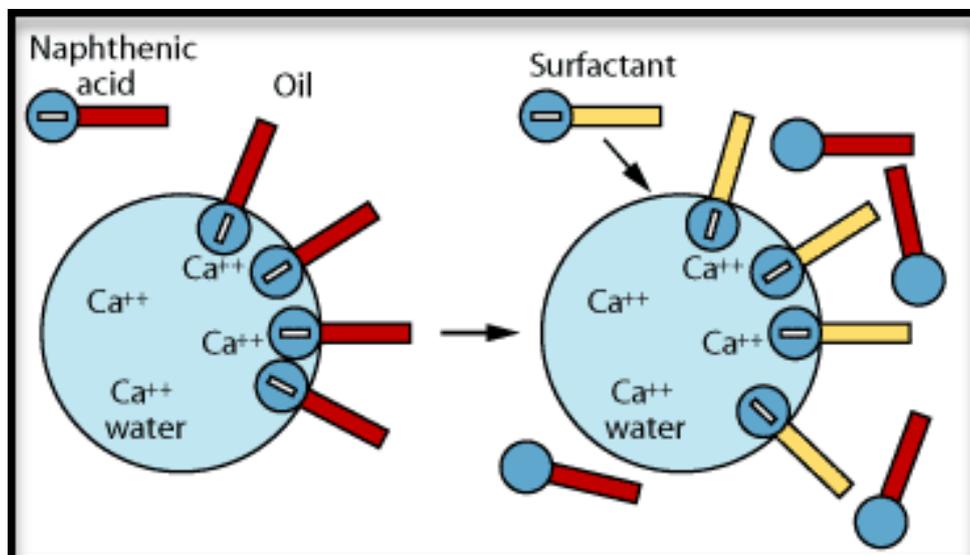


Figure 20: At the left of the picture: Naphthenic acids acting as surfactants. At the right side of the picture: Surfactants known as naphthenate inhibitors use the mechanism of interfacial crowding to inhibit naphthenate salt formation, (source: www.oilfieldwiki.com/wiki/Naphthenates#Production_problems_caused_by_naphthenates)



Figure 21: Naphthenate deposit found in an electrical desalter in Angola (source: www.oilfieldwiki.com)

Naphthenates are sticky solids that get hard upon contact with air, they are insoluble in oil and water and have a density between oil and water. Naphthenate deposits are mostly collected in oil/water separators and desalters but also in tubing and pipelines and they may cause severe damage or even lead to unplanned shut-downs. Very common way to avoid naphthenate deposits is by using surfactants that will occupy the oil-water interface more actively than naphthenic acids (Figure 20) or add an acid in order to decrease pH up to 6.

INORGANIC SCALES

Another kind of solid deposit that can be formed during production procedures is inorganic scales. The most common types of inorganic scales are calcite (CaCO_3) and barite (BaSO_4) and also CaSO_4 and SrSO_4 . The type of inorganic scales that may be formed depends mostly on the composition of the connate water in the reservoir rock. The inorganic ions present in the formation water can precipitate if they exceed their solubility limits. The most common anions that contribute to scale formation are Na^+ , K^+ , Ca^{2+} , Mg^{2+} , Ba^{2+} , Sr^{2+} , Cl^- , HCO_3^- and SO_4^{2-} [28].

These deposits may be found in the reservoir, in the production tubing or at surface facilities. Scale deposition causes formation damage, blockage in perforations or gravel packs, down-hole control or logging devices damage, restricted flow or blockage of flow lines, safety valves, chokes, pump failure and also enhances corrosion problems in production tubing and transportation pipelines.

DISPERSED SYSTEMS IN HEAVY AND EXTRA HEAVY OILS

When viscous oil is produced along with water and gas, they create a multiphase mixture. This mixture typically consists of a dispersed system of either water droplets or gas bubbles in oil continuous phase.

When water droplets are dispersed in oil we deal with a water-in-oil emulsion and when gas bubbles are dispersed in oil then we deal with a foamy oil phenomenon. These dispersed systems are of great importance because they have different flow properties from the single crude oil and affect well production and surface facilities.

Oil emulsions

During production the reservoir fluid undergoes various shear stresses along with the presence of drilling and production additives. This may result in the formation of emulsions [44].

An emulsion is a system of two immiscible liquids in which the one liquid is dispersed in the form of droplets throughout the other liquid that comprises the continuous phase. In a water-in-oil emulsion (w/o), water is the dispersed phase and oil is the continuous phase while in an oil-in-water emulsion (o/w) the opposite happens.

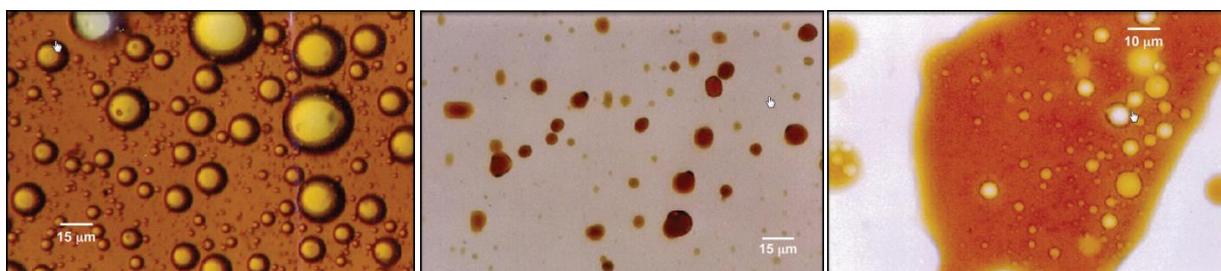


Figure 22: From the left to the right: water-in-oil, oil-in-water and water-in-oil-in-water emulsions. (source: http://petrowiki.org/Oil_emulsions)

When the fluid passes through restrictions such as pumps, tubing or safety valves, chokes and any kind of equipment that may cause high shear rates, the result can be a very stable emulsion. Besides the mechanical energy during shearing, emulsion stability depends on the presence of chemicals that may act as emulsifying agents, such as friction reducers, drilling additives, rate of penetration (ROP) enhancers, corrosion inhibitors etc. Also, heavy oils contain high concentrations of polar compounds with high molecular weights, such as asphaltenes, resins, naphthenates, mercaptans and carboxylates that can act as natural surfactants.

Also, the presence of suspended or precipitated solids in the formation water of the produced fluid can influence asphaltene precipitation, generating a greater mass of solids and stable emulsions. The surface-active resin-asphaltene aggregates adsorb wax and other solids found in the interface of water droplets, stabilizing the emulsion and possibly leading to production plugging (Figure 23).

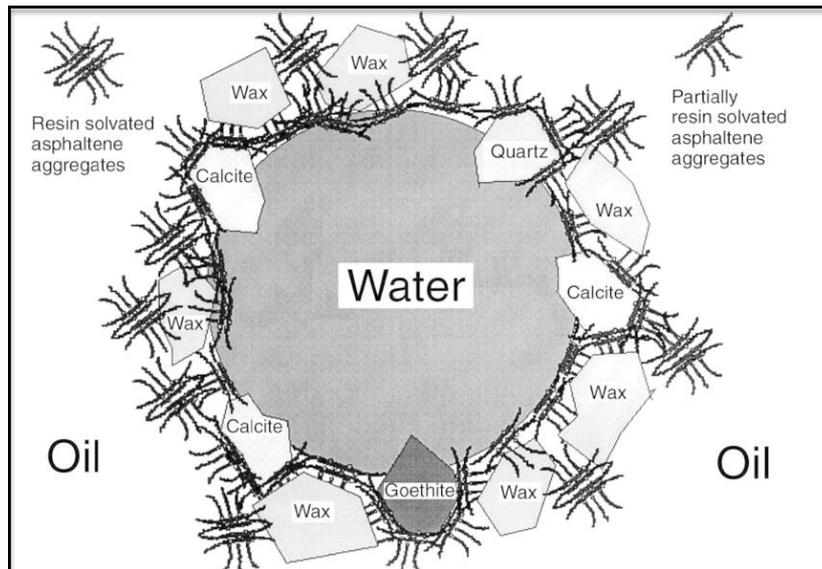


Figure 23: A visualization of the stabilizing effect of asphaltene and other solids on water droplets. (Sarbar, M.A. and Wingrove, M.D. 1997. *Physical and Chemical Characterization of Saudi Arabian Crude Oil Emulsions*. Presented at the SPE Annual Technical Conference and Exhibition, San Antonio, Texas, USA, 5–8 October. SPE-38817-MS) [45].

In addition to natural emulsifiers that are inherent in heavy crudes, enhanced oil recovery methods (see next chapter) that are generally used in heavy oil production contribute to the creation of stable emulsions [27]. These recovery methods include in-situ combustion, steam injection, steam assisted gravity drainage and cyclic steam stimulation. In such cases the emulsions are created through the addition of thermal energy that disperses water in crude oil. Especially CSS and SAGD that are used in the case of heavy oil production create formation damage and this is where stable emulsion are created. Emulsion problems can also emerge from acid stimulation when acid is pumped into a heavy crude oil reservoir. If the necessary additives are missing, mechanical energy combined with the low pH of the stimulation fluid can lead to the formation of viscous emulsions. In any case, mechanical energy is necessary for the creation of small water droplets dispersed in the crude oil and the presence of surfactants - either from the oil itself or from added fluids – is necessary for stabilizing the dispersed phase. Stable emulsions create flow assurance issues, high concentrations of oil in the disposal water, decline of productivity in wells due to increased apparent viscosity and high costs for pipelines, gravitational separation and processing facilities [46].

A good indication of emulsion stability is usually the time required for phase separation. Emulsions with droplets of less than 10 microns in diameter are considered very stable. In addition, high concentration of droplets with small average size leads to an increased apparent viscosity of the oil and decreased flow rates.

The most significant processes in order to breakdown an emulsion is flocculation and coalescence. Flocculation is based on the aggregation of the droplets without any change in the size of each droplet. Coalescence is when two droplets join each other and create a larger droplet by breaking the liquid film around them. This procedure is irreversible (Figure 24).

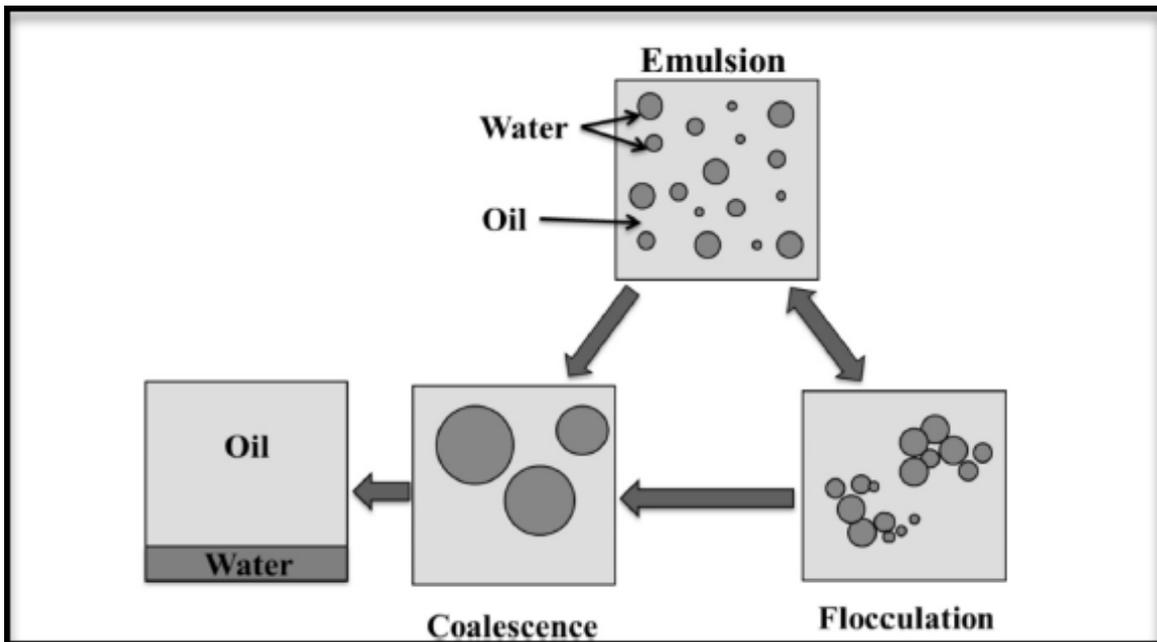


Figure 24: Schematic of flocculation and coalescence process (Treatment of water in heavy crude oil emulsions with Innovative microemulsion fluids 2014) [27].

One important issue concerning emulsions is the emulsion inversion point. There is a critical water concentration that varies according to the oil composition of each fluid above which water-in-oil emulsions (oil continuous phase) invert to oil-in-water emulsions (water continuous phase). The inversion from water-in-oil to oil-in-water emulsion is characterized by a great difference in apparent viscosity as it is shown in the following figure.

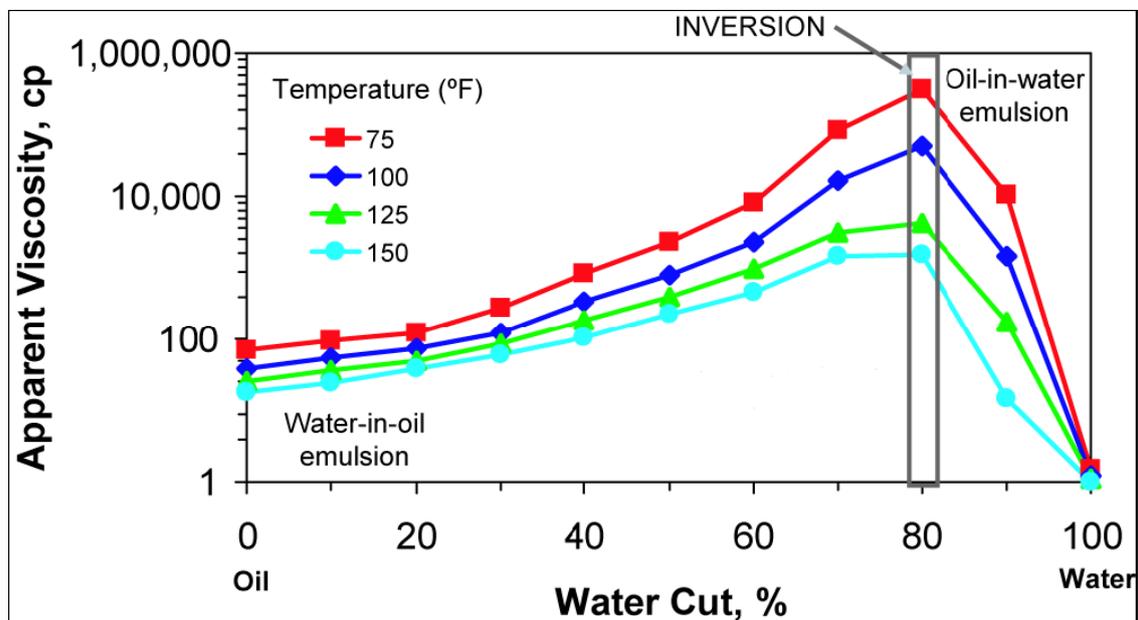


Figure 25: Emulsion inversion points of a tight emulsion in four different temperatures, source: http://petrowiki.org/Oil_emulsions

The exact determination of the inversion point is strongly associated with the collection of representative emulsion samples. In order for the emulsion sample to represent the liquid from which it

is taken, no further emulsification should be allowed after the sample is extracted. For this reason, many ways have been developed to avoid any pressure drop during sampling [47].

The emulsion inversion point is crucial in heavy oils between 21 and 13°API [25]. The reason for this is because in some heavy oils there is a significant rise in the emulsion viscosity as they reach the inversion point concentration in water, followed by a sudden decrease after the inversion point. It is therefore advisable the produced fluid to reach primary separation equipment at water cuts lower than the ones of their inversion point by at least 10% or in water cuts higher than the inversion point. This way, production at the peak of the emulsion viscosity can be avoided.

Several treatment methods exist to break down emulsions: Thermal, by providing heat, mechanical, by increasing residence time in gravitational separation facilities, electrical by applying electrostatic forces to enhance coalescence or chemical by adding demulsifiers.

Foamy oil

Depressurization during production releases the dissolved gas from the oil. There are cases where the released gas does not form a continuous gas phase but creates dispersed bubbles in the crude oil. In the case of light oils these bubbles coalesce very quickly into larger gas bubbles, but in the case of heavy oils the bubbles remain dispersed in the crude and give to the produced oil a foamy form that looks like chocolate mousse [48]. This can be described as oil continuous phase which contains dispersed gas bubbles. It is produced from heavy oil reservoirs under solution gas drive. The term “solution gas drive” means that the driving force that provides the energy for the production and transportation of the reservoir fluid comes from the gas that is dissolved in the oil [49]. Because of the solution gas drive, foamy oil has different behavior than conventional heavy oil. It has low effective viscosity and the expansion of the gas bubbles resulting from the pressure decrease of the reservoir provides the energy to push the oil from the reservoir to the production well [50].

The foamy oil phenomenon has a positive impact during primary cold heavy oil production with sand (CHOPS) of the extra heavy oils in Venezuela Orinoco Belt. It can explain the surprisingly high recovery factors - up to 10% - that are not normally expected in oils with API gravity between 7 and 9.8.

Many theories have been developed in order to explain the connection of the foamy oil phenomenon with the increased production rates [51]. However it is a fact that the apparent viscosity of the foamy crudes depends on the shape of the dispersed gas bubbles and this shape is highly dependent upon the shear conditions. In low shear rates, the bubbles remain spherical and that seems to increase the apparent viscosity of the foamy oil. On the other hand, in high shear rates the bubbles are deformed and elongated which leads to a decrease of apparent viscosity in the direction of the flow. What should be noted is that not all heavy crude oils have a foaming tendency and the phenomenon itself is hard to simulate.

The foamy oil phenomenon has negative impact as well. The entrapped gas into the oil films, which creates the foam, is not easy to be released and separated from the oil. This reduces the capacity of oil/gas gravitational separators because long retention times are needed for a given quantity of foamy oil to be separated [47]. An additional problem of foaming oils is the increased possibility of liquid carry over (liquid escaping with the gas phase from the separation vessel) or gas carry under (gas escaping with the liquid phase from the separation vessel) [52]. Methods that are commonly used for breaking foamy oils are agitation, heating, chemicals and centrifugal forces.

FLOW REGIMS

Slug flow

There are several types of two phase flows both in horizontal and vertical tubing (Figures 26, 27). Slug flow is of special interest because it may create problems in separation facilities and this is the reason why the volume of potential liquid slugs must be established prior to the separator design.

With the term liquid slugs in two phase flow (liquid and gas) we define a series of liquid volumes that are separated by large gas pockets [53]. When the liquid is accumulated in such a quantity that totally blocks the gas flow, then the gas pushes the liquid along the pipe like a slug [52]. The volume of a slug depends on flow rates, flow properties, the size of the pipeline (length and diameter) and the change in elevation.

In vertical pipes, the large gas volumes are axially symmetrical in the shape of a bullet. The resulting fluid switches from high gas to high liquid content.

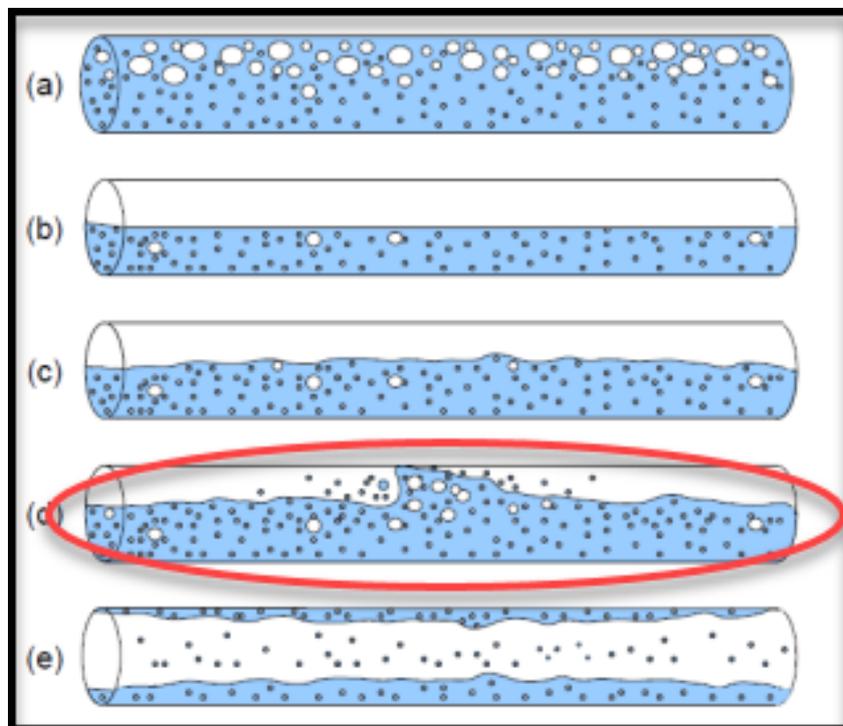


Figure 26: two phase flow regimes in horizontal pipeline: (a) bubble flow, (b) stratified flow, (c) wavy flow, (d) slug flow, (e) annular flow. In the red circle slug flow in horizontal pipe is represented (source: doc.modelica.org/om/Modelica.Fluid.Dissipation.PressureLoss.StraightPipe.dp_twoPhaseOverall_DP.html, image modified by C. Argyropoulou)

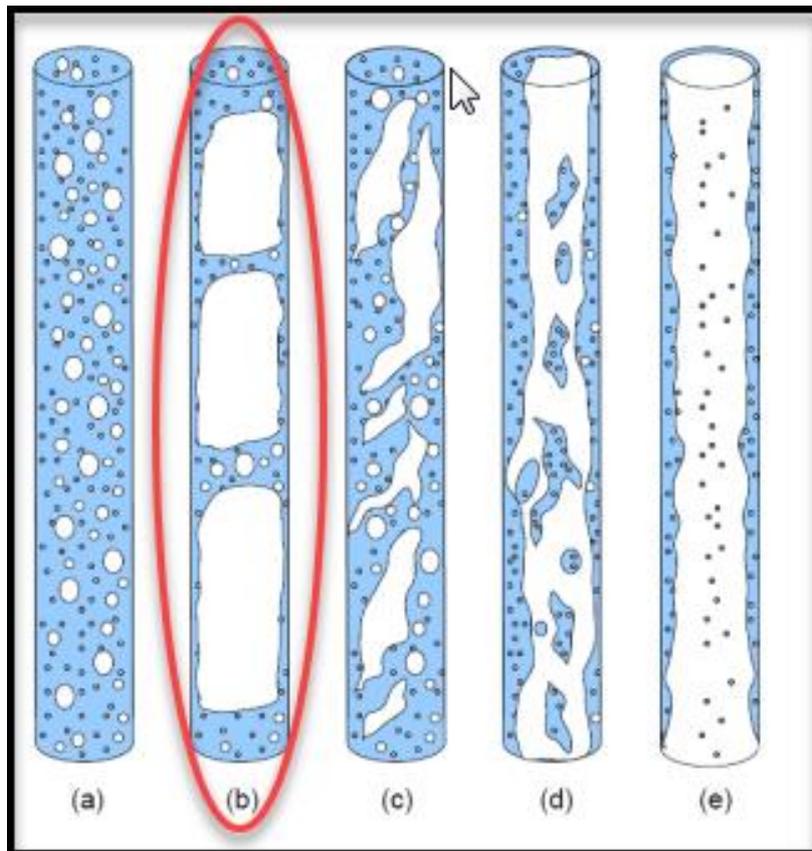


Figure 27: two phase flow regimes in horizontal pipeline: (a) bubble flow, (b) plug slug flow, (c) foam flow, (d) annular streak flow, (e) annular flow. In the red circle slug flow in vertical pipe is represented (source: doc.modelica.org/om/Modelica.Fluid.Dissipation.PressureLoss.StraightPipe.dp_twoPhaseOverall_DP.html, image modified by C. Argyropoulou)

Liquid slugs can cause severe damage in process facilities, flooding of separators, increased corrosion and high back pressures when gas is progressively accumulated behind the liquid slug [54].

When the volume of the anticipated liquid slugs is very big, large volume pipes are installed upstream of the separator. These pipes have such a geometry that can work properly even when they are empty of liquid but they also have enough space to fill with liquid when the slug enters the system [52].

EXTRACTION OF HEAVY AND EXTRA HEAVY OILS

The oil mobility in the reservoir is a very important parameter which describes how easily the oil moves in the porous media. The mobility in porous media is defined by Darcy equation.

$$u_x = \frac{k}{\mu} \cdot \frac{dP}{dx} \quad (1)$$

In equation (1), u_x is the Darcy velocity of the fluid over the x axis, k is the permeability, μ is the viscosity, P is the pressure and x is the distance over the axis. For a single phase fluid k represents the absolute permeability. For multiphase flow, k represents the effective permeability. The phase fluid mobility is given as

$$\lambda = \frac{k}{\mu} \quad (2)$$

The mobility ratio, M , is an important parameter in displacement processes and is defined as:

$$M = \frac{\lambda_2}{\lambda_1} = \frac{k_{rw}\mu_o}{k_{ro}\mu_w} \quad (3)$$

where λ_1 , λ_2 refer to the viscosities of the drained and imbibition fluid respectively. K_{rw} is the relative permeability to water, K_{ro} is the relative permeability to oil, μ_w is the water viscosity and μ_o is the oil viscosity if the drained and imbibition fluids are oil and water respectively [5]. Values higher than unity represent an unfavorable mobility ratio since they imply that the imbibition fluid bypasses the drained fluid. On the other hand values lower than unity represent a favorable mobility ratio since they imply displacement of the drained from the imbibition fluid. [55] [1].

Primary recovery of petroleum depends on the natural energy coming from the reservoir. This natural reservoir drive comes from mechanisms that include water displacing oil towards the well, expansion of the natural gas at the top of the reservoir, expansion of the gas initially dissolved in the crude oil, and gravity drainage resulting from the movement of oil within the reservoir towards production wells. Primary recovery can also be assisted in the well tubing by downhole pumps or gas lift to bring the oil up to the surface. This is called artificial lifting [56].

Secondary recovery is applied when the natural reservoir drive diminishes. Reservoir pressure is not sufficient to bring the oil to the surface, so secondary recovery methods supply extra energy into the reservoir in the form of pressure by water injection and natural gas reinjection [57].

Water Injection is basically an artificial water drive. The energy needed in order to drive the oil towards the production wells is provided by the means of water injection from the surface as it is shown in figure 28. Water injection elevates the decreased reservoir pressure keeping production rates stable for a long time. It is very important the injected water to be chemically compatible with the reservoir fluids so that it does not cause formation damage. For this reason the produced water is commonly used for water injection.

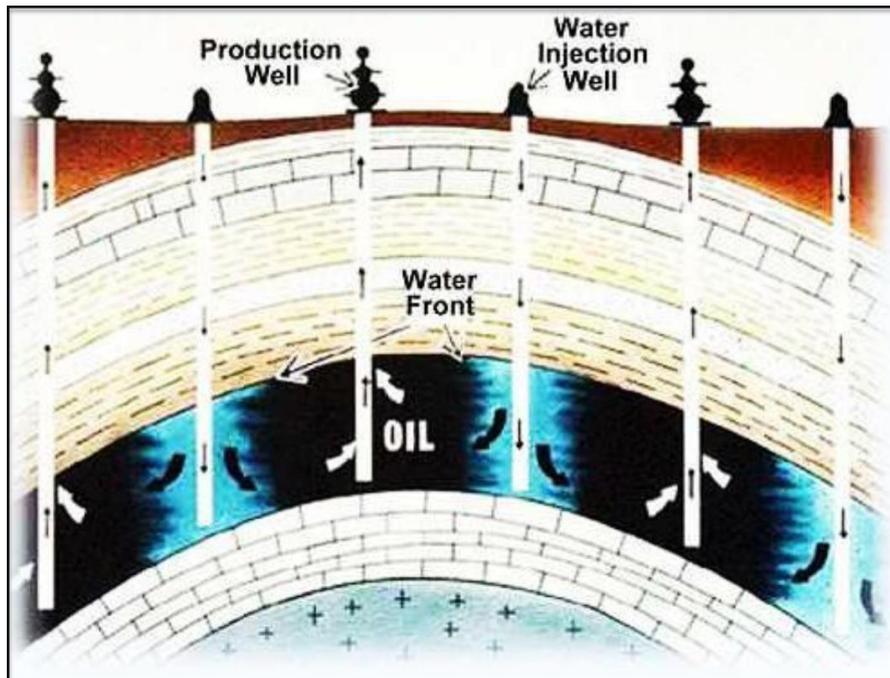


Figure 28: water injection process. Source, newenergyandfuel.com

Gas re-injection serves the purpose to induce the flow of crude oil. This is achieved both by dissolving the gas molecules in the oil so that it lowers its viscosity and by acting as a displacement fluid, similar to the water injection. Once the natural gas is retrieved in the surface it is reused in the injection well. Besides natural gas, carbon dioxide and inert gases can also be used in gas injection processes.

Cold production is a primary recovery method that uses operating techniques and specialized pumping equipment to produce heavy oil without applying heat to the reservoir. In most cases this encourages the associated production of the unconsolidated reservoir sand (UCSS), which is directly connected to significantly higher efficiency of heavy-oil production.

Cold production without sand is applied in heavy oils of relatively low viscosity that flow in reservoir conditions and are produced from boreholes [58]. Economic production rates are achieved by using long horizontal wells completed with slotted liners to exploit large drainage areas. The oil is lifted to the surface by the use of electrical submersible pumps and progressing cavity pumps. In Canada, economic success in oils less viscous than approximately 1500 cp is common. This technology has also found major application in the Faja del Orinoco in Venezuela. In this case, the 12 °API heavy crude has a viscosity of approximately 1500cP with the existing solution gas that can flow at reservoir temperatures [59]. The role of solution gas in the case of heavy oils is very important and it differentiates substantially from the respective of conventional oils: The gas does not generate a continuous phase. The formulated bubbles flow with the fluid and do not coalesce but they expand generating an internal gas drive (foamy flow) that facilitates production. In order to reduce viscosity even further, diluents such as naphtha are injected. Horizontal wells up to 5000 ft long allow the heavy oil to be produced at economic rates at sufficiently low drawdown pressures in order to prevent extensive sand production. Well geometries include wells with several horizontal branches (multilateral wells). The horizontal legs are placed precisely in the target sands using logging-while-drilling (LWD) and measurement-while-drilling (MWD) equipment, enabling a more precise placement of the wells. In some locations, sand production from the unconsolidated formation is minimized by using slotted liners and other sand-control methods. A low drawdown pressure in a long multilateral can also reduce the need for significant sand control. [60], [61].

The main advantage of this method is the considerably lower cost compared to Cold Heavy Oil Production with Sand (presented right after) since it avoids expenses associated with sand handling and disposal. Furthermore, its cost is also lower compared to thermal techniques.

The disadvantages of this extraction method are that it is directly dependent on reservoir geometries and that the recovery factors are usually poor ranging between 5 and 6%. Higher recovery factors, that still do not exceed 10%, can be achieved with optimally designed horizontal wells. In any case, a very significant portion of the original oil-in-place remains in the reservoir.

Cold heavy oil production with sand (CHOPS) is a technique applied both to tertiary recovery of conventional oil as well as to primary production of oil sands and oil shales. In unconsolidated sandstones this technique takes advantage of the fact that the sand influx can enhance oil recovery. For that reason, sand exclusion devices such as screens, liners, gravel packs are not used and sand influx deliberately starts from the beginning of production and continues during the productive life of the well. The sand is produced along with oil, water and gas and is separated on the surface.

This extraction method is characterized by sharp pressure gradients that increase fluid velocity and consequently sand production. Observed sand cuts can reach up to 40% of sand by volume in the beginning of production followed by a drop to 0.5 – 5% after a few months period [62] [63].

Cold Heavy Oil Production with Sand increases productivity for a number of reasons. First of all, the removal of sand grains from the reservoir increases the permeability near the well bore (Figure 29) creating flow paths of high permeability known as wormholes [63]. Additionally, “foamy flow”, which was analyzed in the preceding section, locally destabilizes the sand and improves production. Continuous sand production prevents heavy hydrocarbon clogs, drilling mud and drilling fines from plugging the zone near the wellbore. Lastly, as sand is removed the rock matrix is destabilized and overburden weight acts as an additional natural drive by squeezing sand and oil towards the wellbore.

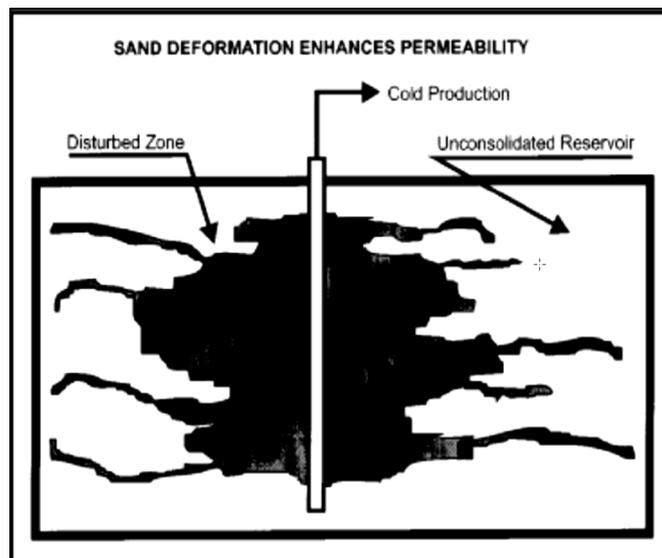


Figure 29: Schematic of sand production with wormholes, Chugh, S., Baker, R., Telesford, A., and Zhang, E., 2000, *Mainstream options for heavy oil: Part I cold production*, [63].

Cold Heavy Oil Production with Sand provides production rates from 10 to 30 times higher than the ones achieved during cold production without sand. The recovery factor is also higher and can reach approximately 12 to 20% of oil in place [58], [64].

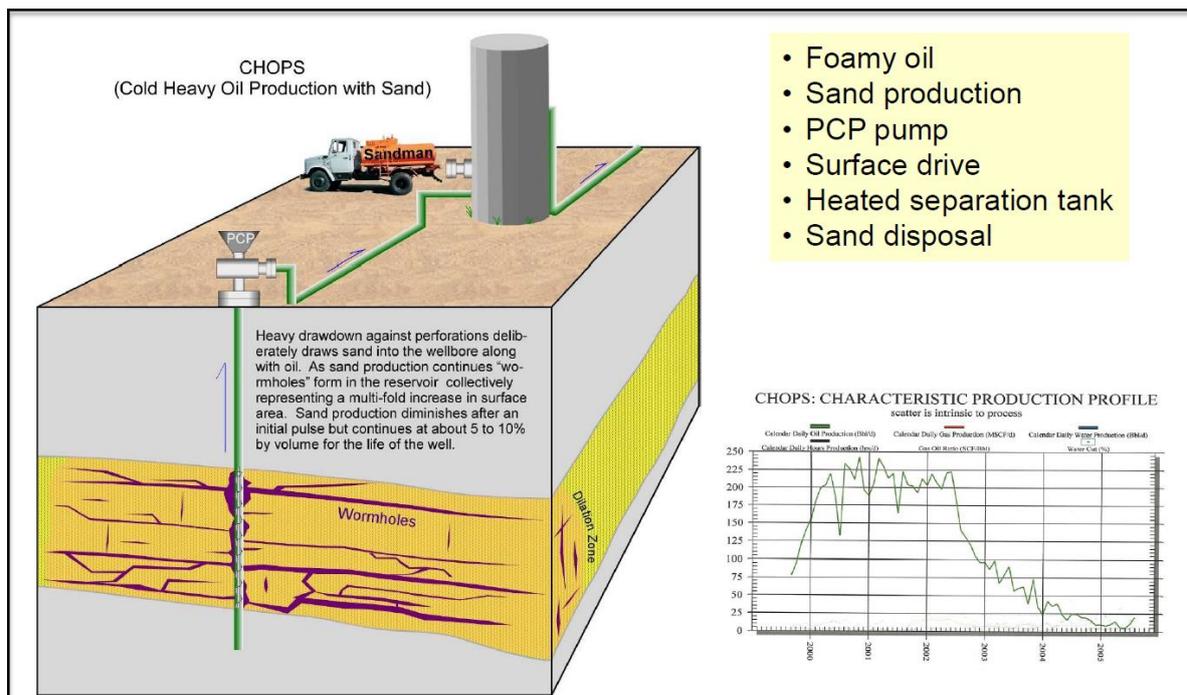


Figure 30: CHOPS method, source BP Alaska Heavy Oil Production from the Ugnu Fluvial-Deltaic Reservoir Josef Chmielowski, Search and Discovery Article #80289, 2013 [65] (Provided by the supervisor).

Disposal of the sand which is contaminated with oil is a serious disadvantage of this method. Waste management expenses can be as high as 15 to 35% of the operating costs [66]. Because of the large volume of sand being produced, pipelines cannot be used for transportation due to the risk of plugging. Instead of that, produced fluids along with sand are stored temporarily in lease tanks that are generally heated to induce separation of sand from the fluids. Trucks are more commonly used to move oil, sand and water to centralized fluid treating and sand disposal facilities. Additional costs related to CHOPS method are caused by short equipment life due to erosion, production interruptions due to well plugging and frequent cleaning of wells and equipment [67]. Furthermore, the stable emulsions that are generated during production and tank cleaning contain water, oil from 20 to 50% and minerals and are very difficult to treat and dispose.

CHOPS is implemented for exploiting 1 – 7 m thick subsurface oil sands containing oil with viscosity higher than 500cP. So far vertical or slightly inclined wells up to 45° and rotary progressive cavity pumps are the standard extraction method in the Canadian heavy oil belt along the Alberta-Saskatchewan borders [68].

Pressure Pulse Technology (PPT) is a relatively new injection technology that was discovered and developed into a production enhancement method in the period 1995 – 2001 [59]. It can be applied in order to enhance recovery rates and reduce clogging in wells and fractured media [69].

Generally when an injected fluid is introduced in the reservoir aiming to displace the hydrocarbons towards the production well, it flows through preferential paths of high permeabilities. Furthermore, the mobility ratio of the displacing and displaced phases often results in the channeling of the displacing phase through the displaced phase which is called viscous fingering. The result is that the injected fluid affects only a portion of the reservoir because the displacement front is not uniform. Consequently fluid injection is partly effective and the sweep efficiency and dispersion are limited.

Pressure Pulse Technology is a high amplitude dynamic excitation of the porous medium which is mechanically accomplished by the injection of a low-frequency pulse of fluid into the medium [70]. This pressure pulse creates different waves one of which is a slow strain wave called porosity dilation wave or fluid displacement wave, similar to a tidal wave. Porosity dilation wave causes dilation and contraction in a pore scale by moving in the porous medium with a velocity from 5 to 150 m/sec [71].

The pressure pulse is repetitive and is created through sudden displacement of the injected liquid in the well bore. The equipment necessary for the PPT is placed close to the perforations and a piston is moved up and down by the service rig creating a mechanical impulse. The repeated flow into and out of the wellbore generates mixing in a pore scale, unblocks the pore throats, disperses the injected fluid, suppresses fingering and increases the contact area of the liquid with the porous medium. As a result, an effective implementation of PPT increases the flow rates of hydrocarbons during production.

Pressure Pulse Technology renders CO₂ injection, water injection and polymer flooding more effective because it creates a uniform injection front [64]. When it is applied in injection wells during water flooding it increases the oil production and decreases the water cut. The method is best implemented in unconsolidated sediments and sedimentary rocks that exhibit elastic properties.

Incorporation of PPT with CHOPS has been tested on the field and the first trial that took place in 1999 was in the Morgan field, a CHOPS field 30 km northwest of Lloydminster. The trial period was 10.5 weeks and the target was 580m deep in a reservoir with 30% porosity sand 88% saturated with 10600cP viscous oil. One well was used for excitation and at the end of the period the depletion curves of most of the 13 surrounding wells were reversed, production rates had increased by 37% and sand production had increased as well [72]. It has generally been found that by applying pressure pulse for 5 – to 30 hours can re-establish economic production rates during CHOPS for many months, even years [73]. In heavy oil reservoirs in Alberta the implementation of PPT has shown reduction in depletion rates, increase in oil recovery ratio and extension in the life of the wells [5].

Enhanced Oil Recovery Methods (EOR), or Tertiary recovery methods intend to increase the mobility of the oil in order to increase its extraction. These methods are classified into thermal, chemical and miscible displacement methods. The miscible displacement methods include injection of hydrocarbon gas, CO₂, or inert gas under high pressure [74]. These three categories of EOR methods are applied in order to extract some of the petroleum that remains in the reservoir after the primary and secondary recovery methods. The rest of the amount that is recoverable is determined by a number of factors including the permeability of the rock, the intensity of the natural drives and the viscosity of the oil, the density differences of the phases, the capillary effects of the fluid interfaces etc. When the viscosity of the oil is high, as in the case of heavy and extra-heavy oils, it is hard for it to flow and be produced. As a result significant portions remain in the reservoir and can't be extracted. For this reason, reservoirs containing viscous crudes are treated with Enhanced Oil Recovery techniques even from the beginning of production [5] [60].

Heavy-oil production is based largely on the thermal stimulation of the reservoir in order to reduce oil viscosity and thus to induce flow. Some of the practices that are commonly employed over this method are Cyclic steam stimulation (huff 'n' puff), steam flooding and wet or dry combustion with air or oxygen injection (*In-situ* combustion). [75]. Besides the thermal EOR processes there are certain cases that non-thermal techniques appear to be highly effective. The most important EOR techniques that are applied today in heavy and extra heavy oil reservoirs are presented analytically in the next paragraphs.

NON-THERMAL EOR METHODS

Alkaline (Caustic) Flooding is a type of chemical non-thermal EOR method that involves the injection of sodium hydroxide (NaOH), sodium carbonate (Na₂CO₃) or sodium orthosilicate (Na₄O₄Si) along with water injection operations. These chemicals react with the acidic components of the crude oil, create surfactants that are adsorbed in the oil water and sand interfaces and reduce the interfacial tension while raising the pH of the injected water. In the case of heavy oils the addition of these chemicals improves the mobility ratio between oil and water and the result is more efficient displacement of the oil to the production wells [76]. However this process is not suitable for carbonate reservoirs because calcium ions react with the alkaline chemicals and create calcium hydroxide (Ca(OH)₂) which precipitates and may damage the formation [58].

Carbon Dioxide Flooding is a recovery method during which CO₂ is introduced into the reservoir. Due to its very low viscosity, carbon dioxide tends to bypass the oil and reach the production wells leaving a large percentage of the reservoir unswept. This problem is particularly intense with very viscous oils where the viscosity differential between the CO₂ and the oil is large and thus, viscous fingering is severe. Additionally, CO₂ escapes easily through relatively permeable regions of the reservoir and significant amounts of recoverable oil remain untouched. Also, there is always the possibility of the low density CO₂ to move in the upper parts of the reservoir. The latter phenomenon is known as gravity override [77]. The carbon dioxide flooding is shown in Figure 31.

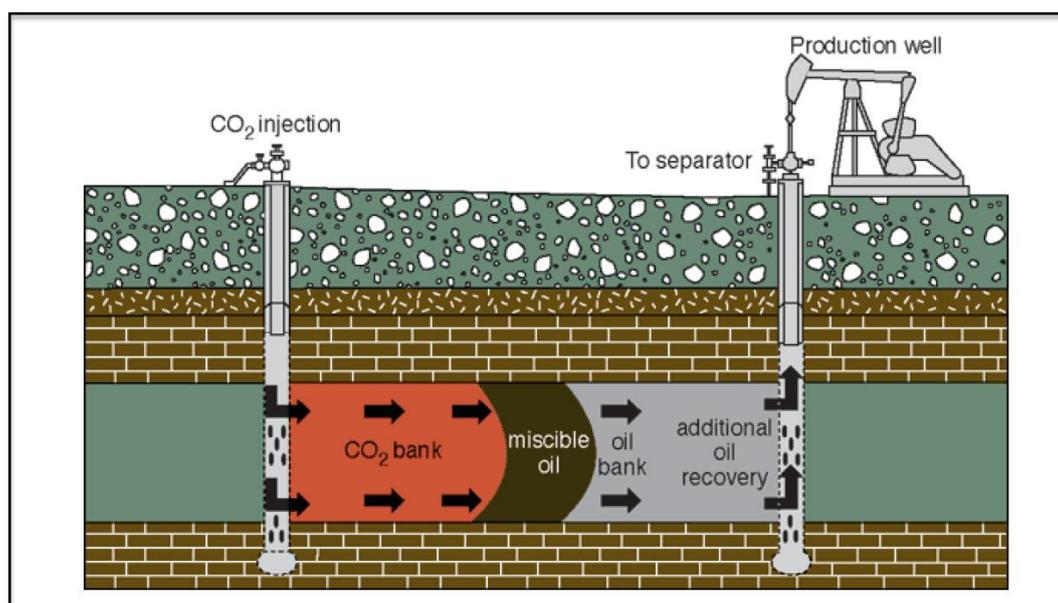


Figure 31: Carbon Dioxide Flooding, Buchanan, R., and Timothy, R., 2011, *Geologic sequestration of carbon dioxide in Kansas*, November 8, 2011, [78].

Even though carbon dioxide is not miscible with the oil on first contact, by forcing it into the reservoir a miscible zone is created from gradual transfer of small and light hydrocarbon molecules from the oil to the carbon dioxide. Provided that the pressure and temperature conditions are favorable carbon dioxide gas is progressively enriched with the lighter fractions of the oil until it condenses into liquid which is miscible with the rest of the reservoir oil. The result is an oil product of increased mobility, lower viscosity and low interfacial tension that flows more easily to the production wells. During fluid production from

the reservoir, the dissolved carbon dioxide re-evaporates to gas phase providing a form of gas lift due to pressure reduction [5].

It is significant to mention that whether the injected carbon dioxide becomes miscible with the oil or not depends on the pressure and temperature conditions as well as the properties of the oil [74] [79]. The CO₂ minimum miscibility pressure (MMP) is the lowest pressure at which multy-contact miscibility can be achieved between carbon dioxide and oil and these two fluids can be mixed in all proportions without the existence of an interface.

At reservoir pressures above MMP, miscible displacement takes place. The mass transfer between the CO₂ and the oil allows the two phases to become completely miscible (one phase) without any interface. The created transition zone is miscible with the oil in the front and miscible with the carbon dioxide in the back [74]. In order for the miscible displacement process to be effective, it must be conducted at or slightly above the MMP when the CO₂ can achieve high degree of solvency with the reservoir oil [5]. This happens in reservoirs that have a minimum depth of 3000 feet (900m) because the pressures and temperatures are then high enough to favor the miscibility of carbon dioxide with the reservoir oil and therefore accommodate high-pressure CO₂ injection.

For reservoir pressures below MMP, immiscible displacement of the oil takes place, which means that the oil and the carbon dioxide cannot form a single phase. However, CO₂ dissolves partly in the oil which then becomes saturated with carbon dioxide and leads to oil swelling, viscosity reduction of the swollen oil/carbon dioxide mixture and reduction of interfacial tension [5] [74] [79] [80]. Solution gas drive is also present as a driving mechanism during immiscible displacement and its combination with oil swelling and reduction of viscosity and interfacial tension enables the reservoir oil to be mobilized and produced.

There are cases where it is difficult to distinguish whether a CO₂ flooding leads to miscible or immiscible displacement process as a result of pressure maintenance failures of the reservoir and/or loss of injectivity [81]. Generally, miscible displacement process is a much more preferred process compared to immiscible displacement. The miscible process is usually applied in light and medium API crude oils whereas immiscible CO₂ process may be applied when the reservoir oil is heavy and cannot easily achieve miscibility with CO₂. Reservoirs that are subjected to immiscible carbon dioxide flooding are usually those containing heavy oils and those that are very thin [82]. In thin reservoirs in particular CO₂ flooding is preferable to the thermal methods because of extensive vertical heat losses [83].

One recovery method that is usually employed in order to reduce gas channeling (viscous fingering), improve sweep efficiency and overcome gas override is Water-Alternating-Gas (WAG) [77]. In this process, a predetermined volume of CO₂ is injected in cycles alternating with equal volumes of water. Another method to increase the CO₂ viscosity in order to avoid viscous fingering is by combining chemicals with gas recovery. This can be achieved by adding surfactants that generate CO₂ foams and thus the viscosity of the injected fluid is significantly increased [84].

Vapor-assisted petroleum extraction (VAPEX) utilizes two parallel horizontal wells where the injection well is located above the production well. Their vertical distance is about 15 ft. VAPEX process is relatively cold (40°) and solvent-based. The vaporized solvents that are injected from the injection well are near dew point, include ethane, propane, butane and naptha and they create a “vapor chamber” as it is demonstrated in Figure 32 [85] [86]. Vaporized solvents start to dissolve in the oil and move towards the cap rock (or any other physical barrier) where they spread along it. At some point they condense into a liquid and the oil and solvent mixture starts to move downwards, because of gravity drainage forces, to

the production well. The space between the wells depends on the viscosity of the oil. The more viscous the oil the smaller the spacing should be. The length of the horizontal wells is also important because long wells can spread more solvent in the porous media.

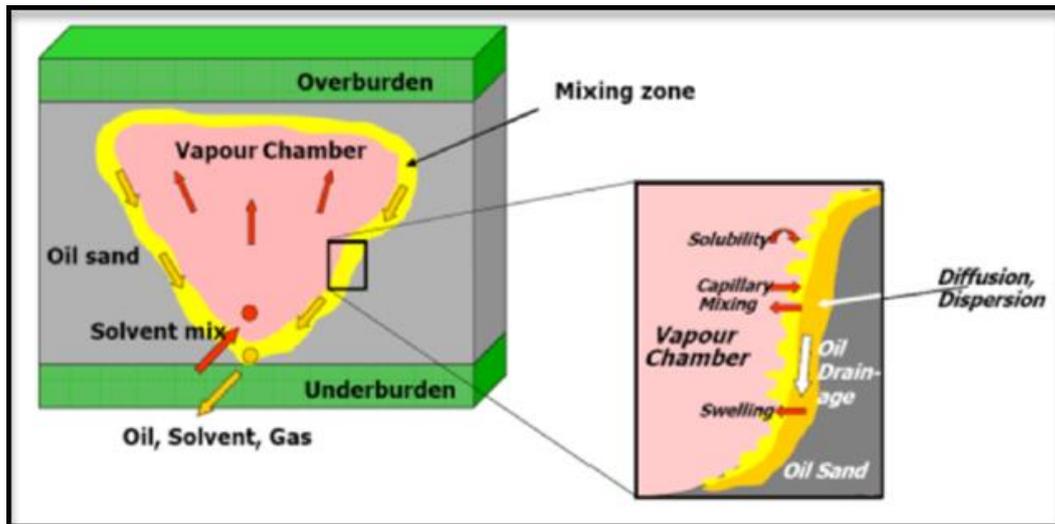


Figure 32: VAPEX process_Source, Canadian Heavy Oil Association

The advantage of Vapor Assisted Petroleum Extraction is the significantly low energy costs. Furthermore, the use of the light hydrocarbon solvents in the case of a heavy oil reservoir can cause the precipitation of asphaltic constituents increasing the potential of in situ upgrading [87]. One more benefit of VAPEX method is that it can be applied in thin reservoirs [5].

Problems that VAPEX method may encounter are possible interaction of the solvent with reservoir minerals since clays adsorb organic solvents very strongly, the loss of solvent through existing faults and the pollution of aquifers [88].

VAPEX can be integrated with SAGD (described later in this chapter) by cycling between steam and miscible gases, by the use of a mixture of steam and solvent or by the injection of heated gas ("warm" VAPEX).

THERMAL EOR METHODS

Cyclic Steam Stimulation (CSS) is a method also known as the "Huff and Puff" method and consists of 3 stages: injection, soaking, and production. Firstly, steam is injected into the production well for a period of weeks, even months, so the formation is heated and the petroleum is liquefied from the heat. Next, the well is allowed to sit for days to weeks, for thermal dissipation and equilibration. At the final step, hot oil is pumped out of the well for a period of weeks or months. Once the production rate falls below a predefined value, another cycle of injection, soak and production is repeated.

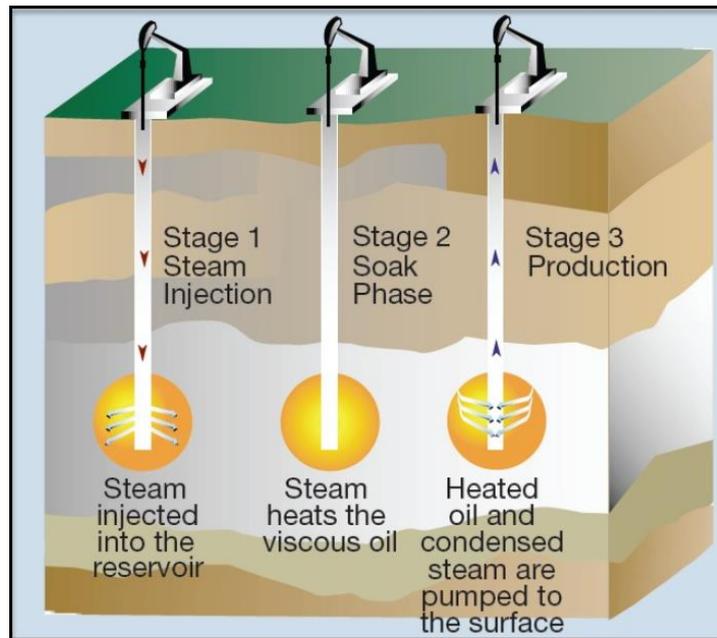


Figure 33: CSS method, Illustration of the Cyclic Steam Stimulation process. Courtesy of the Canadian Petroleum Producers

In **Steam Flooding** different wells are used for steam injection and for oil production. Two mechanisms are at work to increase the amount of oil recovered. The first is to heat the oil and decrease its viscosity so that it will flow more easily towards the producing wells. The second mechanism is the physical displacement in which oil is meant to be pushed by the steam to the production wells. While more steam is needed for this method than for the cyclic method, it is typically more effective at recovering a larger portion of the oil.

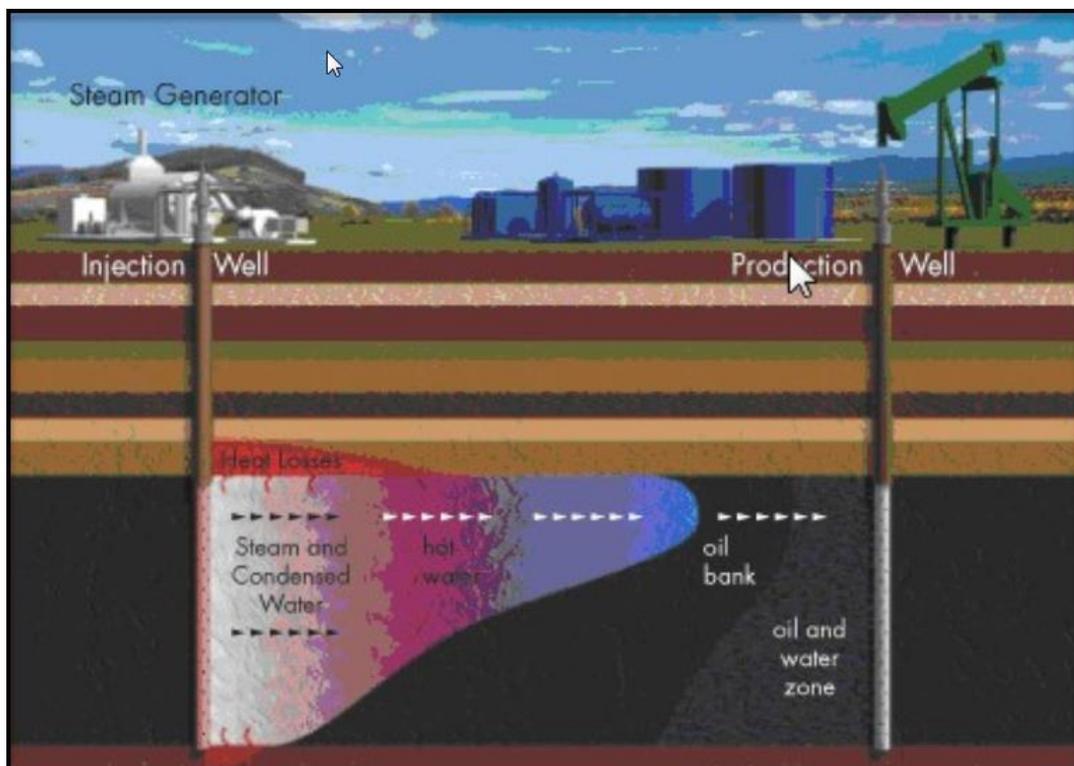


Figure 34: Steam flooding method. Source www.petroleumsupport.com

Steam-assisted gravity drainage (SAGD) was developed in the 1980s to aid in the extraction of oil from the Alberta Oil Sands in Canada. SAGD involves steam injection for viscosity reduction and gravity drainage. To achieve this, two horizontal wells are drilled parallel to one another with a 4 – 6m spacing. Steam is injected into the upper injection well and heats the crude to reduce its viscosity. The liquid crude flows with the assistance of gravity to the lower well where it is collected along with condensed steam and is pumped to the surface. The upper well is often referred to as the “steam chamber” due to the continuous injection of steam which expands in the reservoir. As it is easily understood, existence of vertical permeability is very important for this process to work. [89] [58].

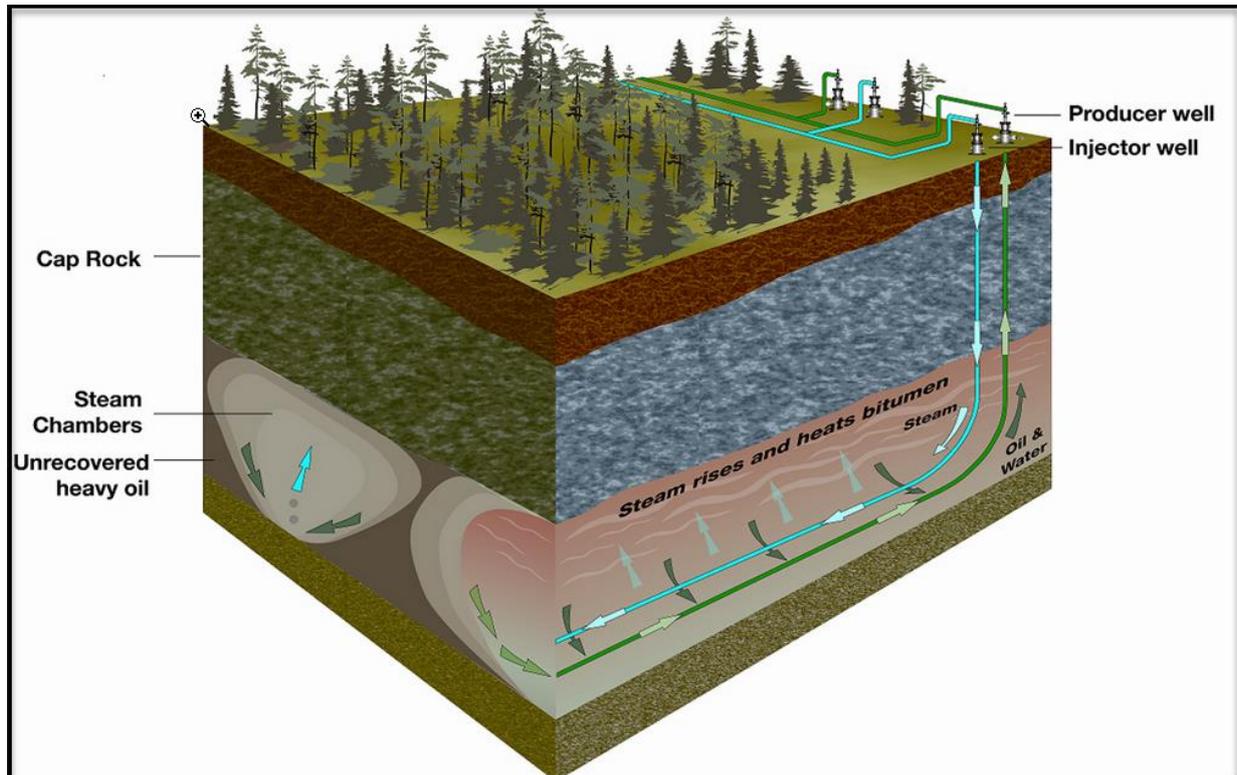


Figure 35: SAGD process, In situ heavy and extra heavy oil recovery Kun Guo et al, 2016 (source: <http://gptarsandsresistance.org/breaking-gptsr-brings-oil-spills-home-for-devon-energy>)

Apart from higher production efficiency, SAGD is also a stable process because it is gravity driven and not pressure driven. This means that instabilities such as coning, fracturing and channeling are absent from this procedure. The produced light gases like hydrogen sulfide, methane, and carbon dioxide rise to the top of the steam chamber and fill the void space acting as an insulation blanket. This reduces significantly the vertical heat losses. The oil and condensate move in the lower part of the chamber by means of gravity and are produced.

A major disadvantage of SAGD is the high cost of steam generation, together with the emission of significant quantities of greenhouse gases, mostly of carbon dioxide. Another disadvantage of this process is the requirement for large volumes of high purity water. This water must be eventually separated from the produced oil, along with the formation water. This method is also not applicable to thin reservoirs where there is not sufficient space for separation between the two wells.

Variations of SAGD include: FAST-SAGD which is a combination of SAGD and CSS wells drilled at the same depth as the SAGD production well and a combination of SAGD and VAPEX which is called Expanding Solvent Steam Assisted Gravity Drainage (ES-SAGD) and includes the injection of both steam and solvent.

In-situ combustion is continuous injection of an oxidizing gas (air or oxygen-enriched air) into the reservoir which ignites by contact with the resident oil creating a burning front that propagates itself through the reservoir. Most of the oil is driven towards the production well by a combination of gas drive (from the combustion gases), steam and water drive. Combustion can be further characterized as “dry,” when only dry air or enriched air is injected and “wet,” when air and water are co-injected.

In-situ combustion has also important benefits. Due to the high temperatures prevailing in the combustion zone, that can reach 345 - 650°C (650 - 1200°F), the heavy oil molecules crack into lower-boiling products and as a result the quality of the produced oil is improved. In this manner a partial upgrading of the oil in the reservoir is realized. Heavy oil upgrading is of big importance and numerous field observations have indicated an upgrading of 2 - 6° API for heavy oils undergoing such type of production. Furthermore, burning a portion of the oil in place to generate heat reduces the cost and avoids the disadvantages of generating steam at the surface.

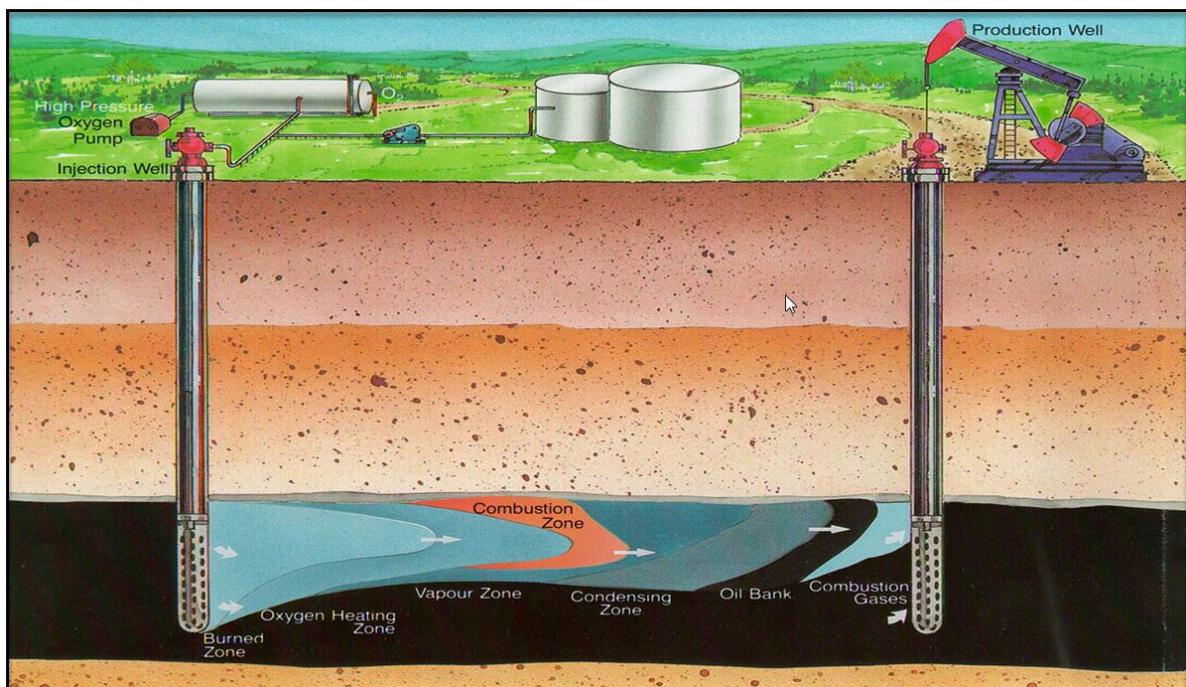


Figure 36: ISC process, source www.insitucombustion.ca

A difficulty of this method is the formation of oxidized hydrocarbons in temperatures below 300°C due to oxygen addition reactions (low temperature oxidation). These hydrocarbons are characterized by increased polarity and higher viscosity than before the oxidation occurs [90]. Further difficulties in the application of the in-situ combustion include the effective control of the fire front towards production wells, the breaking of the formulated oil-in-water emulsions which are contained in the produced fluid and the removal of the heavy-metal compounds that leach from the rock matrix into the production fluids. In addition, production wells should be designed to withstand high thermal stresses.

In-situ combustion has been effective so far for the recovery of viscous oils and it is mostly attractive choice for deep reservoirs. It usually involves no loss of heat and the thickness of the combustion zone is variable depending on each production case. [60] [91].

Toe-to-heel air injection (THAI) is a combustion process with an adjustable fire front which produces an oil partially upgraded and suitable for refinery use.

Toe-to-Heel Air Injection is a combination of a vertical air injection well and a horizontal production well. Combustion products and warm hydrocarbons are swept from the toe to the heel of the horizontal producing well. This method provides in-situ partial upgrading of the oil and increased recovery.

A major advantage in THAI process is that oil recovery occurs through a short displacement mechanism: With gravity, oil flows downward travelling a distance of 15 – 30 ft to the horizontal production well, rather than having to channel through several hundred feet as in the usual combustion process [92]. The process has also the potential to operate in lower pressures and in thinner or deeper reservoirs than the usual steam based processes. No water or gas fuel is required during production and the produced water can be efficiently treated. Furthermore, the production of sulfur and heavy metals is reduced since they remain in the reservoir rock.

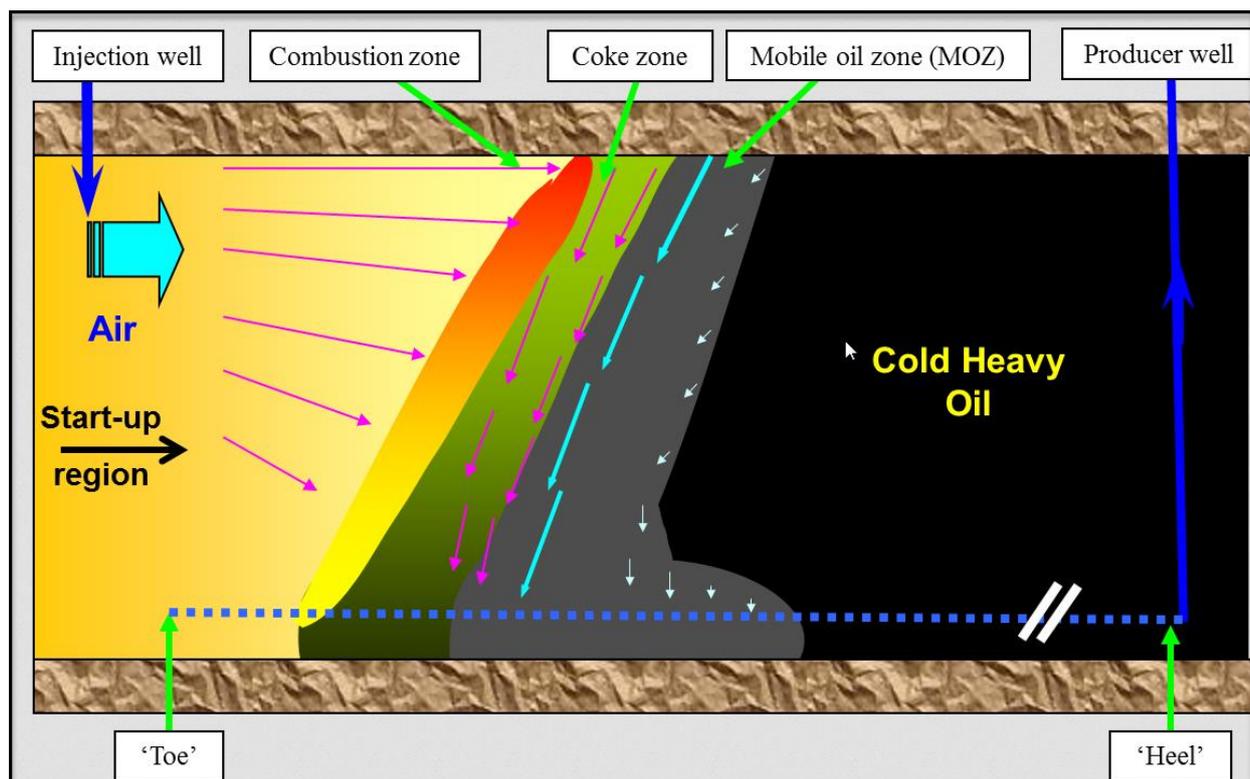


Figure 37: THAI process, Schematics of THAI process in a vertical section (courtesy of University of Bath, UK)

Catalytic upgrading Process In situ (CAPRI) is a variation of THAI method that achieves further upgrading of the heavy oil. The method is based on a catalytic porous layer which is placed between the concentric slotted liners of the horizontal production well. The oil enters the horizontal production well after passing through the catalytic layer, where thermal cracking and hydroconversion reactions take place due to the high pressures and temperatures. CAPRI laboratory tests show significant oil upgrading of about 7° API [93]. The combination of THAI-CAPRI process can be applied to shallow reservoirs with high pressures and achieve operating temperatures of about 400-600°C [94].

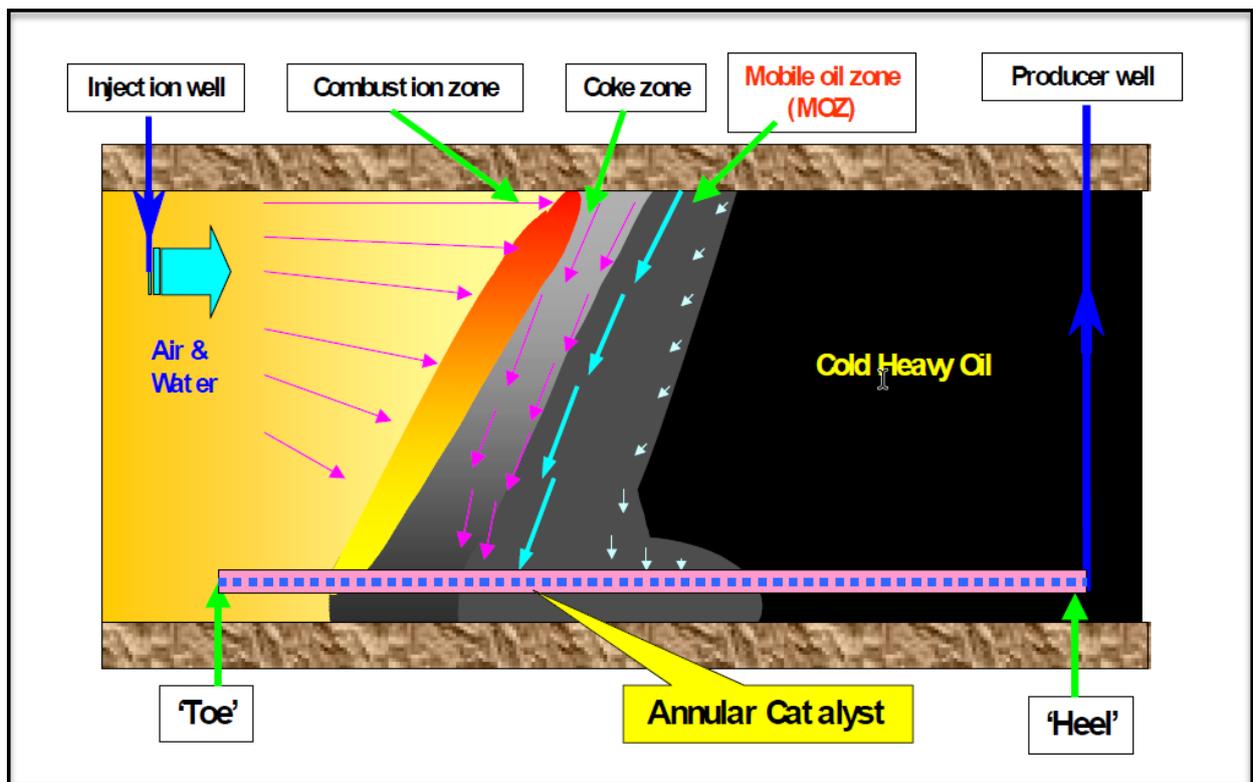


Figure 38: CAPRI process, Schematics of CAPRI process in a vertical section (courtesy of University of Bath, UK)

Petrobank Energy and Resource Ltd. successfully implemented a pilot field trial of the THAI–CAPRI process at Whitesands near Conklin, Alberta, Canada. The P-3B well - as it was called - has been on continuous production of up to 400 barrels per day. They found that THAI–CAPRI effectively increased the produced oil's API gravity by about 2–5° API, the viscosity was further reduced by 10–25 % and the impurities were reduced as well compared to the results obtained when the THAI process was used alone. This significant reduction of viscosity increased the fluidity of the oil, thereby enhancing recovery and pipeline transportation [94] [95] [96].

The advantages in CAPRI method are gravity drainage, small displacement distance for the oil to reach the production well, in situ upgrading, minimization of contaminants associated with heavy oil processing.

The disadvantages are the fast deactivation of the catalyst due to asphaltene and coke deposition and the plugging of catalyst pore matrix leading to high pressure drop and eventually to shutdown of the process for workover operations [94].

All the aforementioned thermal EOR methods can be also classified by the displacement direction of the oil from the injection to the production well. Steam flooding, ISC and CSS use vertical injection and vertical production wells thus they require a long-distance displacement of the mobilized oil to reach the production well. On the other hand, SAGD, VAPEX and THAI methods utilize horizontal wells that require short-distance oil displacement. A schematic representation of the above is given in the figure below.

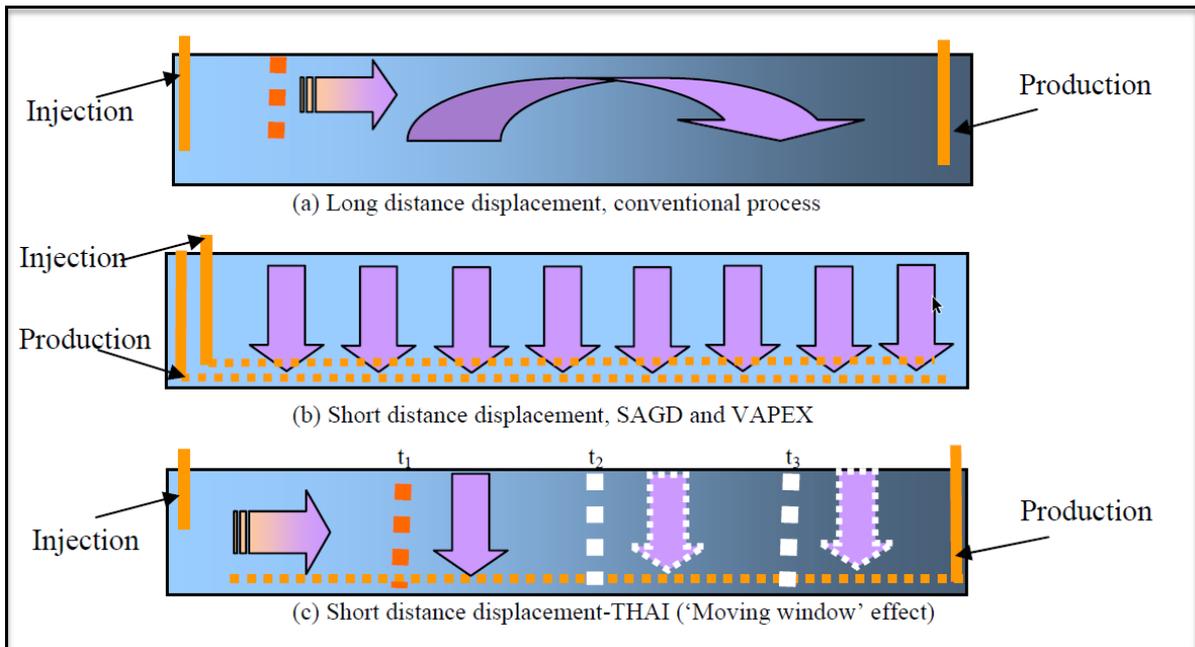


Figure 39: long and short displacement of the mobilized oil, Toe-to-Heel Air Injection, Maximizing Heavy Oil Recovery & In situ Upgrading, University of Bath, UK.

Combination of proven and emerging technologies can be used in hybrid modes to achieve better recovery. For example, CHOPS gives high early production rates, but SAGD gives better overall hydrocarbon recovery, suggesting subsequent or simultaneous use of both methods. SAGD and other thermal methods are very inefficient in reservoirs less than 15 m thick, whereas CHOPS and pressure pulsing technology (PPT) have been successful in such cases.

The following diagram includes all Heavy and Extra-Heavy oil extraction methods

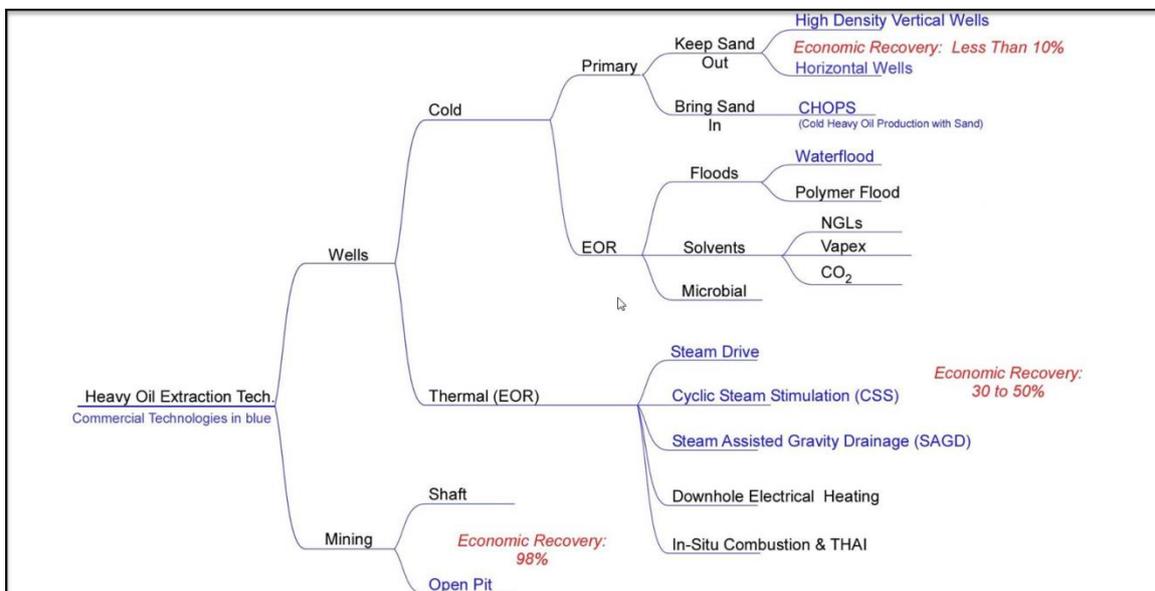


Figure 40: Heavy Oil Challenges & Opportunities North Slope Alaska, Gordon Pospisil - BP Exploration (Alaska) Inc. January 6, 2011

OIL SEPARATION PROCESSES

Oil wells produce a mixture of hydrocarbons, gas or oil, water with dissolved minerals and salt, non hydrocarbon gases including nitrogen, carbon dioxide (CO₂), hydrogen sulfide (H₂S) and solids that can be sand from the reservoir, dirt, scales and corrosive products from the tubing.

The hydrocarbons (gas or oil) must be efficiently separated from the water (dewatering process), the solids and the other impurities (desalting and desanding processes) in order to meet standard specifications. The procedure described is the primary process that is achieved in an oilfield facility and incorporates all the aspects of oil treatment. The secondary process involves water and solid treatment for disposal. Water treatment involves the removal of dispersed and dissolved hydrocarbons (deoiling process), solids and fines from the produced water and solid treatment involves the removal of oil from produced solids. The aforementioned processes are presented in (Figure 41).

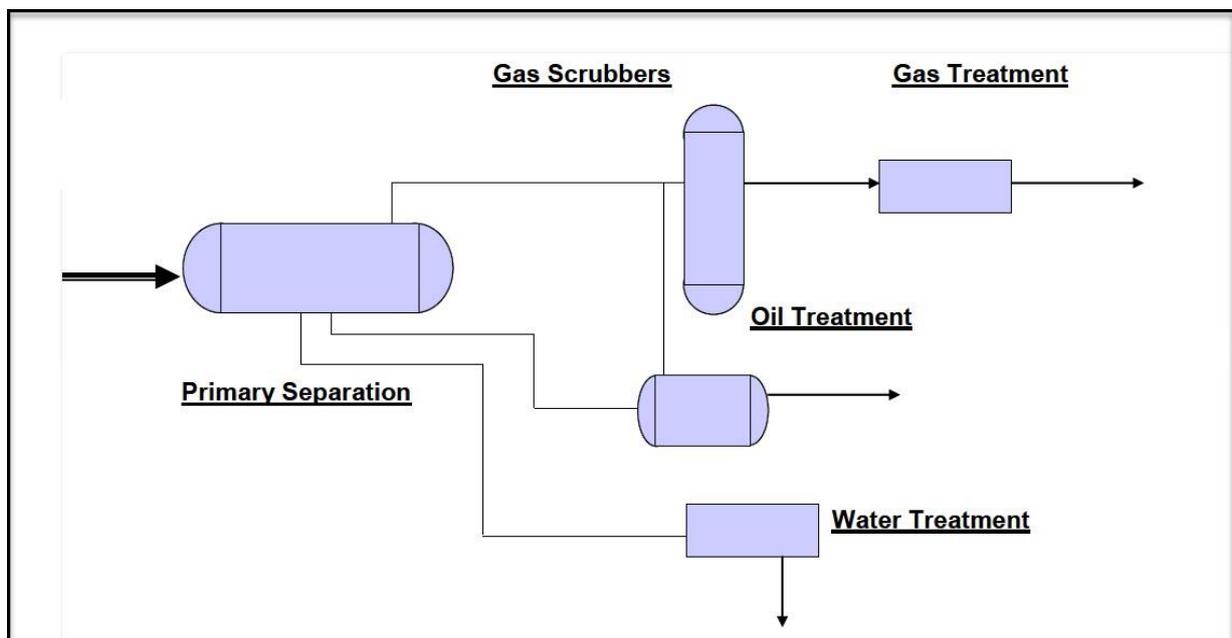


Figure 41: Simplified Schematic illustration of main processes in a separation facility, FMCTechnologies for heavy oil separation_19_Storvik-Subsea separation

The separation depends mainly on the density difference between the phases and it takes place in specialized pressure vessels, the gravitational separators. Separators are the first piece of equipment that the production fluids meet after the wellhead in an onshore as well as offshore platform.

It can be readily proven that multistage separation processes yields in greater volumes of liquid hydrocarbons and reduced volumes of gaseous hydrocarbons than a single stage separation [52]. The hydrocarbon fluid properties along with the wellhead pressures determine the number of separation stages and process conditions under which the oil/gas/water separation is carried out. A typical separation train might consist of a HP (High Pressure) separator (ex.1,100 psig), followed by an IP (Intermediate Pressure) separator (ex. 450 psig), a LP (Low Pressure) separator (ex. 150 psig), and possibly an oil treater (ex. 50 psig) before the atmospheric tank.

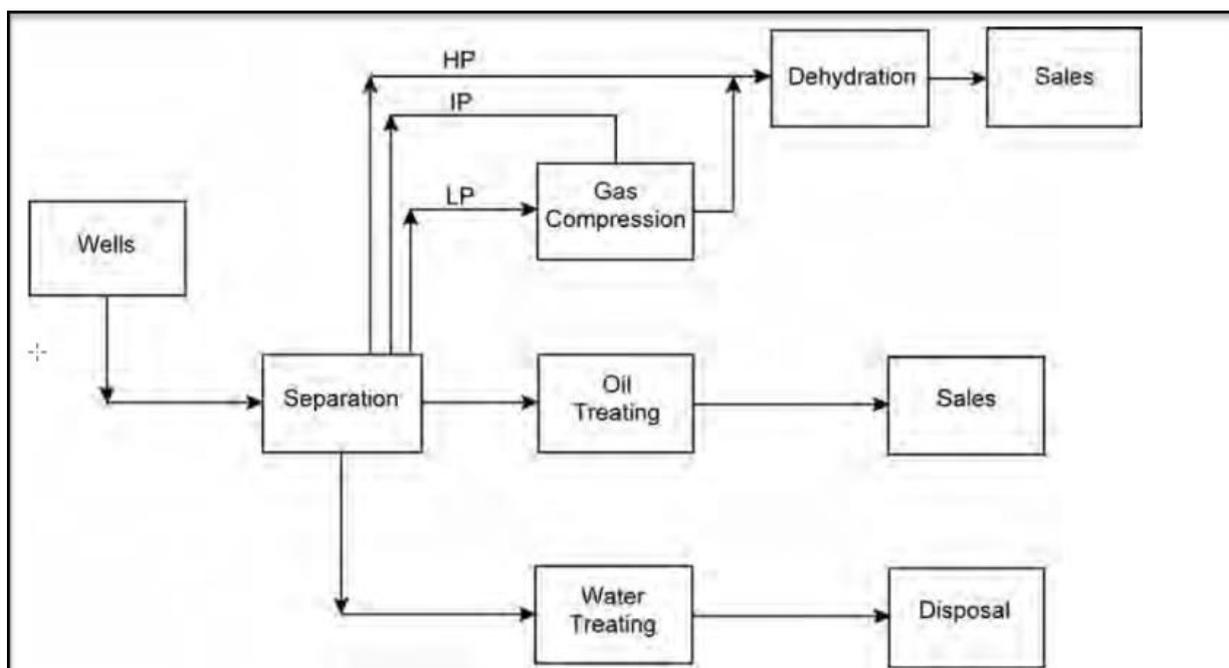


Figure 42: Typical oil facility, *Petroleum engineering handbook Vol.3, Facilities and construction engineering*, Larry W. Lake, [47]

A separator is characterized as two-phase if it separates two phases, such as gas from liquid (oil and water) or three-phase if it separates all phases, gas, oil and water. Produced solids are also separated from the gas and liquid phases by accumulation and removal from the bottom of the vessel. The design and operation of a separator is greatly affected by the characteristics of the flow stream: gas and liquid flow rates (minimum, average, and peak), operating and design P and T, surging or slugging tendencies of the feed streams, physical properties of the fluids, presence of impurities, foaming tendencies of the crude oil, corrosive tendencies of the liquid or gas and designed degree of separation [52].

Based on the vessel configurations separators can be divided into horizontal, vertical and spherical. The most efficient design is selected according to the operating conditions, the cost and the available space.

Horizontal separators (Figure 43) can handle large volumes of gas so they are commonly used for high Gas-Oil ratios. They are smaller and less expensive from the vertical separators for a given flow rate. In a horizontal separator, the liquid droplets fall perpendicular to the gas flow and that makes it easier to settle out of the gas continuous phase. Accordingly, the larger interfacial area of a horizontal vessel compared to a vertical one, aids the gas bubbles to get out of the liquid phase and reach the vapor space. Gas bubbles are spread over a larger area in order to get into the gas phase and that also provides a reduction in foam. In the case of liquid-liquid (oil/water) separation, horizontal separators are more efficient due to the large interface between the liquids.

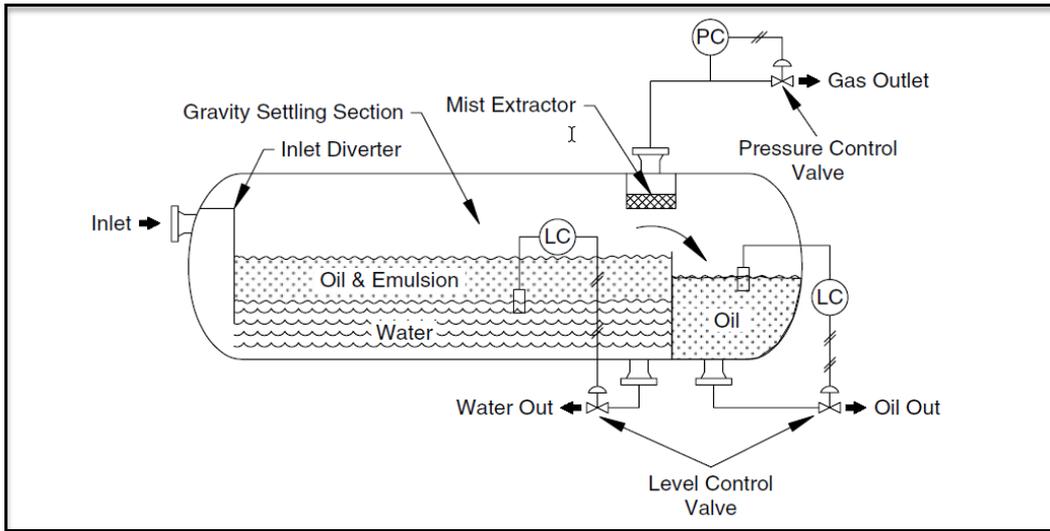


Figure 43: Schematic of a 3-phase horizontal separator, *Surface Production Operations, Design of Oil Handling Systems and Facilities*, Ken Arnold, Maurice Stewart, 2008, [52]

Vertical separators can handle smaller volumes of gas compared to the horizontal separators so they are commonly used for low to intermediate Gas-Oil ratios. For a given flow rate vertical separators have better capacity to absorb slugs of liquid (surge capacity). They are also better at handling solids (sands and sediments) by having an either internal or external cone at the bottom of the vessel that prevents solids build-up. Vertical separators can occupy less space in a facility and that makes them suitable in offshore platforms where there are space restrictions.

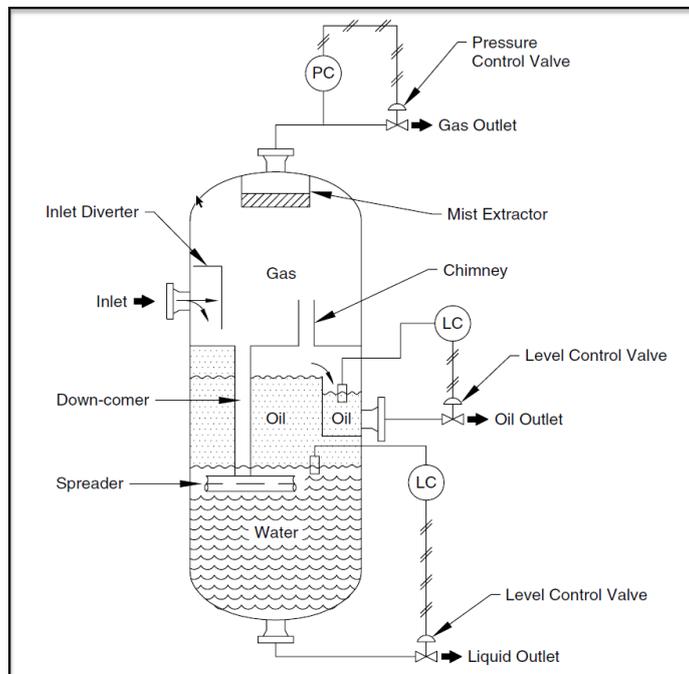


Figure 44: Schematic of a 3-phase vertical separator, *Surface Production Operations, Design of Oil Handling Systems and Facilities*, Ken Arnold, Maurice Stewart, 2008, [52]

Spherical separators were originally designed to withstand very high pressures because their shape increases their mechanical strength. They have more gas capacity than similarly sized vertical separators but less than similarly sized horizontal separators. They also have less surge capacity than similarly sized horizontal separators. Proper design and installation of spherical separators is difficult that's why they are not widely used.

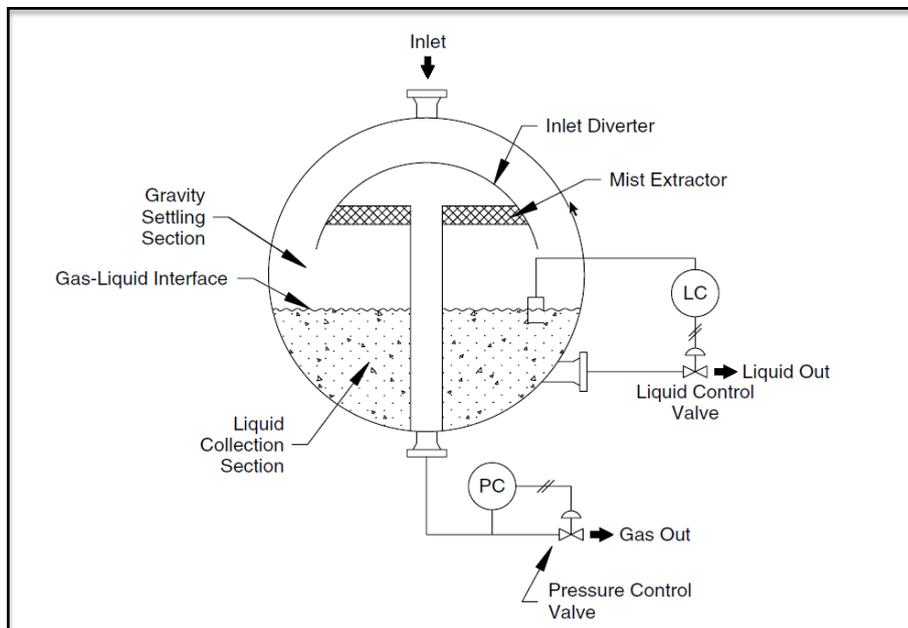


Figure 45: Schematic of a spherical separator, *Surface Production Operations, Design of Oil Handling Systems and Facilities*, Ken Arnold, Maurice Stewart, 2008, [52]

Level control is very important parameter for the proper operation of the separators. Generally, the flow rate of the inlet stream must be equal to the flow rate of the gas and liquid removed through the outlets. When the level rises a control valve opens and lets some liquid out until the liquid returns to the normal operating level. At this point the level valve closes. In a 3-phase separator oil level and water level must be controlled separately so two level controllers and two level valves are used for each phase.

The gas that is contained in a separator is controlled by a pressure control valve. When the pressure in the separator increases the pressure control valve opens and releases the excess of gas. When the pressure returns to the set point the valve closes. These valves are also represented in Figures 43, 44 and 45.

A Free Water Knockout (FWKO) is a 3-phase separator which is commonly used in processing plants when large quantities of water are produced. This can happen for instance in maturing fields that water production is progressively increasing [97], during water flooding, during SAGD when high water cuts are expected mostly in early production or CHOPS that water rates increase over time [98]. In that case the purpose is to remove the free water, i.e. the water which exists either as a continuous phase or in droplets large enough to separate quickly from the oil [99].

The oil and water form three distinct layers while they are settling in the separator; a top layer consisting of relatively clean oil with small amount of water dispersed in tiny droplets, a bottom layer consisting of free water with very little amount of oil dispersed in very small droplets and in the middle a water in oil emulsion layer. The emulsion layer gets thinner when adequate residence time is provided, because the emulsified water droplets coalesce and settle down to the free water phase (Figure 46).

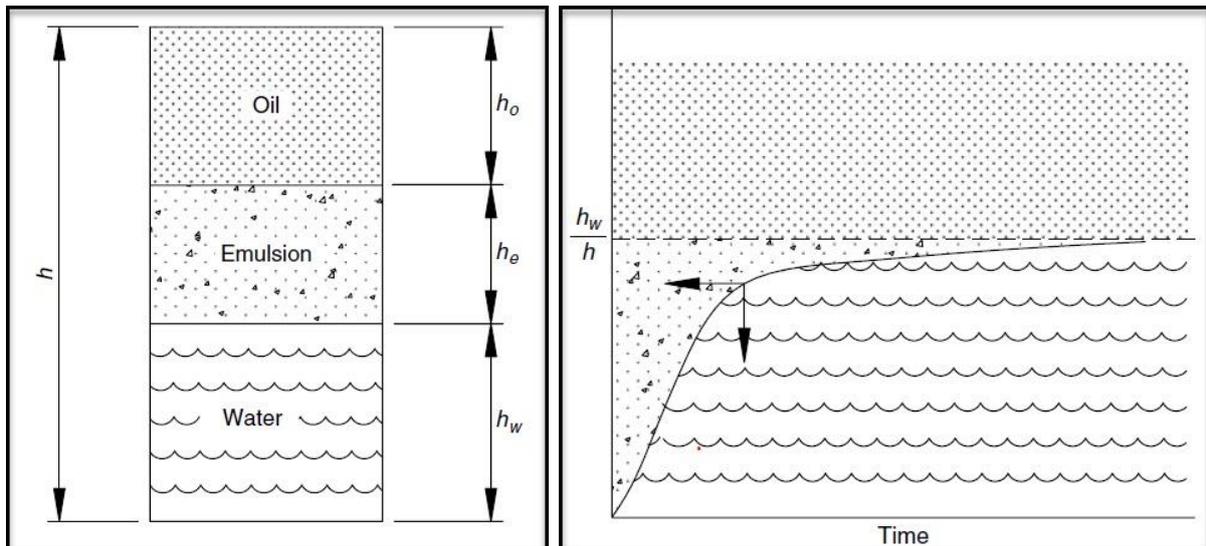


Figure 46: Oil, emulsion and water layers and thinning of the emulsion layer over time, *Surface production operations Vol 1*, [52]

In the case of light oils the water-in-oil emulsion is relatively loose and it is possible to break during the given settling time with the addition of the appropriate chemical demulsifier [47]. However this is not the case with heavy oils which create tight and very stable emulsions that can even withstand the addition of de-emulsifying chemicals. For a more efficient separation a further treatment of the emulsion is required both in heavy and light crude oil processing facilities in order to provide pipeline-quality oil.

Gunbarrels are the oldest equipment in onshore oil treatment and they are very common in heavy crude oil applications. They are nothing more than large vertical flow atmospheric tanks that typically have a gas boot (gas separating chamber) that extends 2–4 m above the top of the tank and a down-comer that leads the water-in-oil emulsion to a spreader placed below the oil-water interface. This configuration forces the emulsion to pass through the free water layer, a procedure known as waterwashing, which helps in the collection of the entrained water and the conversion of the emulsion into distinct oil and water layers. The gunbarrels have very large size and diameter and the settling time is between 12- 24 hours. Most gunbarrels are not heated however heat can be provided to the incoming stream. They are relatively inexpensive but their large footprint prevents their use in offshore facilities [52].

Heaters are vessels used to raise the temperature of the crude oil emulsion to facilitate its treatment. There are two types of heaters, the indirect fired heaters and the direct fired heaters. What is common in both types is the shell and the fire tube which contains hot combustion gases. In the indirect heaters the oil flows through tubes that are immersed in water, the temperature can be kept constant for long time but they need several hours to reach the desired temperature and they are safer than the direct heaters. In the direct heaters the oil is heated directly by a fire box and the desired temperature is reached quickly. Direct heating of the oil is hazardous so direct heaters require special safety equipment [52].

Heater treaters are considered an improvement of Gunbarrels and heaters and can handle various viscosities, densities, high and low flow rates and weather conditions. Compared to the Gunbarrels they are smaller and lighter, less expensive and require minimal retention times (10 – 30 minutes). On the other hand they are more sensitive to chemicals, more complicated and provide less storage space. Horizontal heater-treaters are more common than the vertical ones and they consist of the front section (heating and water wash), the oil surge chamber and the coalescing sections [52].

A Slug catcher is a 2-phase gas-liquid separator that is designed to collect large liquid volumes that may overload the gas/liquid handling capacity of the process facilities. Liquid slugs are often encountered during heavy oil production, as mentioned in previous chapter. They are also the result of uneven distribution of liquid and gas volumes in the pipelines during two phase flow. The liquid slugs burst suddenly and are produced in a rate much larger than the equipment is designed to handle, especially during pigging operations. Pigging is used when the pipelines must be swept of liquids so, consequently, large volumes of liquids must be handled by the downstream separation equipment. Slug catchers provide additional volume that reduces the stream velocity and the excess energy produced by the liquid slug [47]. They are designed in numerous different ways but their mode of operation is the same. The gas and liquid slug enters the two-phase vessel, where gas-liquid separation takes place. Gas exits the top of the separator from the gas outlet while the liquid is distributed in a number of large diameter tubes that are called “fingers” and finally exits from the bottom of the vessel. The produced liquid is routed to the next piece of equipment which is a Free Water Knockout for further liquid-liquid separation [52].

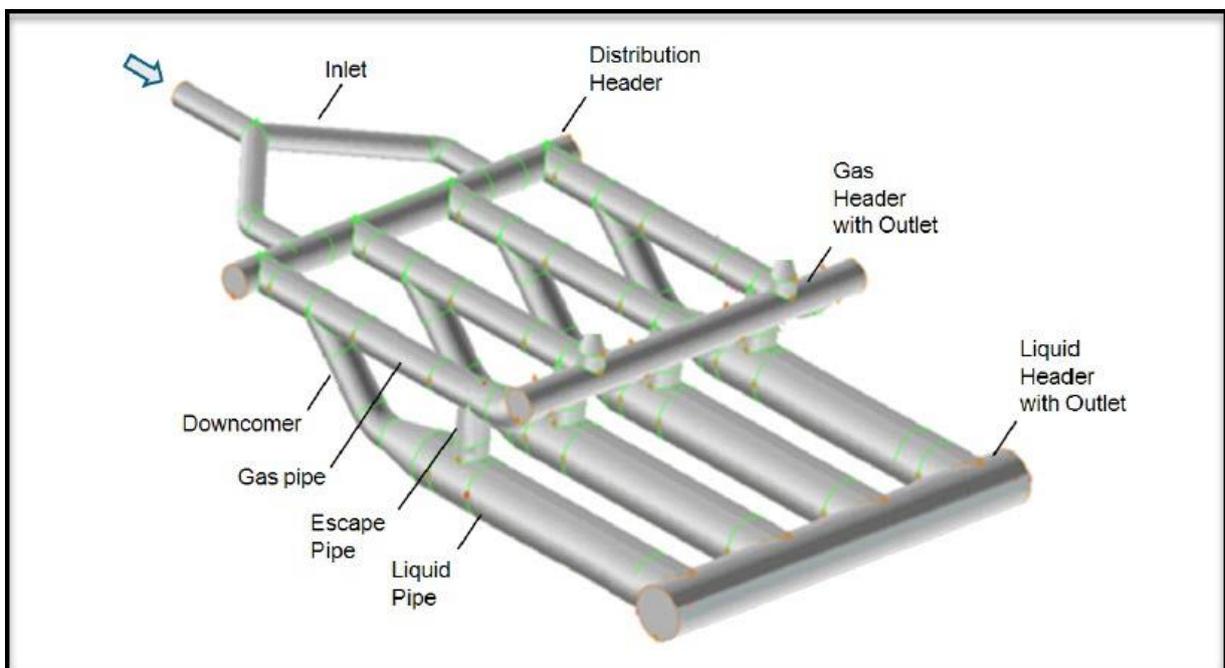


Figure 47: Schematic representation of MultiPipe finger-type slug catcher, *Compact Separation Technologies and Their Applicability for Subsea Field Development in Deep Water*, A. Hannisdal, R. Westra, and M. R. Akdim, FMC Technologies Inc. A. Bymaster and E. Grave, ExxonMobil Upstream Research D. Teng, Woodside Energy Ltd., 2012, *Offshore Technology Conference*, [100].

HEAVY AND EXTRA HEAVY OIL PROCESSING

The treatment of heavy and extra heavy oil is possibly the most important production process and it is directly connected with oil **dehydration** and **desalting**. Dehydration is the liquid-liquid separation of oil and produced water that aims at the reduction of the final water fraction in the crude oil. Desalting is the reduction of the salt content in the crude oil. Since salts are soluble in the water, a reduction of the water content will automatically reduce the salt content as well. If the salt content is high a desalting stage is required, after dehydration, in which the remaining salts are removed by extraction with fresh water in order the crude oil to meet the required salinity specification [101].

The specifications for BS&W (Basic Solids and Water) and salt content may differ from case to case due to local conditions and contractual agreements. A typical range for BS&W specs is from 0.2 to 1% and for salt content ~ 10 – 100PTB [47].

Heavy and extra heavy oil treatment should come across additional challenges compared to the conventional oil treatment due to the often creation of very stable water-in-oil emulsions. A great deal of research has been published concerning the establishment of efficient methods for water/heavy oil emulsion treatment, the most important of which is presented in this chapter.

OIL TREATMENT

DEHYDRATION

Heavy oil dehydration is challenging because crude oil and formation water have similar densities, oil viscosity is high and the droplet size of the dispersed water phase is small. The result is reduced settling velocities and ultimately poor separation, unless very long retention times are applied [102].

Water in oil emulsion is almost always present in the production fluids. The emulsion is either intrigued by the implemented thermal EOR methods, by flow restrictions in valves, chokes and pumps, by the presence of natural surfactants in the heavy oil, like asphaltenes and naphthenates, or by the addition of process chemicals such as emulsifiers, defoamers, corrosion inhibitors, etc [27].

The heavy oil recovery is often characterized by high water cuts and high content of fines which are further increased during the depletion of the mature oil fields. All these have a direct impact on the emulsion and its viscosity which is increased as more water droplets are contained in the emulsion. A 30% water cut may produce emulsions with viscosity which is 40% higher compared to the one of the dry crude [103]. A typical production profile of crude oil and produced water is given in the following Figure 48.

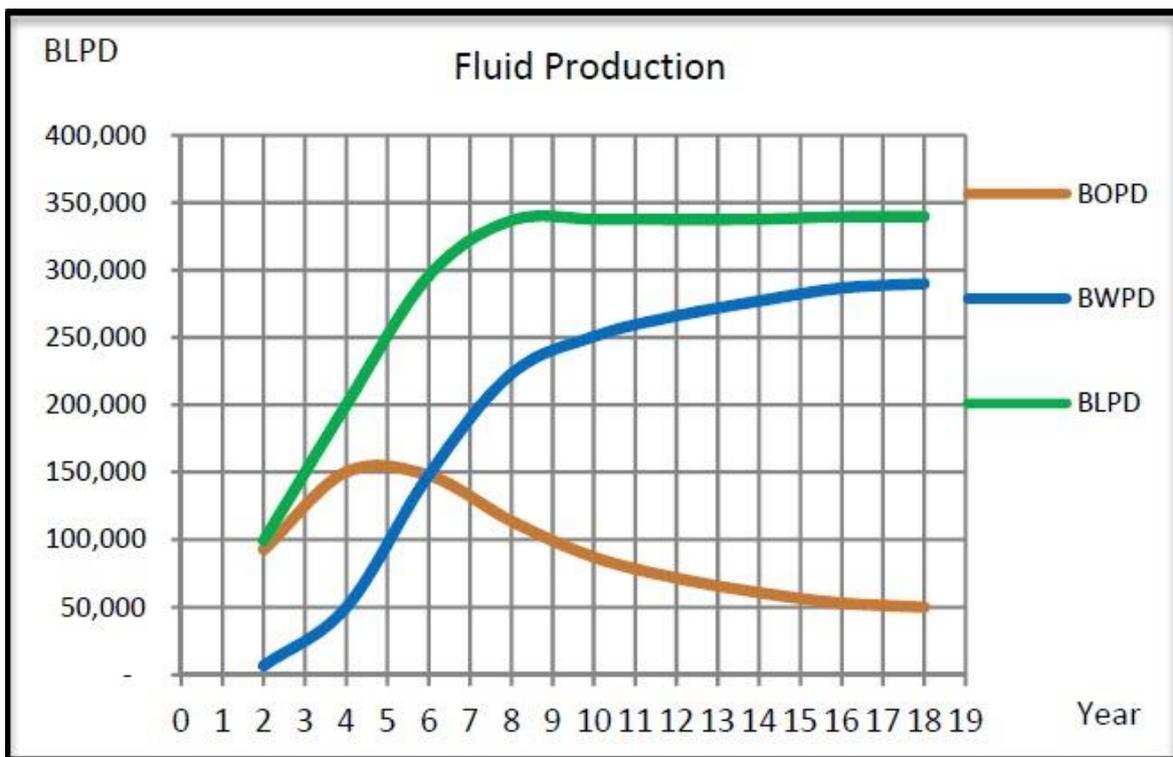


Figure 48: Typical production profile, *Improved Dehydration and Desalting of Mature Crude Oil Fields*, Erik Sellman, SPE, Gary Sams and S. Pavan Kumar Mandewalkar, SPE, *Cameron Process Systems*, SPE 164289, 2013, [103].

The high viscosity of the heavy oil plays a double role on the emulsification process. On the one hand, it interferes with the migration of the emulsifying agents to the water droplet's surrounding film which means that the water droplets can be susceptible to coalescence. On the other hand, the high friction results in low droplet velocities with an adverse effect on their collision frequency [47] [27], resulting in less coalescence and more difficulties in the separation of oil and water phases. However the negative effect of high friction overpowers the emulsifier's migration obstruction, consequently it is very hard to treat high viscosity oils [52]. The viscosity parameter is very significant for crude oils with API gravity between 20 and 11 and in certain cases it gets impossible to treat high viscosity crudes without the addition of a compatible diluent. Diluent addition in crude oils of 14°API and less is a very common practice [52].

The removal of salt – either in suspended or diluted form – from the crude oil is called desalting and is applied in cases where the dehydration solely is not able to reduce the salt content to the specified values. The type and quantities of salts depend on the formation, although their major portions are usually sodium and chloride's ions [104]. Heavy crude oil accompanied by low salinity produced water makes dehydration process challenging due to the low density difference [105]. The following table demonstrates the viscosities and specific gravities of two different API crudes and two different brines (saline water) with respect to different temperatures.

Temp (°C)	API 18		4% Brine Specific Gravity	API 30		12% Brine Specific Gravity
	Viscosity (cP)	Specific Gravity		Viscosity (cP)	Specific Gravity	
40	214.23	0.9308	1.0193	11.51	0.8598	1.0721
60	72.28	0.9180	1.0103	5.30	0.8464	1.0630
80	31.27	0.9052	0.9989	2.96	0.833	1.0516
100	16.15	0.8924	0.9854	1.89	0.8196	1.0381
120	9.48	0.8796	0.9701	1.31	0.8062	1.0229
140	6.12	0.8668	0.9531	0.97	0.7928	1.0059

Table 8: Fluid properties for API 18 crude oil with 4% salinity produced water and API 30 crude oil with 12% salinity produced water, *Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters*, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, Cameron Process Systems, SPE 156299, 2012, [106].

If the properties of Table 8 are projected in a diagram, the density variation of API 18 and API 30 crude oil, 4% and 12% salinity brine and the differential crude oil/brine densities as a function of temperature are shown in the following figure.

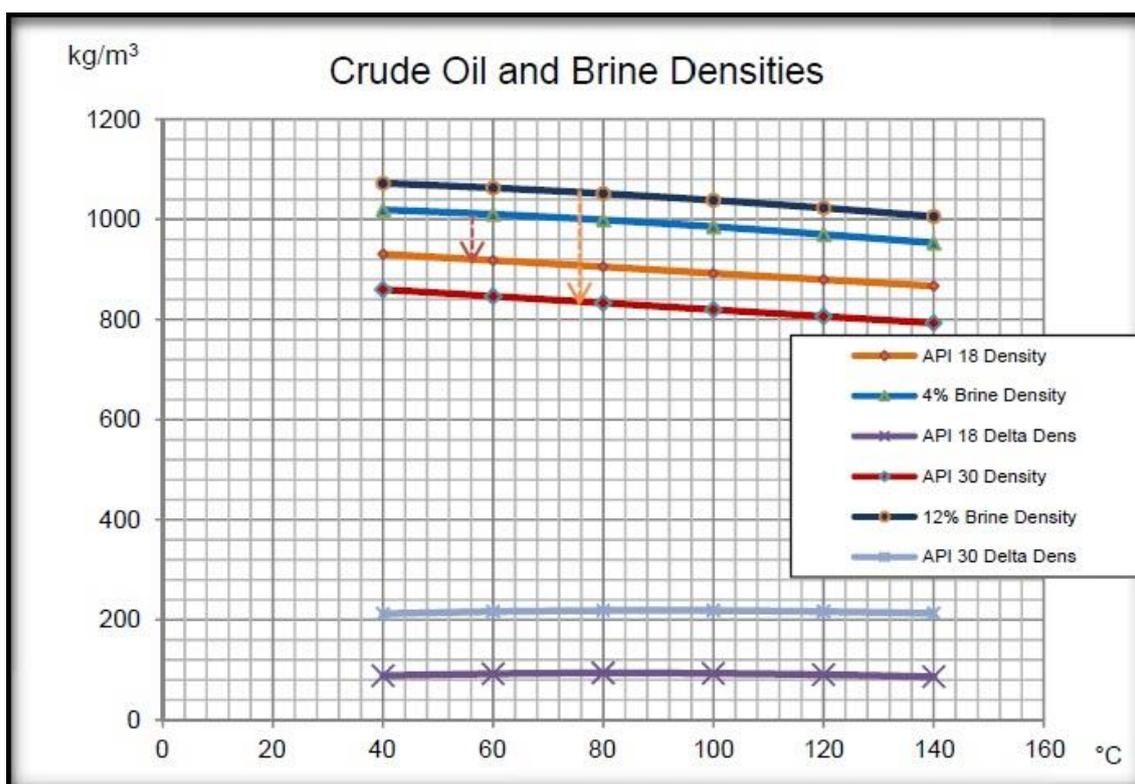


Figure 49: the density variation of API 18 and API 30 crude oil, 4% and 12% salinity brine in addition to crude oil/brine differential densities as a function of temperature, *Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters*, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, Cameron Process Systems, SPE 156299, 2012, [106].

As it can be seen in Figure 49, the lighter API 30 crude with the 12% salinity brine has 2,4 – 2,5 times higher density difference from the API 18 crude with the 4% salinity brine.

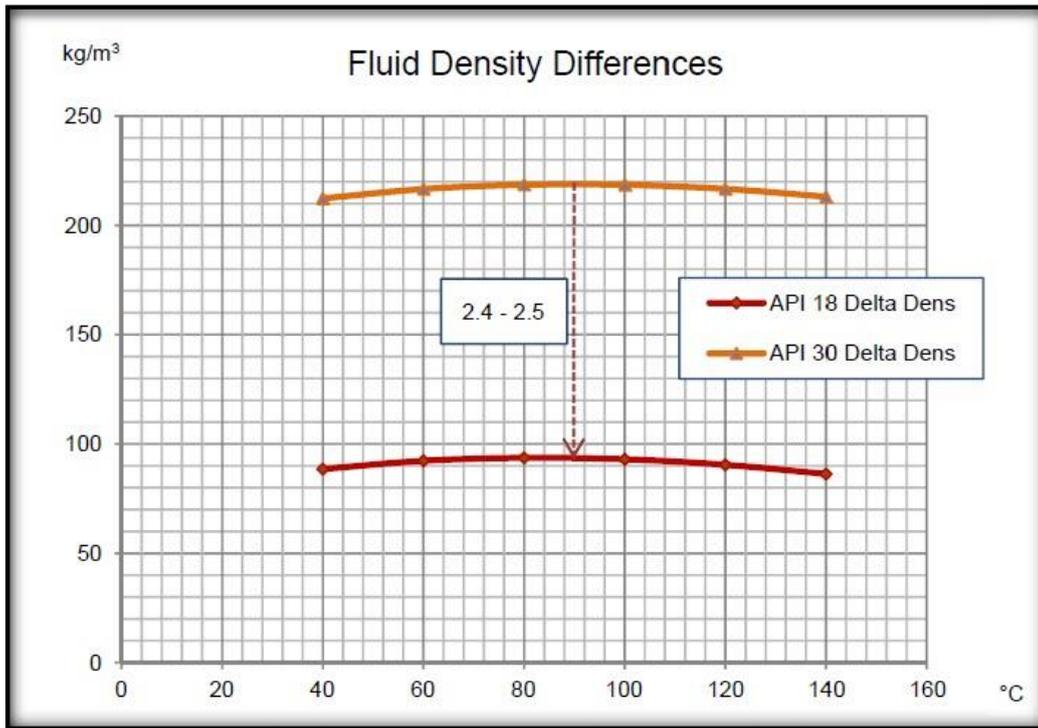


Figure 50: the density difference between the continuous crude oil phase and the dispersed brine phase for API 18 and API 30 crude oils with typical formation water salinities, *Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters*, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, *Cameron Process Systems*, SPE 156299, 2012, [106].

The density differences in Figure 50 show an apparent maximum in 95°C. In conclusion, there is a significant reduction in the separation efficiency when heavier crude oils are mixed with low salinity formation waters. In the following table the density differences between produced water of variable salinity and crude oil of variable API gravity in 95°C are demonstrated

Density Difference PW / Crude (kg/m³)

Density difference is max at 95C $v_s = d^2 (d_w - d_o) g / 18\mu$

PW Salinity %	API Gravity of Crude Oil										
	10	12	14	16	18	20	22	24	26	28	30
0	0	1.4	2.7	4	5	7	8	9	10	11	12
1	16.6	18	19	21	22	23	24	26	27	28	29
3	50	51	53	54	55	56	58	59	60	61	62
5	83	84	86	87	88	90	91	92	93	94	95
7	116	118	119	120	122	123	124	125	126	127	129
9	149	151	152	153	155	156	157	158	160	161	162
11	183	184	185	187	188	189	190	192	193	194	195
13	216	217	219	220	221	222	224	225	226	227	228
15	249	250	252	253	254	256	257	258	259	260	261

Higher density difference with higher crude API gravity and higher formation water salinity.

Table 9: Density differentials between produced water and crude oil in 95°C, *2-Prong Approach for Improved Operating Range of Electrostatic Crude Oil Treaters*, Pavan Mandewalkar and Erik Sellman, *Cameron Process Systems_OTC_26348*, Brazil, 2015, [107].

Interfacial tension is also a factor that directly affects coalescence. It exists when two phases are present and it can be defined as the force that holds together the surface of a particular phase. When the interfacial tension between oil and water is low the water droplets can easily coalesce when they contact each other. When the interfacial tension is high, which can happen when a surfactant is present, the water droplets may aggregate but they don't merge to form larger droplets thus, coalescence is prevented [52]. Therefore, whatever lowers the interfacial tension promotes coalescence and aids phase separation. However, when interfacial tension is too low the water droplets are very easily broken up after they coalesce and form new small droplets. As a result, the interfacial tension must be maintained in a certain range in order not to affect negatively the growth of the water droplets [108] [106].

One more factor that must be taken into consideration during oil/water separation is the age of the water-in-oil emulsion, or, in other words, the time between the emulsion creation and the emulsion treatment. When the emulsion is created, the emulsifying agents that are dispersed in the oil start to gather around the water droplets, increase interfacial tension and impede coalescence. This process of film development continues until there is no more free space on the droplet surface for another stabilizer to attach (Figure 51) [52].



Figure 51: Photomicrograph showing two droplets touching but unable to coalesce because of the emulsifying skin surrounding the droplets, Surface production operations, Ken Arnold, Maurice Stewart, 2008, [52].

A way to prevent emulsion from aging is to treat it as close to the wellhead as possible. Because the high viscosity crudes slow down the emulsifier's migration, the aging progresses are slower than in the lower viscosity crudes.

Besides the physical properties, every crude oil has a chemical composition that makes it unique and in the same manner, the chemistry of the water that is produced with the oil is also unique. Consequently every water-in-oil emulsion is different in its composition, the way it was made, the amount and type of additives, its production process and the operating conditions during processing [109].

SEPARATION METHODS

Production design, treating processes and equipment must be selected after the physical properties of the heavy, viscous oil and water have been accurately determined. However, such properties are difficult to be measured, because it is difficult to obtain a representative sample due to the low mobility, the high compositional gradients, the w/o emulsion formation and the sand production [110]. In order to obtain the oil properties, both water and solids must be reduced to concentration of less than 1% by volume [111]. Further difficulties in oil PVT testing are imposed from the high viscosity, the low saturation pressure and the slow liberation of gas at pressures below the bubble point [112].

The emulsified water content in heavy crudes, with API less than 20, can be from 10 to 35 vol%, while the BS&W (Basic Sediment and Water) limits are set according to local conditions and contractual agreements but typically range from 0.2 to 1%. It is therefore crucial to accomplish complete separation by implementing some form of treatment. [47].

The most common separation methods that are used in water-in-oil emulsions include: gravity settling, centrifugal settling, thermal treatment, filtration, chemical demulsification and electrostatic treatment. Less frequently used methods are pH adjustment, membrane separation, microwave radiation and ultrasonic energy [113] [109] [114]. The conventional separation methods for both light and heavy oils which include gravity and centrifugal settling (mechanical treatment) as well as the mechanical internals commonly used in gravitational separators are described in Appendices A and B respectively. In this chapter, thermal, chemical and electrostatic processes are presented [115]. Special emphasis will be given in electrostatic treatment since it is widely used in heavy oils.

Thermal treatment

The addition of heat to the incoming oil-water stream reduces the viscosity of the crude oil, makes the water droplets move faster thus colliding with greater force and partly deactivates the action of natural emulsifiers like paraffin crystals that dissolve upon heat addition. The action of treating chemicals is also enhanced with the temperature increase, because they work faster and more effectively. According to the API Specification 12L [116] it is recommended that the crude oil is heated so that its viscosity is 50cSt (centistokes, $1\text{cSt} = 1\text{mm}^2\cdot\text{s}^{-1} = 10^{-6}\text{m}^2\cdot\text{s}^{-1}$) for dehydration and for desalting the viscosity should be <7cSt. Treating temperatures usually range from 100-160°F (38-70°C) and in most cases are below 200°F (93°C). For heavy oils below 20°API the treating temperatures are higher than 180°F and sometimes reach 300°F (150°C) which may have negative effect in the density differential [52]. In some cases, very high treating temperatures can lead to a water density lower than the oil density (Figure 52) [47].

Drawback to the heating process is that it is energy intensive and results in a larger system footprint. Fuel is required which often represents an important cost both in capital and in operating expenditures (CapEx and OpEx). Heating can be also impractical in subsea, arctic or remote applications, where the temperatures are too low and the heating of process fluids is difficult [109]. Increased temperatures also result in higher solubility of the water in the oil phase. Then as the oil is transported, it gradually cools down in the pipeline and free water comes out of the solution and separates, a phenomenon which can create serious flow-assurance issues [109].

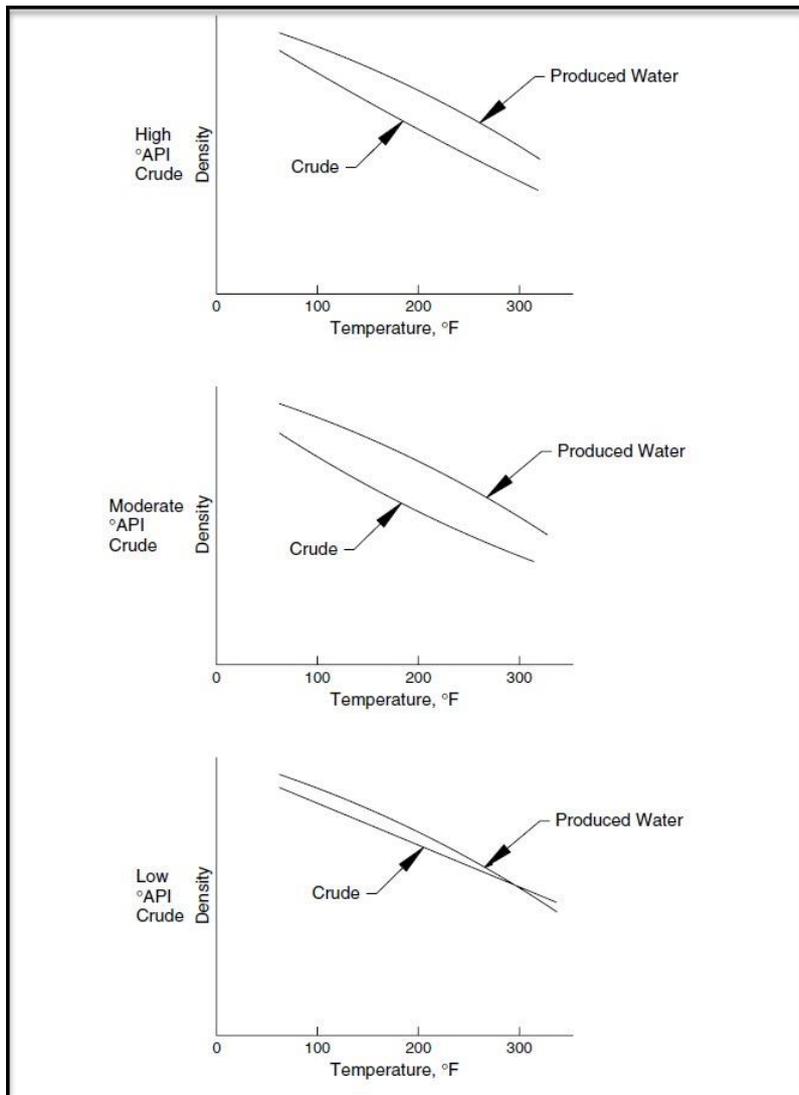


Figure 52: Relationship of density and temperature for three different API gravity crudes and produced water, *Surface production operations, Ken Arnold, Maurice Stewart, 2008, [52].*

Heat can cause the loss of the lower boiling-point hydrocarbons which represent the light ends of the crude. That, on its turn, causes the so called “shrinkage” of the oil, i.e. less produced oil volumes and thus less income. The vaporized hydrocarbons can be collected and sold with the produced gas after they are compressed but even this way there is a net income loss [52] [47]. The liberated gas may produce flow and settling issues, such as turbulence during its rise to the coalescing section or attachment to solids or water bubbles.

Chemical demulsification

Chemicals are added in w/o emulsion to induce coalescence and cause rapid separation of oil and water phases. Their purpose is to reach the droplets, to be absorbed in the surrounding film and cause its rupture. In case there are emulsifying agents surrounding the droplets, the chemical demulsifiers should displace them in order to proceed to the droplet coalescence [52].

A chemical demulsifier is effective when it is strongly attracted to the oil/water interface, it migrates rapidly through the oil and reaches the dispersed droplets. A demulsifier can attract multiple water

droplets with similar charge in order to form clusters. A demulsifier must also promote the rupture of the film between the water droplets and the oil phase if the flocculation force is not enough for that. With film rupturing the droplets in the clusters start to merge and create droplets of larger sizes.

An effective chemical demulsifier must also change the wettability of the suspended solids (iron sulfides, clays, drilling muds) to remove them from the oil/water interface. As the solids get water-wet, they leave the interface and diffuse in the water droplets. In the same way paraffins and asphaltenes can be made oil wet and disperse in the oil phase [47].

Since the field conditions change with production time there is no uniform set of chemicals that covers efficiently the entire life of the field. Seasonal changes or changes in the solid content of the produced fluids can alter the emulsion stability and consequently the chemical requirements. Generally, the amount of chemical that is added varies from 1 gallon per 150 barrels to 1 gallon per 1000 barrels [52].

A thorough mixing of the demulsifier with the emulsion is essential in order to diffuse to the dispersed droplets. When heating is also applied the reduced viscosity facilitates the mixing of the emulsion with the chemical additives and the reactions at the oil/water interface progress more rapidly. The additives should be added at a continuous rate and in amounts directly proportional to the production rates. The optimum location for the injection is at the production manifold before the fluid enters the separator [52]. Uniform dispersion of chemical demulsifiers can be also achieved when the small volume of the demulsifier is mixed with a diluent prior to injecting it to the emulsion. Turbulence accelerates the diffusion of the additives in the emulsion and increases the number and the force of collisions between the water droplets. However, excessive turbulence can cause further emulsification so its duration and intensity must be controlled. The use of excessive chemicals not only increases the production costs for purchasing, handling and injecting the demulsifier but it can also contribute to the emulsion stability.

The combination of heat and chemical demulsifiers is the most common way to enhance oil dehydration. Chemical addition should be carefully designed in conjunction with heating so that the optimum separation can be achieved in the lower operating cost possible. Since chemical treatment is relatively cheaper and less energy demanding than heat treatment, the objective is the addition of appropriate volume of demulsifiers to minimize heating demands during treatment [109]. A major drawback of the demulsifiers' dehydration method is that the supply of chemicals must be continuous to the production sites. Constant market supply is sensitive to weather conditions, market availability and even political factors, so the cost is increased. This cost can turn into an economical challenge when the supply is at remote, Arctic or subsea fields.

Electrostatic treatment

Crude oil production industry has a long experience on using electrocoalescer packages for dehydrator and desalter vessels [115]. The final oil should contain less than 0.3% water-in-oil content and 10 PTB of total dissolved salts.

Electrocoalescence is an energy-efficient and cost-effective way to destabilize water-in-oil emulsions in heavy oils. It is also considered as enabling technology for subsea separation of produced water from heavy oil in deepwater developments [114] [117]. Electrostatic treatment can be used in conjunction with chemical and/or heat treatment and reduce the costs of chemical additives and energy consumption [118] [109].

The technique involves the use of electric field to separate the liquid phases. For that to happen, the thin film of the continuous liquid phase that surrounds the droplets must be reduced in thickness and break when the droplets approach one another. The electrostatic forces must be strong enough to outweigh the interfacial tension barrier and induce coalescence. Generally, the application of an electric field increases the coalescence rate, improves the migration speed of the droplets towards the electrodes and consequently facilitates separation [97].

The first publications over the use of electrocoalescence for phase separation appeared in the first third of the last century [97]. The contemporary research over electrocoalescence sought to make separation faster by affecting on the current type (AC or DC) and the frequency of the applied field, the physical and electrical properties of the fluids, the chain formation and the fluid motion. The physical and electrical properties of the fluids differ according to the field location and the production time [119].

Advances in electrocoalescing technology allowed its application in three-phase separators by using fully insulated electrodes. The dual frequency electrostatic treaters with composite electrodes, the horizontal flow electrostatic treaters and the compact electrostatic treaters represent new technologies that have achieved high separation efficiencies. Before analyzing these advances, some basic theoretical background on electrocoalescence is presented in the following paragraphs.

Electrical Forces

The application of an electric field imposes additional forces on the suspended droplets (besides gravity and drag force) which affect their motion and make them collide with sufficient velocity to overcome these barriers [106]. Depending on the charging and polarization of the water droplets and the distribution of the external electric field and assuming a fixed spherical shape for the drops, three different forces may act.

- **Dipolar attraction** between two induced dipoles. When a water droplet is exposed to an electric field it will respond by reorienting the dipolar water molecules so as to align with the electric field. This situation leads to the formation of an induced dipole [120] [113]. When two induced dipoles approach each other they will either attract or repel. For dipole-dipole interaction between two spherical droplets, the classical equation for the dipolar electrostatic force is the following [121]

$$F_{DIP} = \frac{24\pi\epsilon_c E^2 r_1^3 r_2^3}{s^4} \quad (4)$$

where ϵ_c is the dielectric constant of the continuous phase, E is the intensity of the applied electric field, r_1 and r_2 are the radii of the two droplets and s is the distance between the centers of the droplets. If we consider that $r_1 = r_2 = r$ in similar droplets, equation (4) indicates that:

$$F_{DIP} \propto (r^6, \epsilon_c, E^2, s^{-4}) \quad (5)$$

From Eq. 5 it is evident that the dipolar force is strongly dependent on the droplets size and the distance between them. If we consider a homogeneous water-in-oil emulsion the distance, s, between the centers of two similar droplets with the same radius r is inversely proportional to the dispersed water volume fraction X [122]. That can be described with equation (6):

$$s = r \left(\frac{4/3\pi}{X} \right)^{4/3} \quad (6)$$

Consequently, the dipolar force becomes less effective when the water content is reduced because the distance between the droplets is increased, while it gets more effective in high water dispersions where the droplets are close to each other [115]. The dipolar electrostatic force is illustrated in the figure below. The dipolar attraction is common to all types of electrostatic fields.

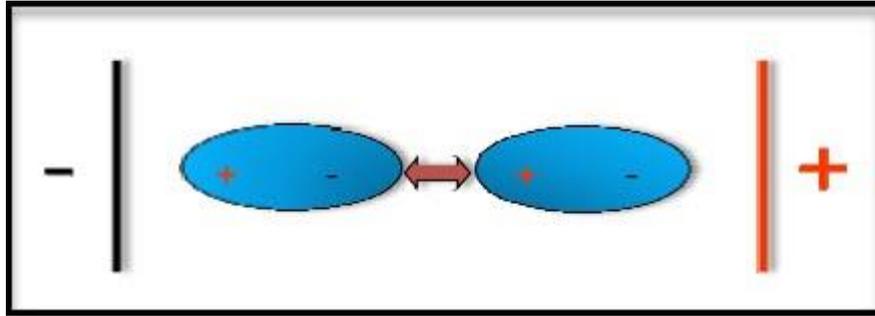


Figure 53: Dipolar electrostatic force produced by positive and negative centers on induced dipoles of water droplets, *Compact and Light Weight Equipment for Floating Production Systems*, Erik Sellman, Pavan Mandewalkar, Gary W. Sams, *Cameron Process and Compression Systems*, OTC-24727-MS, 2014, [104].

• **Electrophoresis** is an attractive or repulsive force between the charged electrodes and the charged water droplets in a uniform electric field. The electrophoretic force is given by the following equation

$$F_{EP} = qE = \frac{2}{3} \pi^3 r^2 \epsilon_c E^2 \quad (7)$$

The direction of the force depends on the polarity of the charge q and the direction of the electric field. In the equation 7, r is the radius of the droplet ϵ_c is the dielectric constant of the continuous phase and E is the intensity of the applied electric field. The intensity of the electrophoretic force is then:

$$F_{EP} \propto (r^2, \epsilon_c, E^2) \quad (8)$$

Unlike the dipolar forces, the electrophoretic ones are independent of the distance between the droplets [115].

If the field is a homogeneous AC field the droplets will oscillate around a mean position but no net movement will be observed. On the other hand, if the field is a homogeneous DC field, there will be a net movement of the droplets and finally a contact with an electrode [123]. However, in order for this charge not to leak away to the surrounding oil the relaxation time, τ of the continuous oil phase must be large compared to the droplet contact time with the electrode [113]. The relaxation time is given by the expression $\tau = (\epsilon_{oil} / \sigma_{oil})$, where ϵ_{oil} is the permittivity of the oil (dielectric constant) and σ_{oil} is its conductivity. In a uniform DC field the net movements of the droplets between the electrodes will provide them with many more collision opportunities, enhancing their coalescence significantly. It has been shown that the electrophoretic attraction force is the strongest electrical force available for moving the droplets (up to four orders of magnitude larger than the dipolar attraction force) but the corrosion mechanisms that emerge from the unchanging polarity and the tendency to short circuit (chain formation, see in the

following paragraphs) limits the use of high DC field in electrocoalescing technology [115] [106]. The electrophoretic force is illustrated in the figure below.

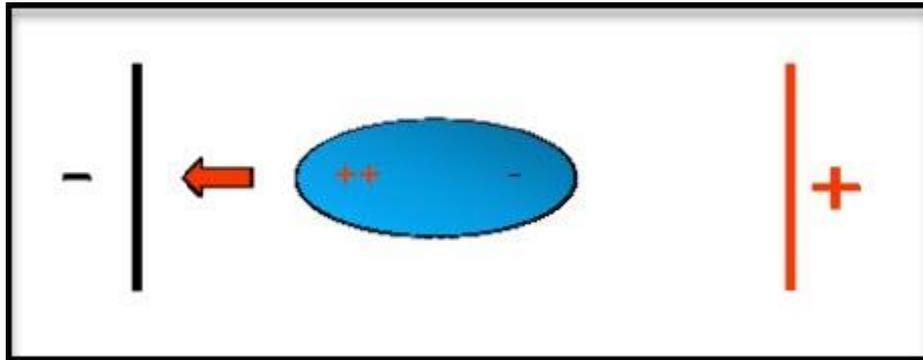


Figure 54: Electrophoretic force moves the charged water droplets towards the closest electrode, Compact and Light Weight Equipment for Floating Production Systems, Erik Sellman, Pavan Mandewalkar, Gary W. Sams, Cameron Process and Compression Systems, OTC-24727-MS, 2014, [104]

- **Dielectrophoresis** is the tendency of the droplets that have different dielectric constant than the continuous phase to move towards or away from the place with the highest field intensity [113]. This force arises only in non-uniform AC or DC fields. The direction of the force depends on the dielectric constant of the droplets and the continuous phase. If the dielectric constant of the droplets is higher than the dielectric constant of the oil, which is always the case, the force will pull the droplet towards the highest field intensity (positive dielectrophoresis) [123]. The dielectrophoretic force is given by the following equation:

$$F_{DP} = 2\pi r^3 \epsilon_c \beta \nabla (\vec{E} \vec{E}) = 2\pi r^3 \epsilon_c \beta \nabla |\vec{E}|^2 \quad (9)$$

where β is a factor defined as $\beta = \frac{\epsilon_d - \epsilon_c}{\epsilon_d + 2\epsilon_c}$, ϵ_d is the dielectric constant of the dispersed droplets and ϵ_c is the dielectric constant of the continuous phase. In a homogeneous field $\nabla |\vec{E}|^2 = 0$, therefore $\vec{F}_{DP} = 0$. The variables that determine the dielectrophoretic force are shown in the following equation:

$$F_{DP} \propto (r^3, \epsilon_c, \nabla E^2) \quad (10)$$

The dielectrophoretic force is independent of the inter-droplet distance, but it depends strongly on the gradient of the electric field and also on the droplet's electrical properties and size. Dielectrophoretic forces are proportional to the droplet radius raised to the power of three, so they may be ineffective to remove the smallest droplets, which constitute the residual fraction of water in gravity separated emulsions [115]. Dielectrophoretic force is the weakest of the aforementioned forces (around half the strength of the dipolar attraction force) and it is common in all field types. The dielectrophoretic force is illustrated in the following figure.

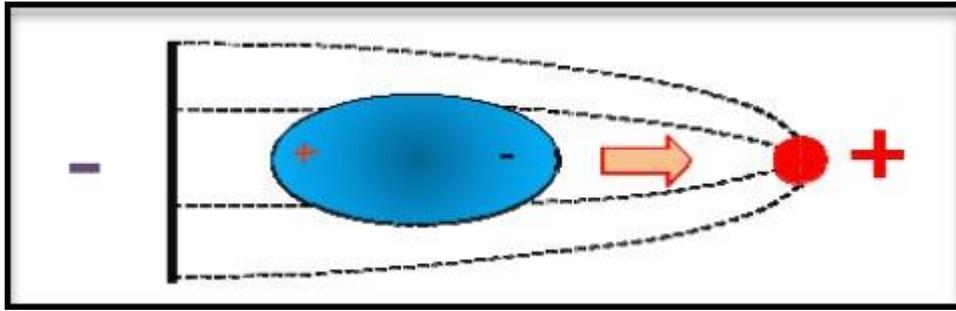


Figure 55: The dielectrophoretic force moves the polarized water droplets in a non-uniform field towards increasing gradient, *Compact and Light Weight Equipment for Floating Production Systems*, Erik Sellman, Pavan Mandewalkar, Gary W. Sams, *Cameron Process and Compression Systems*, OTC-24727-MS, 2014, [104]

• **Deformation** is induced from the droplets' polarization due to the opposite charge of the two sides of the water droplet. This force competes with the interfacial tension force that tends to keep the droplet in a spherical shape and eventually deforms and elongates the droplet in the direction of the electric field. As a result the droplet's shape turns to an ellipsoid of eccentricity e which is given in the following expression [121]

$$e^2 = \left(\frac{9\epsilon_c r E^2}{16\pi\lambda} \right) \quad (11)$$

In equation (11) the introduced variable λ represents the oil/water interfacial tension. The droplets can deform up to a critical axis ratio defined as the ratio between the critical length of an elongated droplet and its original diameter. At this point the droplets become unstable and shatter. The critical deformation ratio was found to be constant with the value of 1.9 and it is connected with a corresponding threshold field E_c expressed by the following equation:

$$E_c = 0.648 \sqrt{\frac{\lambda}{2\epsilon_{oil}r}} \quad (12)$$

Above the threshold field E_c the interface becomes unstable and dispersion takes place generating droplets of a much smaller radius [120]. Equation (16) indicates that only droplets with diameters larger than approximately 650 μ m can shatter. Even though droplets of this size are neither encountered in emulsions nor exhibit separation difficulties, the interface deformation must be taken into account. This is because the local increases of the electrostatic field can be high enough to cause significant deformation and destabilization of the droplets leading either to coalescence or to shattering.

Electrostatic fields

There are several ways by which the electrostatic fields can enhance coalescence. The types of electrostatic fields are the alternative current (AC), the direct current (DC) and combinations of them such as the Dual Polarity and the Dual Frequency technologies [108].

• **AC electrostatic field** has been widely used for dehydration during the last 75 years. The AC field is applied at 50 to 60Hz and causes the droplets to deform due to dipolar attraction forces. Coalescence is accelerated by the attraction between the oppositely charged ends of the water droplets. Due to the nature of the dipolar attraction force the treaters using AC fields are very effective for bulk water removal whereas their performance deteriorates when lower water content is encountered. That is because in low

water cuts the dispersed droplets are spaced further apart and thus, the dipolar attraction force weakens. This is also observed during dehydration process, as the water content decreases, the dipolar attraction force loses effectiveness in removing remnant water from the w/o emulsions. Coalescers using AC fields usually have horizontal electrodes which are made from steel rods and that is the reason why they are often called grids [102].

Typical examples of AC electrostatic treaters are TriVolt® and Bilectric® from Cameron Process and Compression Systems (currently acquired by Schlumberger). These two AC treaters are illustrated in the figure bellow

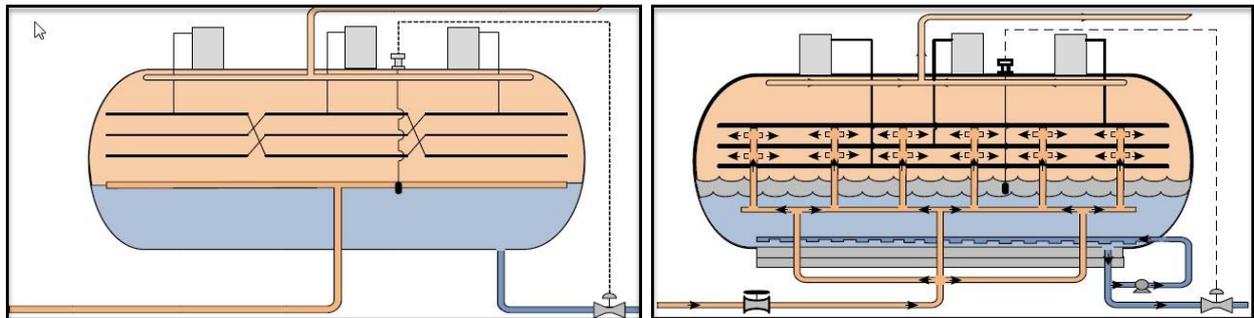


Figure 56: (left) TriVolt® and (right) Bilectric® AC treaters, *Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters*, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, Cameron Process Systems, SPE 156299, 2012, [106].

AC electrostatic treaters typically have upward vertical oil flow (up flow electrostatic treaters) so the fluids enter from the bottom of the treater and the oil outlet is on top. Common elements of all treaters are the inlet distribution system, the electrode grids, the power units, the collector pipe, the level control and the sand jetting/mad wash system. There are options for one, two or three grids in TriVolt® and Bilectric® AC treaters, that can be energized or grounded and each one of them has its own power unit. Bilectric treaters typically operate with 12.5 and 16.5 kVAC voltages and TriVolt have few more taps than the Bilectric. The inlet distributors can either be a conventional pipe or an open bottom distributor.

- **DC electrostatic fields** are capable to provide very effective coalescence due to the electrophoretic forces that increase the collision frequency of the droplets. However, the application of DC electrostatic fields in w/o emulsions with high water content results in electro-corrosion (galvanic corrosion currents). The risk of an electric breakdown is also high and in order to avoid chain formation (see next paragraphs) the electrodes are insulated. Even so the DC electric field causes all ions and charge carriers like water drops to move to the insulating barriers and the voltage drop between the electrodes is located mainly across the solid insulation. To face this effect pulsed DC and high frequency AC fields are used [97]. Pulsed DC field can energize droplet surfaces. Modulated High Frequency AC fields were developed by Cameron Process Systems and provide higher average electrostatic field inside the AC treater with significant dehydration improvement.

In conclusion, treaters that apply DC electrostatic fields are considered suitable only for processing refined products with very low water content [106].

- **AC/DC electrostatic fields** were initially used in combination with the Dual Polarity technology 45-years ago. The AC field is used for bulk oil removal and DC field for the removal of remnant water from the emulsion. In this process the incoming w/o emulsion is at first subjected to a weak AC field for bulk water removal and subsequently to a stronger DC field, where remnant water droplets are removed. Since

the DC field is applied only between the electrode plates the possibility of electro corrosion is minimized. The electrode plates are oriented vertically across the treater's length with alternate plates charged positive and negative (Figure 57).

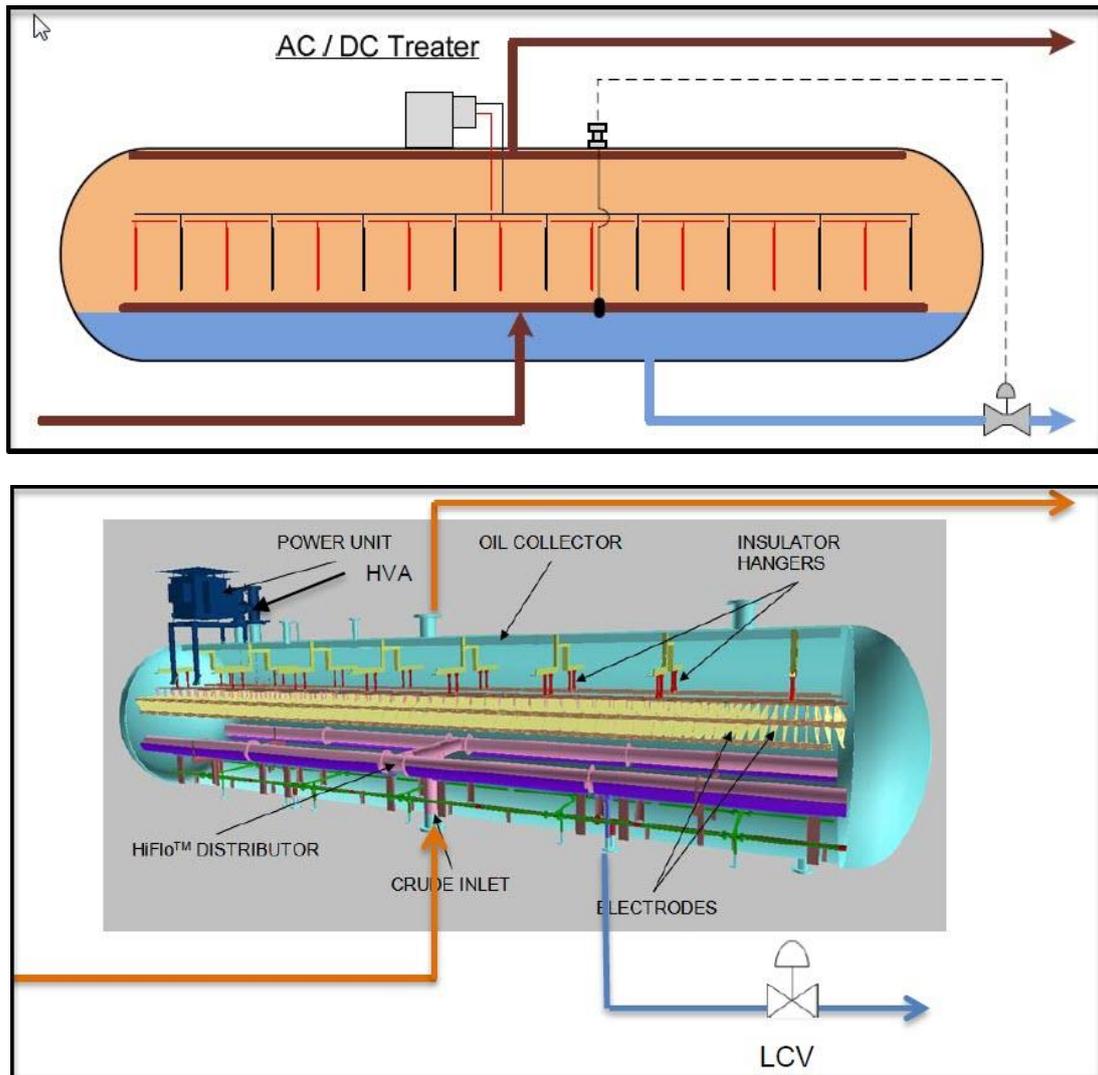


Figure 57: 2D (above) and 3D (below) schematic representation of an AC/DC electrostatic treater, (Above) Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, Cameron Process Systems, SPE 156299, 2012, [106] ,(below) 2-Prong Approach for Improved Operating Range of Electrostatic Crude Oil Treaters, Pavan Mandewalkar and Erik Sellman, Cameron Process Systems_OTC_26348, Brazil, 2015, [107]

The power supply is designed in such way that the positive and negative plates are not charged at the same time (half cycles) and thus, the possibility of sustained DC current is eliminated [108].

In the strong DC field the droplets are polished to lower BS&W specifications. They are charged and attracted to the electrode of the opposite polarity. When a droplet approaches the electrode of the opposite polarity is charged with the same polarity and moves towards the opposite electrode. The droplet moves in the DC field mainly due to electrophoretic force and then in the AC field is deformed due to dipolar force, then collides, coalesces and finally settles to the separated water phase in the bottom of the treater (Figure 58).

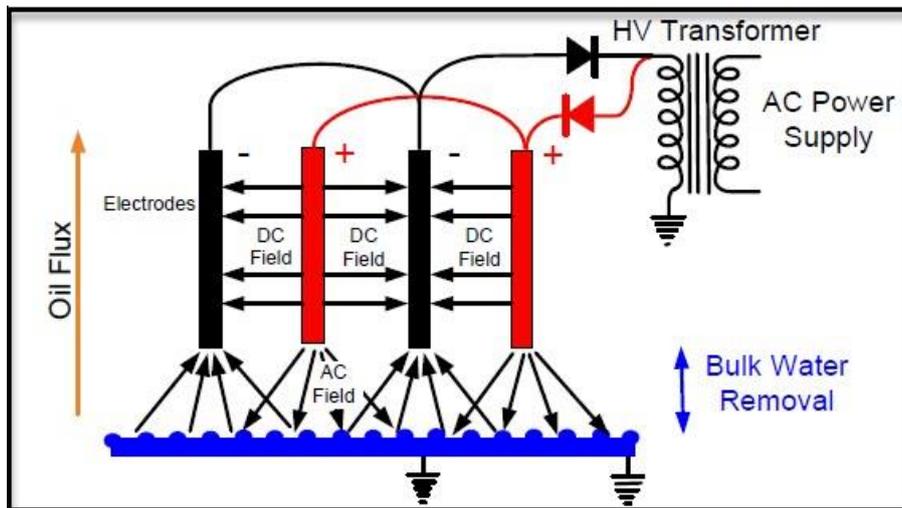


Figure 58: Typical Dual Polarity configuration utilizing AC and DC fields, Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, Cameron Process Systems, SPE 156299, 2012, [106].

The AC/DC treaters require fewer power and offer twice the voltage gradient compared to conventional AC treaters while they can handle 25% higher flux [107].

Crude oil Conductivity

Crude oil conductivity is very important as it affects the efficiency of the electrostatic fields inside the treater. When the oil conductivity is very low, the electrostatic charge cannot reach the droplets easily and when the conductivity is very high some charge dissipates into the oil (voltage decay) and does not affect the water droplets (Figure 59) [106].

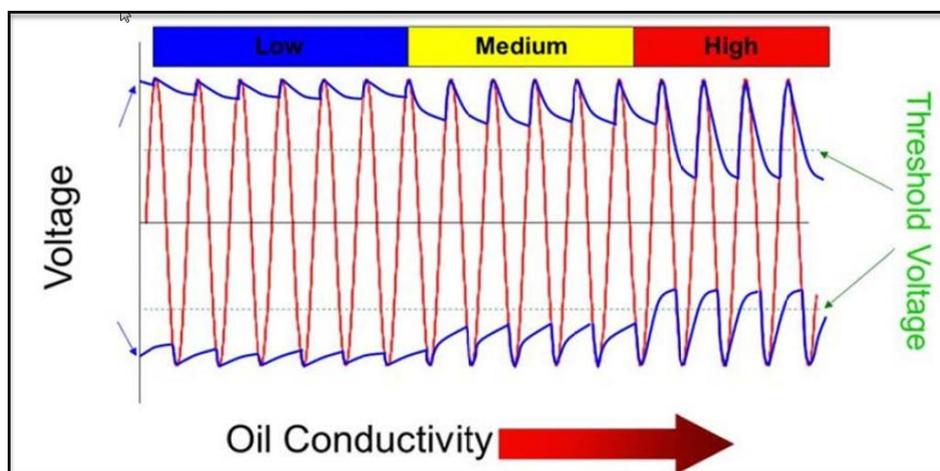


Figure 59: Voltage decay as a function of increasing oil conductivity, Field Implementation of New Electrostatic Treating Technology, John M. Walsh, Shell International, SPE, Gary Sams, Cameron, SPE, Joseph Lee, Cameron, SPE, OTC 23200, 2012, [108].

The conductivity is measured in NanoSiemens per meter. Crude oil conductivity can be divided in:

- Low, which is below 500 nS/m
- Medium, which is between 500 and 1000 nS/m
- High, which is between 1000 and 2500nS/m and

- Ultrahigh, which is above 2500nS/m

An important factor that affects crude oil conductivity is the process temperature. Crude oil conductivity increases up to 8 - 12 times as the temperature increases from the ambient one. The conductivity versus temperature effect is shown in the following figure:

Another factor that affects conductivity is the mixing of different crude oils. It has been observed that the crude oil blends have much higher conductivity than the original oil before blending [107]. The very heavy oils, with API gravity less than 14, are very often diluted, resulting in a mixture with a much higher conductivity than the original heavy crude.

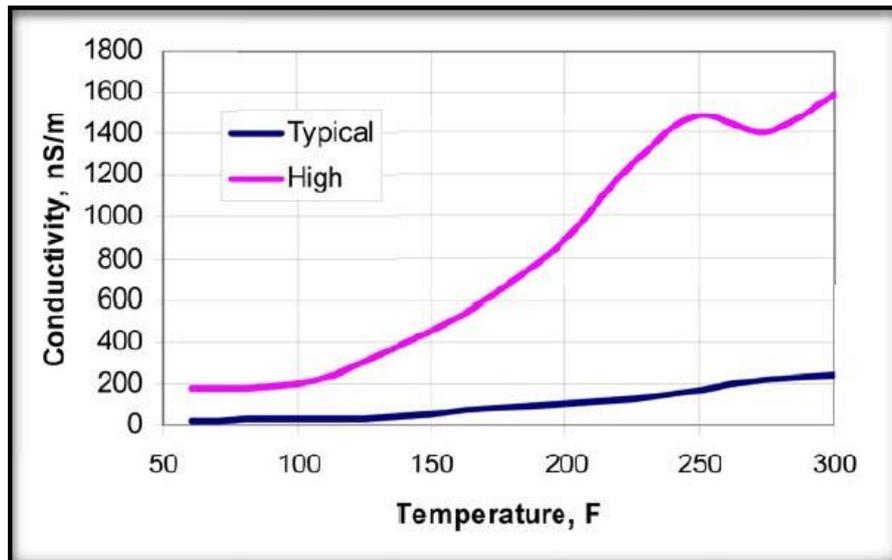


Figure 60: Conductivity as a function of temperature for a typical and high conductivity crude, *Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters*, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, *Cameron Process Systems*, SPE 156299, 2012, [106].

The result of high conductivity crude oil in an AC electrostatic treater is the low intensity electrostatic field and the low dehydration efficiency, since the smallest droplets cannot be reached from fields of lower voltage. The problem with the low intensity fields is that they do not have the energy needed to move the very small droplets. An impractical solution would be to use larger power units, but more often AC/DC treaters of modulated Dual Polarity are used, which are more effective in high conductivity crude oil. Modulated Dual Polarity is a 25 year-old technology in which the Dual Polarity treater can be provided with an additional voltage amplitude modulated power unit for improved water droplet coalescence and dehydration [124]. Voltage amplitude must be between a Threshold Voltage which is the minimum voltage necessary to initiate coalescence and Critical Voltage which is the limiting maximum voltage where a drop of specified diameter can exist (see previous Figure 60) [113]. The modulated Dual Polarity technology has higher coalescing power than the conventional AC/DC technology and is more effective on smaller droplets. It can be used in locations where external heat source is not provided because it can increase up to 50°F the operating temperature. It also exhibits tolerance to high water cuts and conductive oils. This technology has been used in crude oils that exhibited significant dehydration difficulties and the use of conventional dehydration technology was considered ineffective [107]. It has also been proved a very useful tool in cases where the properties of the water-in-crude oil emulsion changed during production. The fact that the treater does not work with a fixed Voltage but it can be modulated to match the specific properties of the inlet emulsion offers flexibility to the process equipment and can be optimized to the separation efficiency [113].

Chain formation

Chain formation has been observed mainly in DC fields and has been considered as one of the main retarding factors in electrocoalescence [97]. Basically, it is the formation of chains of droplets that connect to each other and align in the direction of the field creating a “bridge” between the electrodes leading to short circuit [121].

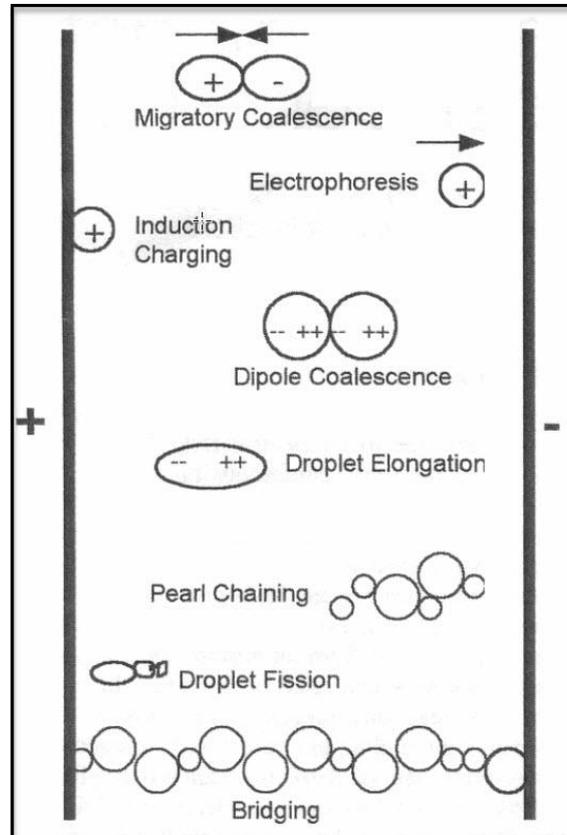


Figure 61: Different effects an electric field can have on droplets. At the bottom chain formation (bridging) is illustrated, Urdahl, O.; Nordstad, K.; Berry, P.; Wayth, N.; Williams, T.; Bailey, A.; Thew, M. *Spe Production & Facilities*, 16, (1), 4- 8, 2001, [125].

The chains may be formed without being in contact with one or both electrodes. The droplets of the chains do not coalesce because the film of the continuous phase between the adjacent droplets does not break. The existence of rigid films between adjacent droplets has been attributed to the existence of inherent surfactants of the crude oil such as asphaltenes [126] [127]. Electrode bridging results in a current leakage through the chains (short circuit) and, along with the continuous phase conductivity are the main factors influencing the power consumption of an electrostatic dehydrator. A method to prevent short circuiting is to insulate the electrodes. Insulated electrodes make sure that if bridging occurs between them the charge is only reduced locally on the electrode surface. On the other hand, with bare electrodes, bridging would fully discharge the capacitor and the whole electrostatic field would break down [128].

Fluid motion

In a w/o emulsion, even without the application of an electric field, there are random collisions between the droplets which depend on the flow conditions. Application of an AC field along with turbulent flow increases coalescence rate [120] and the median diameter of the dispersed phase is increased with moderate turbulent mixing even if no voltage is applied [129].

Before the 1960s all the commercial electrostatic treaters were designed in a way that maintained a degree of turbulence inside the vessel. The aim was to disrupt chains of droplets in order to prevent short circuit between the electrodes. However, the modern electrostatic coalescers are designed to maintain laminar flow conditions in the vessel so as to facilitate the simultaneous coalescence and gravity settling of the dispersed phase [113]. A compromise is necessary concerning turbulence intensity which must be high enough to prevent chain formation and low enough not to cause redispersion and breakage of water droplets [113].

DESALTING

Produced crude oil contains suspended and dissolved salts. Dissolved salts are responsible for corrosion, fouling, plugging and scaling which have detrimental effects on production facilities and refining processes [130]. Desalting is the treatment process that reduces the crude oil's salt content to be within the acceptable specifications and it is applied in electrostatic coalescer vessels. Since the salts are soluble in water a reduction in the water fraction will automatically reduce the crude's salt content. Desalting often takes place in two stages. The first stage is dehydration through gravity separators and the first electrostatic treater (dehydrator). The second stage is desalting and is performed in an electrostatic treater (desalter) which is placed right after the dehydrator. The salts are diluted by the addition of fresh wash water between the first coalescer (dehydrator) and the second coalescer (desalter). The main purpose of having a second electrostatic treater in the process train is to ensure a high dilution rate which is the ratio of wash water and water fraction in the crude. This dilution is essential in cases where dehydration alone is not sufficient to reach the typical salinity values in the export crude that range from approximately 10 to 100PTB.

Electrostatic treaters that are used as desalters treat the crude oil which is mixed with fresh water. This is shown in the following figure with a schematic representation of an Electro Dynamic Desalter (EDD) developed by Cameron Process Systems (currently acquired by Schlumberger).

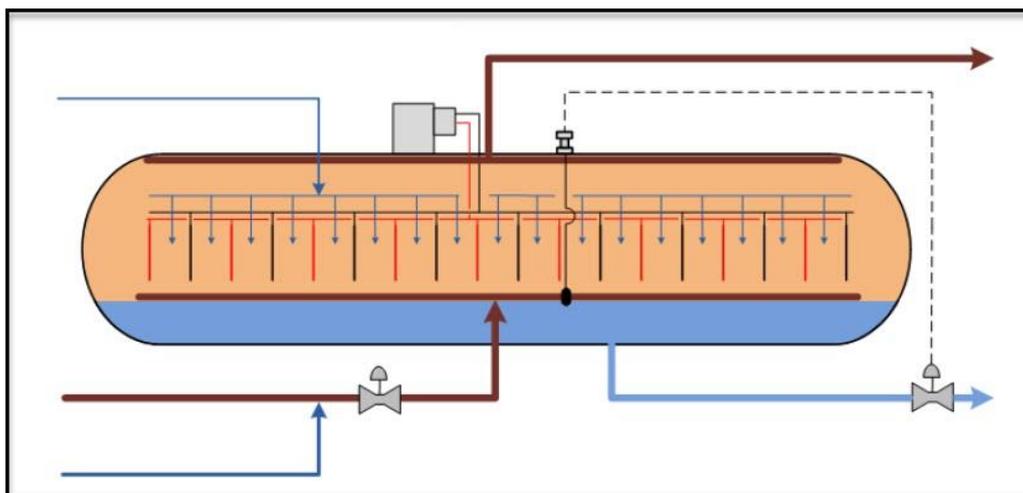


Figure 62: Schematic Representation of Electro Dynamic Desalter (EDD) from Cameron Process Systems, 2-Prong Approach for Improved Operating Range of Electrostatic Crude Oil Treaters, Pavan Mandewalkar and Erik Sellman, Cameron Process Systems, OTC-26348, 2015, [107].

The success of the desalting process depends on the efficiency of the fluid mixing. Conventional desalting process includes injection of wash water utilizing static mixers or mixing valves which cause pressure drop

and shearing. This fact may lead to the creation of undesirable stable emulsions of water and crude oil. An effective wash water-crude mixing operation must distribute the injected fresh water into uniformly sized and separable water droplets in the oil phase [130].

New technologies have been to market that ensure the optimal mixing of fresh water and crude oil, and guarantee gentle mixing and homogeneity of the fluid that enters the Desalter. One such device is shown in the figure below.

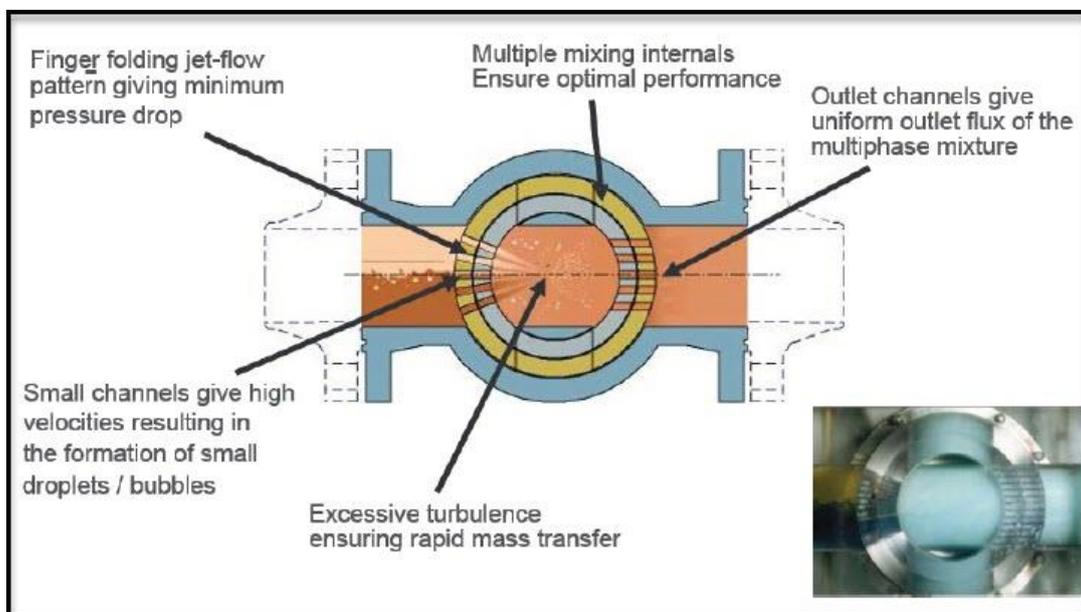


Figure 63: Schematic representation of ProSalt technology developed by ProSep Oil Technologies. ProSalt guarantees narrow water droplet size distribution and low pressure drop for high efficiency separation, Pro Sep Oil Technology Portfolio, 2015, [99], New Mixer Optimizes Crude Desalting Plant H. Linga, ProPure; F. A. Al-Qahtani, SPE and S. N. Al-Qahtani, SPE, Saudi Aramco, SPE 124823-PP, 2009, [130].

TECHNOLOGIES USED IN HEAVY OIL DEHYDRATION

HEAVY AND EXTRA HEAVY OIL DEHYDRATION WITH CENTRIFUGATION

In a usual oil dehydration process there are three main stages. In the first stage, most of the water is removed in a FWKO vessel. Then, in the second stage, the (containing emulsified water) “wet” crude oil is heated to facilitate separation in the second separation vessel, where the produced gas is flashed. The “dry” crude that exits the second stage separator is cooled before it enters the storage tank which is the third stage of the process. During this procedure the dehydrated oil meets the BS&W (Basic Solids and Water) specifications, which is essential in order to avoid pipeline corrosion and facilitate downstream upgrading processes.

As the oil density approaches that of water, conventional static gravitational separators become extremely large, heavy and require long retention times, excessive heating and more demulsifiers in order to break the stable water-in-oil emulsions and deliver crude oil that meets specifications. In heavy oil dehydration and desalting processes the final treatment stage normally takes place in an electrostatic treater after which the BS&W must be below 0.5%. In offshore facilities heavy oil dehydration is more demanding due to the restricted space and the weight limitations that require compact and lightweight

equipment. Even so and despite the heavy, large and expensive equipment heavy oil production may fail to meet BS&W specifications and thus deliver a product of lower value [131].

An alternative way to achieve heavy oil dehydration with reduced retention times, reduced vessel sizes and reduced or even without demulsifiers is the use of centrifugal separators (Centrifuges) [132].

Centrifuges have already been used in the huge oil sand's deposits in Canada, with API gravity as low as 8° and very high viscosity, in a process developed by Syncrude Canada Ltd. By using steam, hot water and caustic soda with the oil/bitumen they result in a slurry that is led to the primary separation vessels of the onshore facility. There it is treated with naphtha so that the bitumen floats on the surface as froth. At this point a highly efficient and compact solution is needed to turn the bitumen into refinery-standard crude oil. In cooperation with Alfa Laval a two stage separation process was developed for the treatment of bitumen froth using centrifugal separation technology. The schematic layout of the froth treatment is shown in the following figure

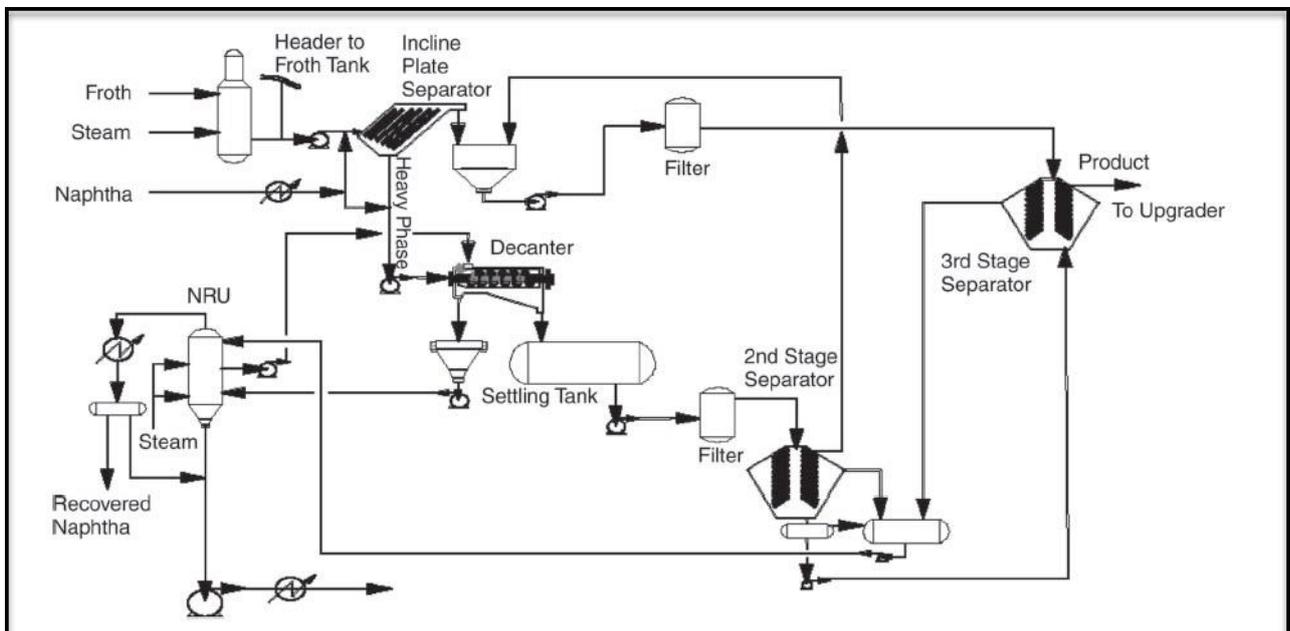


Figure 64: Schematic layout of the froth treatment process at Syncrude Canada Ltd., Heavy oil and bitumen dehydration-A comparison between Disk-Stack Centrifuges and Conventional separation technology, J. Agrell, Alfa Laval, Tumba A.B.;and M.Faucher, SPE, Alfa Laval Inc., SPE 97913, 2007, [131].

The centrifugal separators are used to treat a high throughput of bitumen froth under extremely erosive conditions in order to minimize the amount of water and solids and send a clean product to the upgrader. The feed in the first centrifuge consists of approximately 60% diluted bitumen, 30% of water and 10% of solids. The product leaving the second (polishing) stage contains less than 0.6% solids and less than 2% water.

In the same way as with the treatment of bitumen froth, disk-stack centrifuges are used in heavy oil treatment. In a disk-stack centrifuge oils with API gravity as low as 11.5° can be efficiently treated without excessive use of demulsifiers and meet BS&W specifications. The advantage of the disk-stack centrifuge is that it has a compact size and it can be used in offshore facilities where the available space and payload is at minimum. Heavy crude oil normally contains more solids than the lighter crudes, so the treating equipment must be able to treat high solid load. Intermittently discharging centrifuges are less suitable

for treating heavy crude oils due to their limited solids-handling capacity whereas the disk-stack centrifuge discharges continuously the solid load due to its specially designed nozzles.

In the disk-stack centrifuge the separation process takes place in a rotating bowl at 5000 – 8000g's force field. The internal components that consist of a stack of conical disks closely packed offer the needed area for the heavy crude to be continuously treated. The spaces between the disks form channels of approximately 5mm distance and can be regarded as a number of parallel separation vessels. These inserts act also as emulsion breakers [132]. Although the residence time inside the centrifuge is on the order of a few seconds the degree of separation is very high because of the centrifugal force.

The main parts of a disc-stack centrifugal separator are the frame, the drive motor, the separator bowl and the inlet/outlet arrangements. These parts can be seen in the following figure.

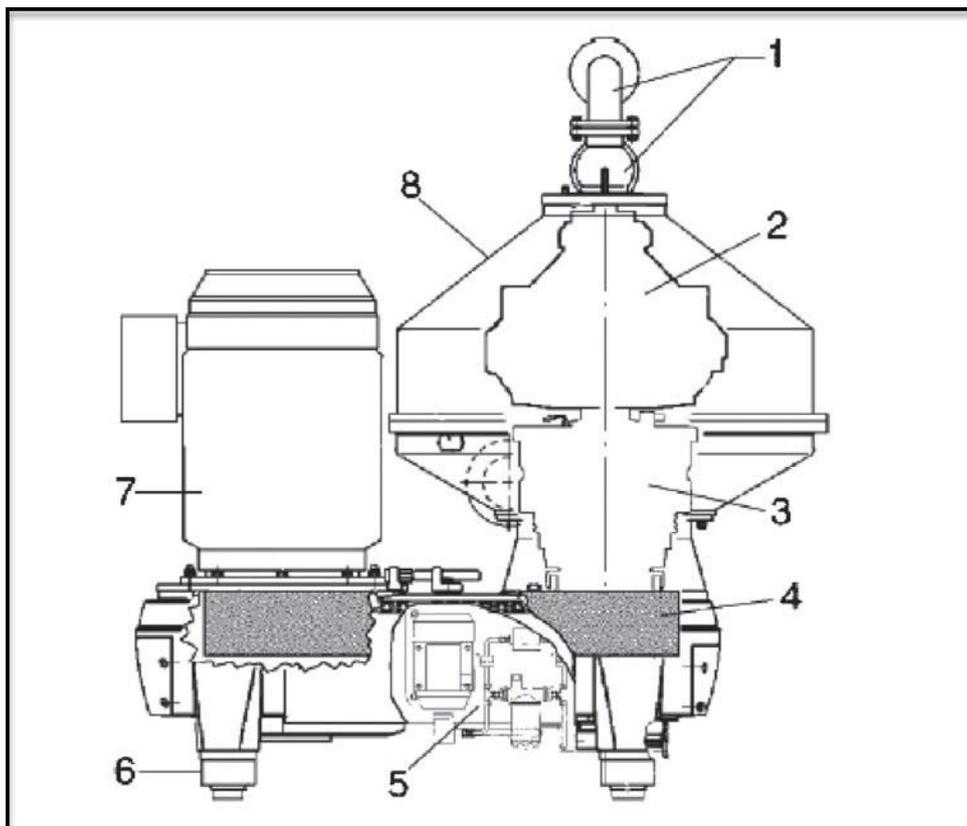


Figure 65: Main parts of the centrifugal disc-stack separator: (1) inlet and outlet, (2) bowl, (3) spindle, (4) belt and belt pulley, (5) oil circulation system, (6) foundation feet, (7) motor, (8) frame hood, Heavy oil and bitumen dehydration-A comparison between Disk-Stack Centrifuges and Conventional separation technology, J. Agrell, Alfa Laval, Tumba A.B.;and M.Faucher, SPE, Alfa Laval Inc., SPE 97913, 2007, [131].

A brief description of the liquid flow in the disk-stack centrifuge is as follows:

The liquid enters through the stationary inlet device and bushings into the rotating separator bowl accelerated to full rotational speed. The rectangular sections created from the disc spaces minimize shear. The feed is evenly distributed between the discs and due to the high g-force heavy particles and water separate outwards along the disc surfaces, towards the periphery of the bowl and are guided to the nozzles. Solids and water are discharged continuously towards the screen inside the frame hood and exit through an outlet in the bowl casing. The oil, free of water and solids exits through the paring disc into the outlet pipe.

A cross section of the disc-stack centrifuge and a detailed presentation of the parts and the distribution of the liquids are shown in the following figure.

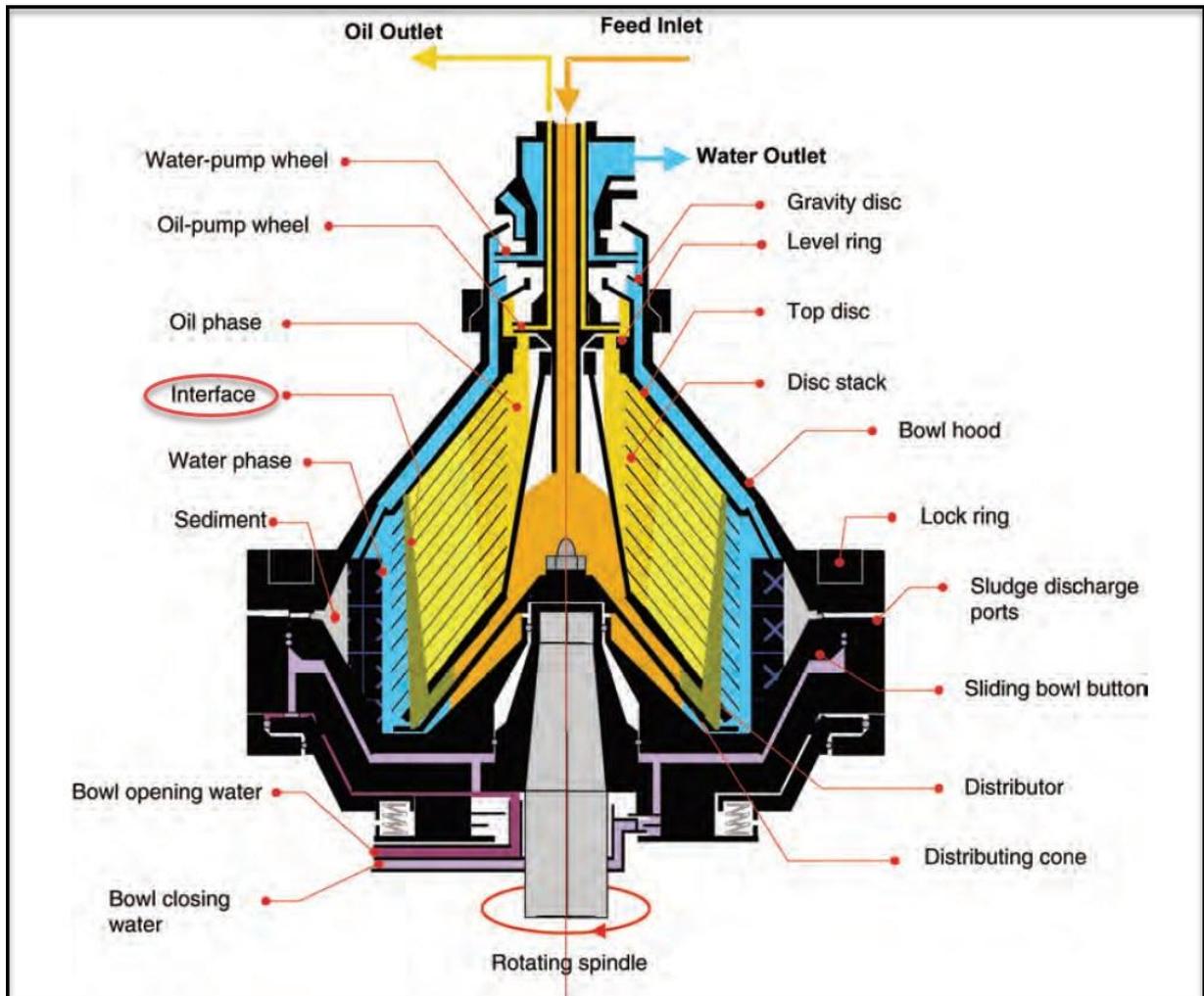


Figure 66: Cross section of the disc-stack centrifuge. The internal parts of the device are shown in detail, the interface close to the periphery of the bowl is denoted with a red circle, Utilization of heat and centrifuge technology to recover crude for sales from slop oil, Babita Angeli Dubay, Michael Mottley, Petroleum company of Trinidad and Tobago Ltd.; Ravi Ram, Kendall Mungal, Schlumberger Ltd, SPE-180906-MS, 2016, [133], (Courtesy of Alfa Laval Inc.)

The discs are designed in such a way that enables separation of water and solids from oil at high flow velocities. The result is removing efficiently particles and droplets of water of less than a few μm in size.

The design of the inlet section of the centrifuge is very important because shearing, droplet splitting and foaming must be avoided. For this reason the inlet is equipped with a stack of closely packed circular discs with the help of which the entering liquid can be gently accelerated with minimum splashing and foaming. The non-rotating liquid flows into the space between the inlet pipe and the circular discs and then moves outwards to the periphery of the discs and is set in rotation. The number of the discs depends on the inlet flow rate so for low flow rates a small number of discs is needed whereas for high flow rates a greater number. This configuration increases the efficiency of the separation because the droplets are not broken up and less gas becomes trapped in the liquid.

The efficiency of the separation process also depends on the position of the oil/water interface which must be maintained in a position that optimizes the usage of the available separation area. To achieve

this, the interface needs to be controlled by altering the pressure balance between the liquid phases in the bowl. That is accomplished by maintaining the level of the liquid in the paring chamber independently of flow rate. Two constant-pressure modulating valves connected with the water outlet, work in interplay with the paring device and keep the liquid level stable in the paring chamber.

The valves add or remove water through the paring device so that the nozzles are kept filled with water regardless of the water content of the feed. Consequently, when the feed contains less water than needed, the sufficient quantity of water is imported whereas an excess of water in the feed is exported [131].

Loop tests to estimate the efficiency of the disc-stack centrifuge device in heavy oil dehydration were conducted in 1998 using crude oils from the Mariner field prospects of 11.2° and 14.6° API [132]. The results of the tests showed that the disc-stack centrifuge was able to effectively dehydrate both crudes to BS&W specifications. Additionally the addition of chemical demulsifiers even improved separation in higher flow rates. A schematic layout of the heavy oil dehydration process using a disc-stack centrifuge is given in the following figure.

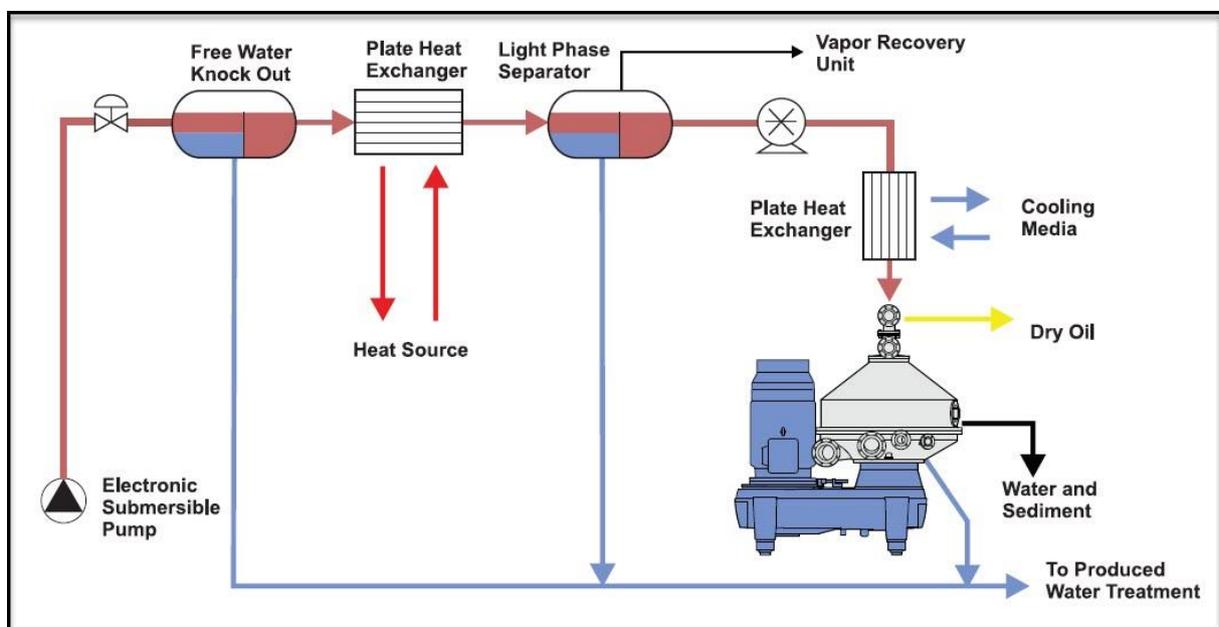


Figure 67: Schematic layout of the heavy oil dehydration process, Making heavy crude oil profitable, Alfa Laval

The use of disc-stack centrifuges in dehydration of heavy crude oils is considered as a reliable technology and a great number of centrifuges are installed in Offshore dehydration facilities of intermediate and heavy crude oils. The reliability in handling solids loading efficiently and resistance in wearing has already been proven in oil sands separation for the past 25 years. [131].

Besides oil dehydration the disc-stack centrifuge technology has found application in produced water deoiling as well. The only parameter that changes when the centrifuge is applied in the deoiling (removing oil droplets from water continuous phase) procedure is the oil-water interface position in the bowl. In **deoiling** processes, the interface is closer to the center of rotation whereas in **dehydration** processes the interface is closer to the periphery of the bowl (Figure 68).

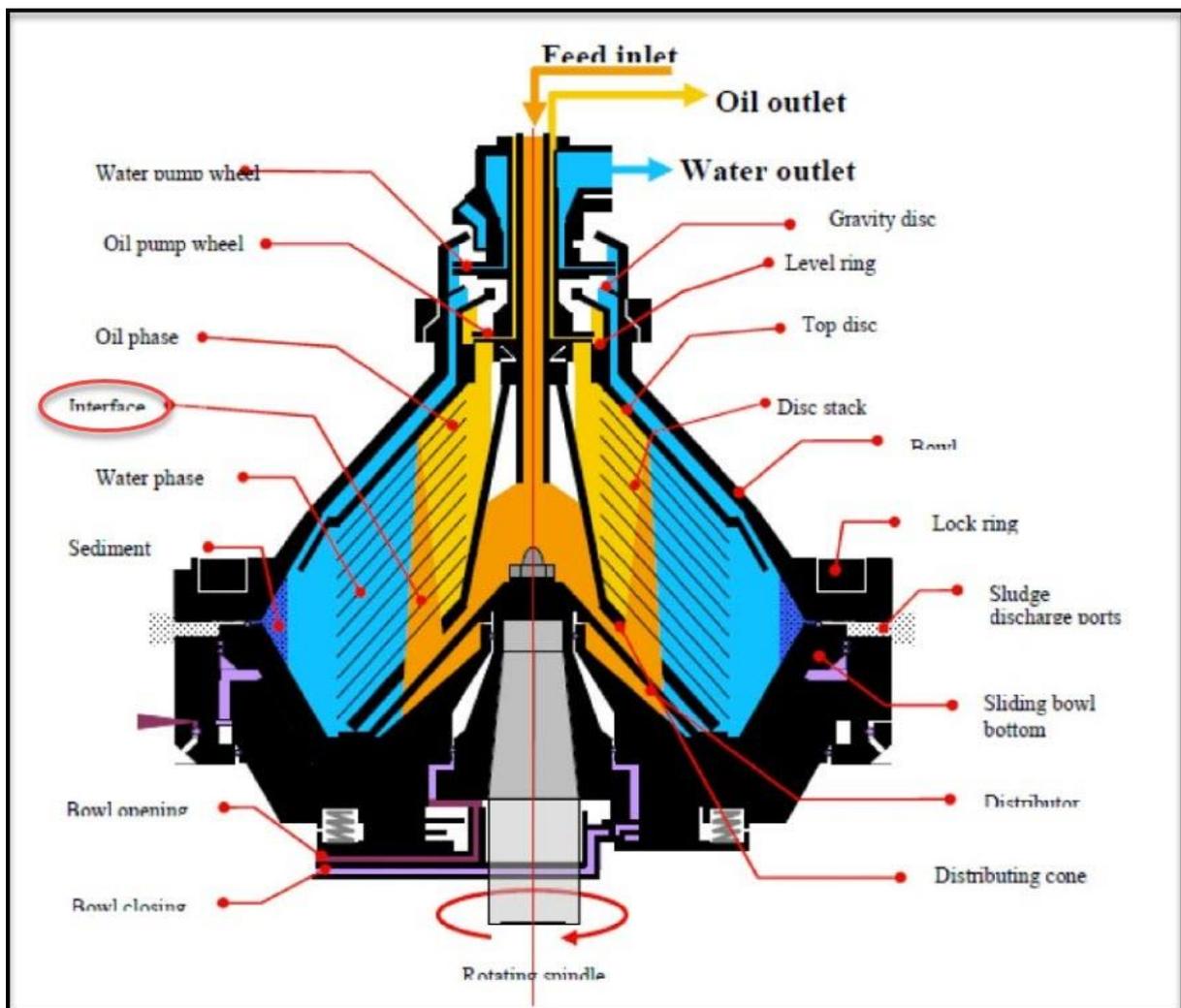


Figure 68: Disc-stack centrifuge in deoiling process. Compared to Figure 66 the interface is close to the center of rotation. Interface is denoted with a red circle, (Petroleum Engineering Handbook Volume 3, Facilities and Construction Engineering, Larry W. Lake, SPE, 2007), [47] (Courtesy of Alfa Laval Inc.)

Oil droplets of up to $1\mu\text{m}$ in diameter can be removed and the oil content in the produced water is in the range of 10 – 12ppm [132].

Disc-stack centrifugation is also applied in the recovery of oil from slop oil. Due to the higher contamination in water, solids and fines and the presence of very tight emulsions, the separation is usually enhanced by the application of heat. The disc-stack centrifuge is used in the last and most demanding stage of the procedure in order to break the emulsion and separate the oil from the water [134].

Current technologies deliver very versatile disc-stack centrifuges both for onshore and offshore applications. The specific equipment is used in almost all heavy crude oil processes such as In-situ recovery processes (e.g. SAGD - steam assisted gravity drainage), and mined oil sand production, as well as slop oil. The equipment can be modified or even change usage to adapt to the changes in feed composition during the production life of the field. So, for example, a centrifuge which was initially used in heavy oil processing can be configured later on to process the produced oily water [135]. [136].

CURRENT ELECTROSTATIC COALESCENCE TECHNOLOGIES

The electrostatic coalescence is the most effective demulsifying technology of water-in-oil emulsions and has been fully adopted in dehydration and desalting crude oil processes [124]. It results in the reduction in size and weight of the processing vessels as well as in heating demand and consequently in the carbon footprint of the processing facilities. [137].

The past few years there have been significant improvements in the capacity and performance of the electrostatic treaters through continuous improvements of their mechanism and the research over the electric field development. The development of novel in line and inlet electrostatic devices and their successful trials and installations in heavy oil processing plants have made electrostatic coalescence technology the most widely adopted solution for the demulsification of heavy oil [113].

Dual Frequency® AC/DC electrostatic treating technology

In order to treat high conductivity and low interfacial tension crude oils, Cameron Process Systems (currently acquired by Schlumberger) incorporated in the voltage-modulated combined AC/DC technology a variable frequency power supply. The objective was to control in an effective way the electrostatic field decay and the interfacial tension. In a conventional AC/DC treater when one set of plates is being charged the alternate experiences charge decay which, in high conductivity oils can result in the loss of the DC field. To counter this phenomenon the idea was the time between the charges to be reduced by increasing the frequency of the power source (Figure 69).

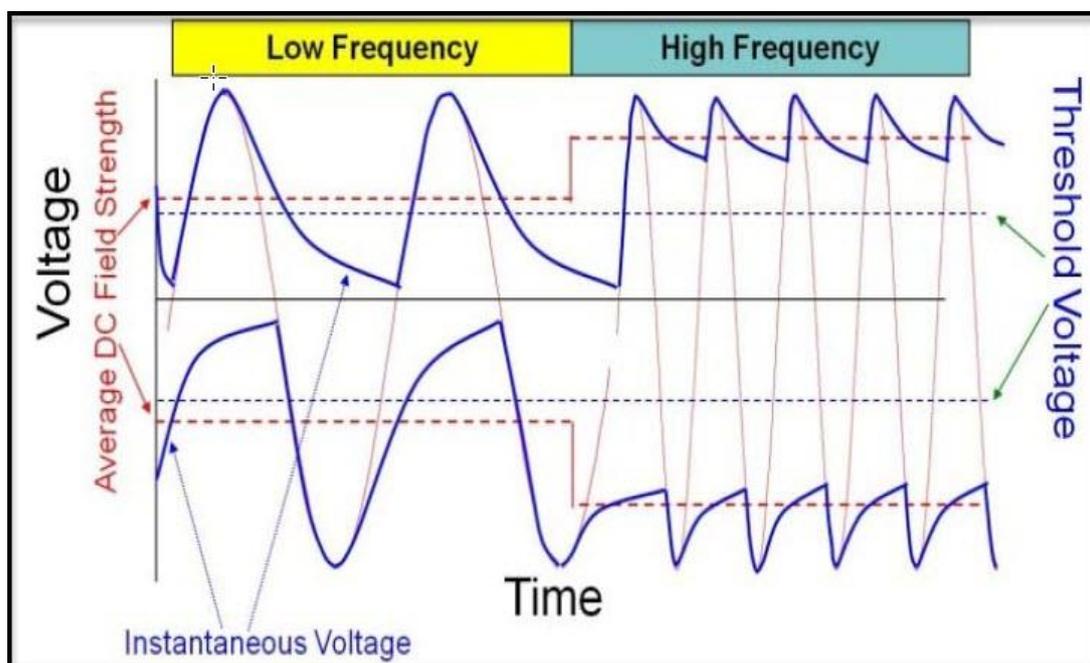


Figure 69: Average field strength increase with higher frequency, Compact and lightweight equipment for floating production systems, Erik Sellman, Pavan Mandewalkar, Gary W. Sams, Cameron Process and Compression Systems, OTC-24727-MS, 2014, [104].

As it is already mentioned, interfacial tension is an important parameter that controls coalescence. High interfacial tension between water and oil makes the droplets resist deformation and they don't coalesce while low interfacial tension allows the droplets to deform and coalesce easily but they are also broken

up as easily as they coalesced due to the low energy barrier. Typically, interfacial tension has been controlled within a desired range with the use of chemical demulsifiers.

With the application of an electric field the droplet surface must be energized sufficiently to overcome high interfacial tension. That can be accomplished by setting the power supply at a frequency close to the resonant frequency of the dispersed droplets. At the same time, the power supply frequency (base frequency) is set to a value high enough to minimize field decay and then is modulated (pulse frequency) at a rate that energizes the droplet surfaces. This combination creates sustained high field strength and droplets energized enough to coalesce easily. Similarly, for low interfacial tension, the voltage oscillation must be minimized in a frequency below the resonant frequency of the droplets. This way the droplets coalesce without the application of excessive electrical force that would lead to their dispersion [108]. The base frequency can be modulated in terms of amplitude as it is shown in Figure 70 and the curve of the pulse frequency modulation can vary [106].

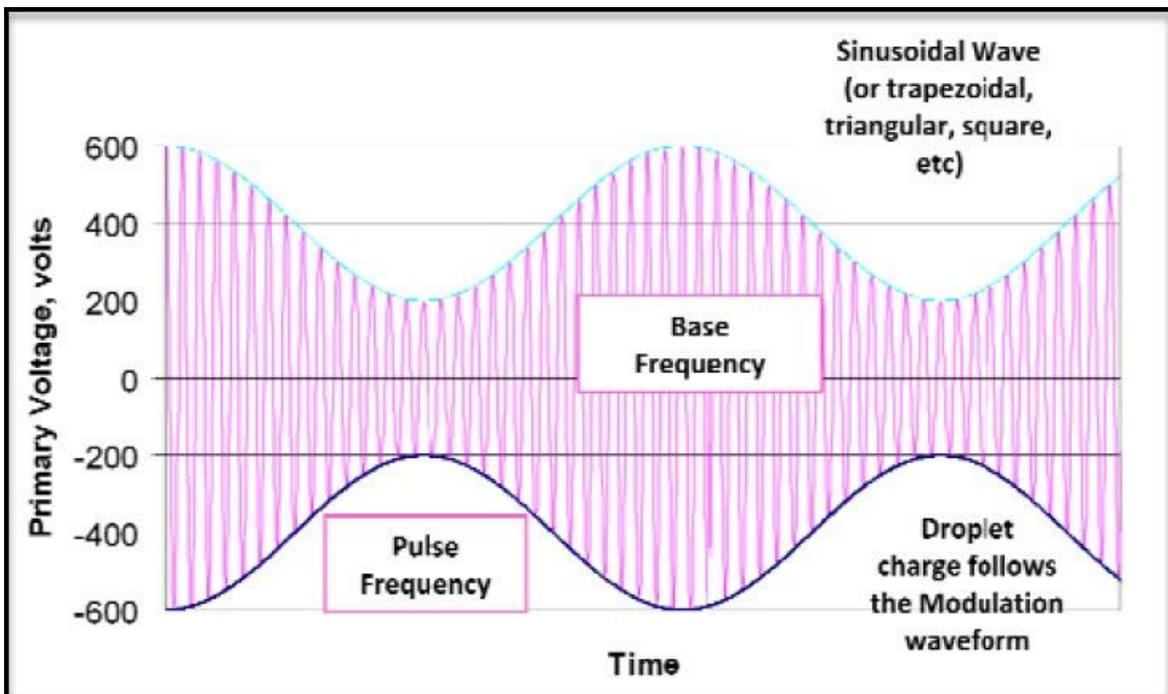


Figure 70: Frequency modulation in the Dual Frequency System, Compact and lightweight equipment for floating production systems, Erik Sellman, Pavan Mandewalkar, Gary W. Sams, Cameron Process and Compression Systems, OTC-24727-MS, 2014, [104].

This process of treating high oil conductivity and low interfacial tension with modulated voltage/frequency is called Dual Frequency crude dehydration and is applied over the last 12 years. The configuration of the electrodes is the same as the one in Dual Polarity technology (Figure 71) and it can increase the base frequency from 800 to 1600Hz to modulate the DC field according to the crude characteristics. A Dual Polarity unit can be upgraded to a Dual Frequency unit by changing the power unit and the electronic control system in order to adjust the base frequency and modulate the DC field.

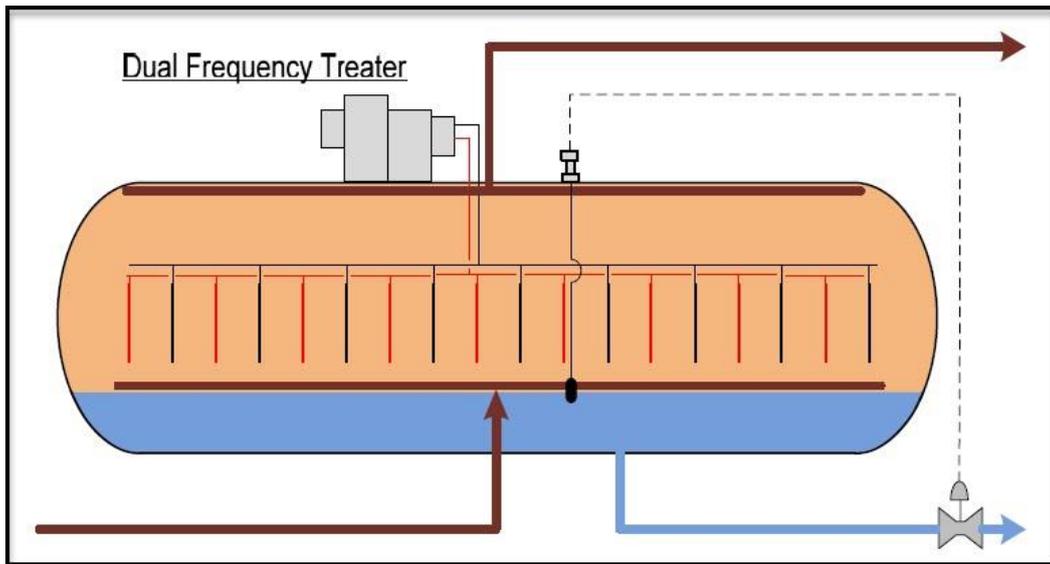


Figure 71: Dual Frequency combined AC/DC electrostatic treater that uses the same grid configuration with the conventional AC/DC treater, 2-Prong Approach for Enhanced Dehydration of Heavy Crude Oil and DilBit, Erik Sellman, SPE, Terry Murtagh, PE, Pavan Mandewalkar, SPE, Cameron Process Systems, World Heavy Oil Congress, Canada, 2015, [107].

It is known that a system (water droplet) in resonance absorbs the maximum energy when the modulated frequency is close to the resonant frequency. This energy causes the droplet to oscillate internally and heavily deform with respect to the initial spherical configuration. The deformed droplets interact and coalesce easily [106].

Dual Frequency technology incorporates high electrostatic field strength and coalescing power. The result is that Dual Frequency treaters are smaller in size than AC treaters or conventional AC/DC treaters and require less demulsifier dosage [104].

Other important technologies that are implemented alongside the Dual Polarity electrostatic treater include the composite plate electrodes, the Load Responsive Controller (LRC) II, the Electrostatic Susceptibility Tester (EST) and the HiFlo® distributor. Each one of these technologies contributes in the effectiveness of a Dual Frequency treater. Equally important is the separation efficiency of the FWKO placed upstream of the electrostatic treaters. Reduced performance of the FWKO may deliver more water to the downstream electrostatic treaters that are designed for lower inlet water load.

– Composite plate electrodes are used instead of electrically conductive materials because they prevent chain formation and enhance coalescence. During chain formation the bridging between the electrodes occurs at single point on the metal plate but the associated electric field collapses over the entire electrode area. The result is a large energy consumption and the interruption of the dehydration process. Furthermore, chain formation distorts the electric field and creates point charges with high electric field gradients that lead to droplet dispersion and limited droplet growth. Unlike AC/DC treaters that operate with steel electrodes, the composite electrodes were primarily developed to withstand high water cuts. The electrodes that consist of electrically conductive materials (ex carbon steel electrodes) diffuse the applied potential in the whole body of the electrode providing a uniform charge over the surface. However, very large electric field gradients are created at the edges of the electrodes which lead to droplet shattering and arcing in high water cuts [124].

The composite plate electrodes consist of a conductive region surrounded from a nonconductive region and can overcome the aforementioned problems effectively (Figure 72).



Figure 72: Array of composite electrodes. The conductive region in the middle is surrounded by the non conductive material, Compact and lightweight equipment for floating production systems, Erik Sellman, Pavan Mandewalkar, Gary W. Sams, Cameron Process and Compression Systems, OTC-24727-MS, 2014, [104].

Compared to conductive electrodes, composite electrodes have peak charge at the center and lower charge at the edges due to their high electrical resistance.

The electrical connection is made through the conductive material that distributes the charge along the electrode and the voltage at any point on the electrode is determined by the electrical current passing through that point. This way, bridging of the electrodes can be minimized. When the upward-moving emulsion enters the area between the electrodes, they draw more current and their voltage decreases. Further upward flow subjects the emulsion in an electric field of decreasing intensity from high to low. This produces a variable field gradient that adjust to the conductivity of the w/o emulsion. The configuration of the composite plate electrodes is demonstrated in the figure below.

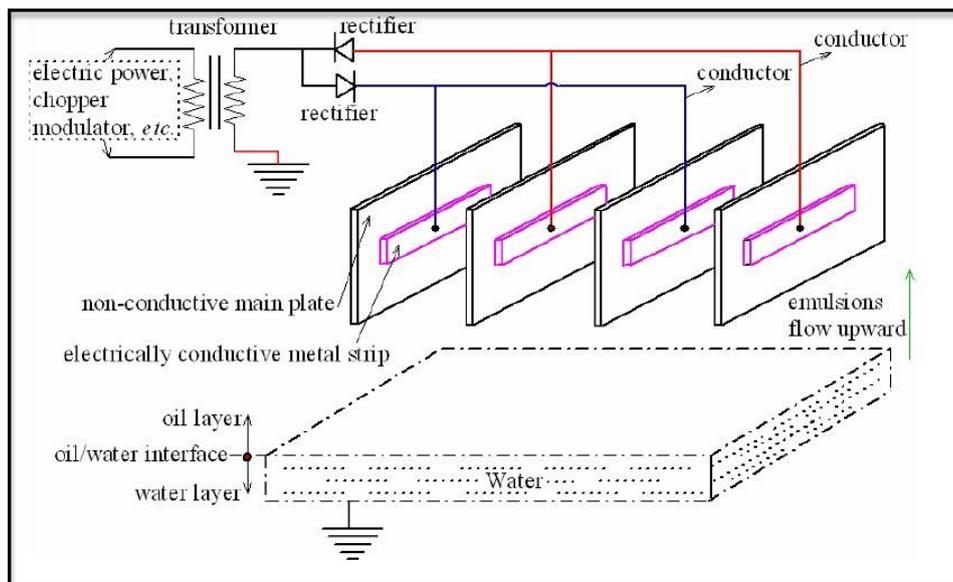


Figure 73: Original construction of composite plate electrodes, Distributed charge composition electrodes and desalting system, Prestidge F. L., Johnson B. C., United States Patent No. 4702815, 1987, [138].

Composite plate electrodes that are fabricated of fiber-epoxy material with included conductive material show great effectiveness. In addition to arcing (bridging) suppression, composite electrodes can tolerate longer exposure time in high voltage field, improved water droplet removal and they demonstrate improved handling of high water cut emulsions, even twice as high compared to treaters that use steel electrodes [139]. Comparison of dehydration efficiency between steel and composite plate electrodes shows an improvement in dehydration of approximately 56%. This can be seen in the following example in Table 10 for an API 20 crude oil with 15% inlet BS&W.

Flux ₂ (BPD/ft)	Inlet BS&W %	Outlet BS&W % Steel Electrodes	Outlet BS&W % Composite Electrodes	BS&W Reduction (%)
33	15%	1.2	0.40	0.82
53	15%	1.3	0.75	0.55
76	15%	1.2	0.60	0.60
107	15%	1.2	0.60	0.60

Table 10: Comparison of dehydration efficiency between steel electrodes to non-ferrous electrodes, API 20 crude oil with 15% inlet BS&W, *Improved Dehydration and Desalting of Mature Crude Oil Fields*, Erik Sellman, SPE, Gary Sams and S. Pavan Kumar Mandewalkar, SPE, Cameron Process Systems, SPE 164289, 2013, [103].

Composite plate electrodes are also used in counter-flow dilution during desalting. Conventionally dilution water is injected in the crude oil flow line upstream the desalting vessel. However, the electric field produced by the composite electrodes can be used for droplet breakup (mixing) and for droplet coalescence (settling). In order to achieve this, the fresh water is introduced directly from the top of the desalter and passes downwards and between the vertically positioned composite electrodes, while the oil emulsion is introduced below the plate electrodes and flows upwards. This counter flow procedure in combination with the applied electric field provide intense mixing of the dilution water and the emulsion and increase the contact area between the water droplets and oil [113].

– Load Responsive Controller (LRC) II Smart Interface was based on the patented technology of Prestidge and Schuetz [140] and developed by Cameron Process Systems for use with the Dual Frequency and Modulated Dual Polarity technologies. The objective was to optimize the performance of the electrostatic treaters. The LRC-II Smart Interface regulates the flow of electrical current by providing full power regulation even during process upset conditions. It consists of a reactance transformer combined with an electronic device that senses the variations of power in the coalescer. If short bursts of high power are detected they are interspersed with off periods because the electronic device senses the load and adjusts the power to the transformer accordingly. The result is that the power is delivered to the process even in upset conditions and without problems in the power supply. Such surveillance of power supply safeguards the integrity of the dehydration process. [113].

– The Electrostatic Susceptibility Tester (EST) is a tool used for the selection of the optimum dosage and the dosage rate of the demulsifier, in combination with the electrostatic field. EST works with 200ml samples of emulsion on which it applies various AC and AC/DC electrostatic fields. As it can be seen in figure 74 the samples are subjected in a voltage gradient which is imposed between two electrodes by the EST device.

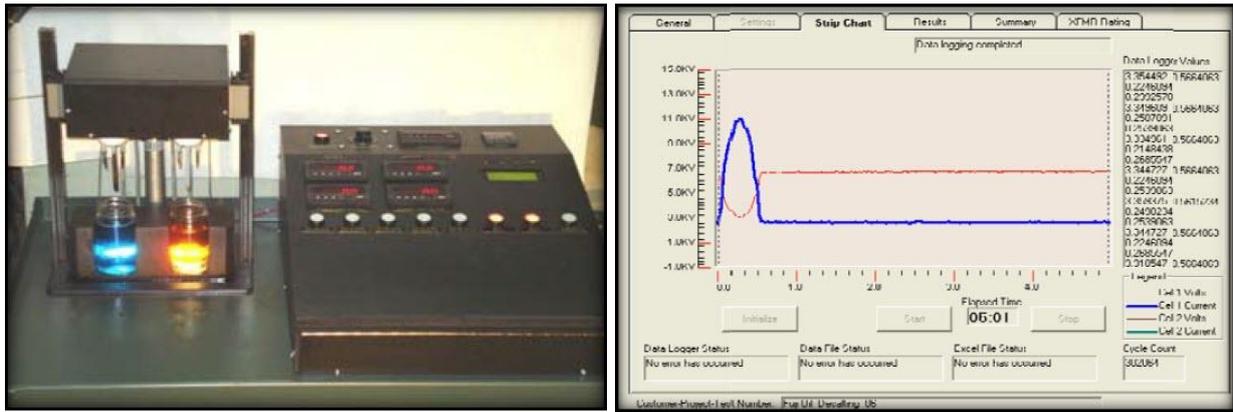


Figure 74: Electronic Susceptibility Tester and output, 2-Prong Approach for Enhanced Dehydration of Heavy Crude Oil and DilBit, Erik Sellman, SPE, Terry Murtagh, PE, Pavan Mandewalkar, SPE, Cameron Process Systems, World Heavy Oil Congress, Canada, 2015, [107].

The resolution of the emulsion as a function of time is plotted for all the applied electrostatic fields. By measuring the time required for the resolution of the emulsion in each case, the test engineer is able to recommend the appropriate demulsifier, the dosage and the dosage rate.

– Flow distribution inside the vessels is a very important parameter for the optimization of the dehydration process both inside the treater but also upstream, in the FWKO vessel. HiFlo Distributor is a proprietary spreader developed by Cameron Process Systems that contributes to the improvement of the treater efficiency. According to the manufacturer, the HiFlo distributor (spreader) can provide 35% better utilization of the vessel volume because it eliminates fluid recirculation and prevents fluid bypassing of the electrodes. With the use of CFD analysis an improved hydraulic design can be produced resulting in the improved utilization of the treater vessel and, consequently, longer effective residence times. FWKO vessels are facing increased challenge in removing small water droplets and sometimes the result is that the downstream electrostatic treaters receive larger water quantities than their designed values. The process trains for heavy oils usually include a FWKO vessel, diluent addition usually for crudes with API less than 14, a heater and an electrostatic treater. The inefficiency of the FWKO vessel arises from the high viscosity of the oil continuous phase. To counter this effect, Cameron Process Systems has included in the FWKO highly efficient internals to improve fluid distribution and enhance droplet coalescence. These consist of two perforated distribution baffles, one located right after the inlet section and the other right before the oil and water outlet. By doing this, an improved utilization of the vessel and an improved bulk water removal is achieved [107]. In the following figure, a schematic illustration of HiFlo spreader and FWKO internals is presented.

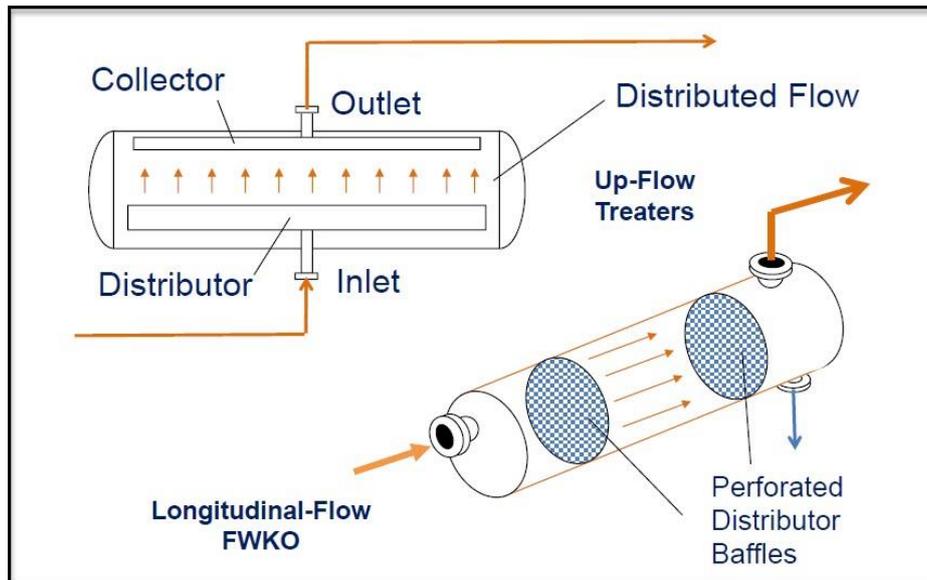


Figure 75: Flow distribution with HiFlo spreader and FWKO internals, Improved Dehydration of Heavy Oil and DilBit Using Modulated Electrostatic Treaters, Erik Sellman, Dan Senft, Terry Murtagh, Cameron Process Systems, World Heavy Oil Congress, 2016, [141].

Vessel Internal Electrostatic Coalescer (VIEC™)

Vessel Internal Electrostatic Coalescer (VIEC) is a technology that was originally developed by ABB Corporate Research Center in Norway in 1998 - 2001 and later by Vetco Aibel Technology & Products that acquired ABB Offshore Systems. Later, Hamworthy (a Wärtsilä Oil & Gas Systems AS company) has developed successfully VIEC technology. Since December 20, 2016, Sulzer has acquired VIEC technology from Wärtsilä Oil & Gas Systems AS. From its original development up to today, VIEC technology has been covered by worldwide patents [142].

Conventional electrostatic coalescers are made with bare metal electrodes. Consequently they cannot tolerate high water cut or increased gas content so they are placed after first and second stage gravity separators in a traditional gas/oil/water separation train (Figure 76).

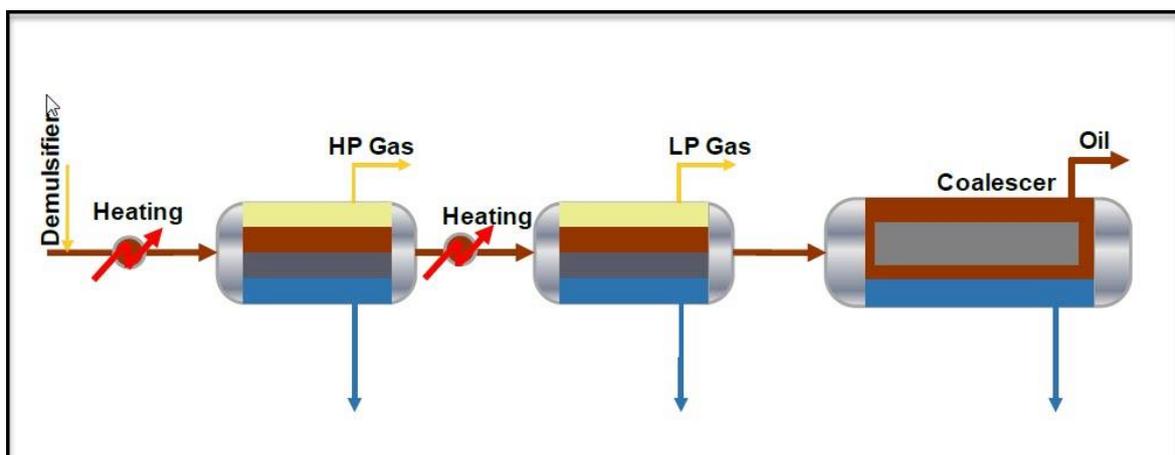


Figure 76: Typical separation train for processing crude oil, Efficient and Flexible Crude Oil processing technology for Oil Separation facilities, MORAD AMARZGUIOUI & PER CHRISTIAN JACOBSEN, Wärtsilä Oil & Gas Systems AS, 2014, [143].

Normally, most of the free water is removed in the first stage separator and gas is removed in both stages. However, even if the gravity separators are successful in removing all the free water, significant amounts of emulsified water can pass to the electrostatic treater, especially in the case of heavy crude oils. It is noteworthy that heavy crude oils can contain emulsified water up to 40 – 50%. This percentage exceeds significantly the designed limit of the electrostatic coalescer and there is a serious risk that the separation scheme will fail in meeting the required specifications. [143]. [144].

VIEC is claimed to be the first electrocoalescence technology that can be applied in the first stage separation, or as an inlet device of a three-phase separator and also work as a flow straightener [113]. The high voltage electrodes are subjected to the harsh environment of the first stage separator, where the water content can reach up to 90% and significant quantities of gas are released [145]. Just like most of the commercial electrocoalescer devices nowadays, VIEC™ includes insulated electrodes to prevent short-circuit between them [115].

VIEC™ combines the characteristics of a perforated distribution baffle and an electrocoalescer. The individual electrocoalescer elements have the form of tubular channels and the emulsion flows through them. The two electrodes extend helically and continuously over the outer periphery of each one of the tubular elements and apply an electric field to the emulsion. Each tube can be characterized as a small individual separator in which the droplets settle as a liquid film in the inner walls. The flow inside the tubular channels is turbulent and the residence time can be adjusted between one and three seconds depending on the fluid properties [113]. A stack of coalescing elements comprises the coalescer module. The size and weight of a VIEC module is 15x45x45 cm and 33kg respectively [115]. A number of modules are located in a supporting wall inside the separator. The entire assembly is put in a mold and cast into a solid block of epoxy under vacuum conditions [145]. In the next figure a VIEC module and a close up of the tubular channels is shown.

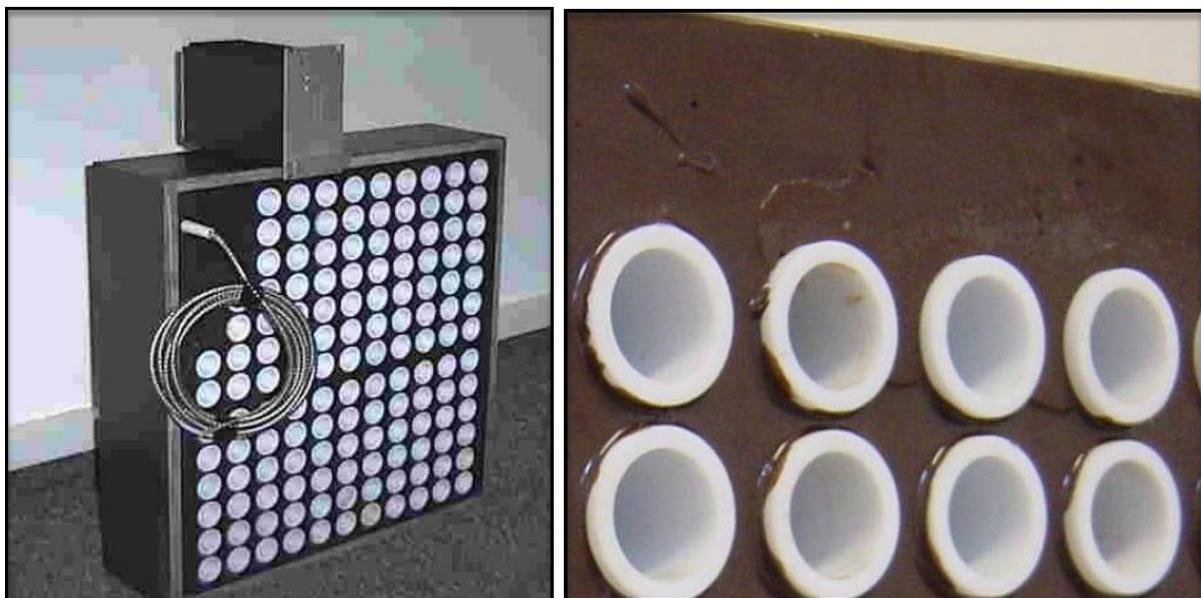


Figure 77: VIEC module (left) and close up (right), Advanced Electrostatic Internals in the 1st Stage Separator Enhance Oil/Water Separation and Reduce Chemical Consumption on the Troll C Platform, Erik A. Wolff/ABB Offshore Systems, Trond L. Knutsen/Norsk Hydro, Wojciech Piasecki/ABB Corporate Research, Peder Hansson/ABB Offshore Systems, Pål J Nilsen/ABB Offshore Systems, OTC 16321, 2004, [145].

Each one of the modules in this configuration is provided with a low voltage connector, a high voltage transformer (transforms the low voltage feed from the connector to high voltage, [146] and it is connected to its own individual frequency converter. The converter, which is controlled from an operating room, shifts the operating frequency in the kHz region to prevent high voltage drop across the insulated electrodes. Additionally, the converter allows frequency and amplitude adjustments to the optimal operating conditions. The power consumption of each module is typically 250W and the operating temperature up to 90° C. The turbulent flow through the VIEC flow channels increases the collision frequency of the droplets and consequently enhances coalescence. The VIEC wall is designed as a perforated wall, with an open area from 20 to 24% of the total VIEC, which reduces the sloshing effect of wave motion. This feature makes the VIEC suitable for offshore, floating facilities [113]. Above the VIEC assembly, perforated baffles are installed. Besides the individual tubular channels, other kinds of flow channels have been developed like the ones used by Wärtsilä Oil & Gas Systems AS and are shown in the following figure. The configuration with slotted flow channels (rectangular) is more efficient in clogging prevention [147].



Figure 78: VIEC wall installed in 1st stage separator with tubular flow channels (left), Advanced Electrostatic Internals in the 1st Stage Separator Enhance Oil/Water Separation and Reduce Chemical Consumption on the Troll C Platform, Erik A. Wolff/ABB Offshore Systems, Trond L. Knutsen/Norsk Hydro, Wojciech Piasecki/ABB Corporate Research, Peder Hansson/ABB Offshore Systems, Pål J Nilsen/ABB Offshore Systems, OTC 16321, 2004, [145], VIEC installed in a separator with rectangular flow channels (right), image source: <http://www.hamworthy.com>, Electrostatic phase separation: A review, S. Mhatre, V. Vivacqua, M. Ghadiri, A.M. Abdullah, M.J. Al-Marri, A. Hassanpour, B. Hewakandamby, B. Azzopardi, B. Kermani, Chemical Engineering Research and Design, 2015, [97]. In both cases perforated baffles are placed above and below the VIEC electrocoalescer.

The location of VIEC electrocoalescence system inside the separator is shown in the following cross sectional schematic representation. It is obvious that the emulsion layer is forced to flow through the VIEC. The standard design for VIEC is to cover the entire oil/emulsion phase.

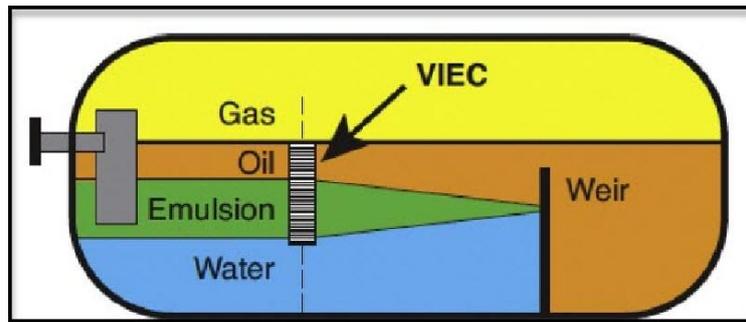


Figure 79: Schematic cross-sectional representation of the liquid phases around the VIEC electrocoalescer. The fluid flow is from the left to the right, *Electrostatic phase separation: A review*, S. Mhatre, V. Vivacqua, M. Ghadiri, A.M. Abdullah, M.J. Al-Marri, A. Hassanpour, B. Hewakandamby, B. Azzopardi, B. Kermani, *Chemical Engineering Research and Design*, 2015, [97].

The whole VIEC system includes the electrode modules inside the separator, the electrical connectors through the separator wall and the operating room which is located in a safe area where the cables are terminated.

Vetco Aibel AS (later Hamworthy Technology and Products AS) has also developed two more VIEC products, the **High Temperature VIEC (HT VIEC)** and the **Low Water Content Coalescer (VIEC LW or LOWACC)**.

The **High Temperature VIEC (HT VIEC)** has a similar design to VIEC but it can operate up to 150°C and 150 bar. The perforated plate consists of ceramic instead of epoxy material. It has been developed to enhance processing of fields that operate in high temperatures and experience separation difficulties. This makes HT VIEC suitable for the treatment of heavy and viscous crude oils. The first use for the HT VIEC electrocoalescer equipment was onboard the FPSO Petrojarl Cidade de Rio das Ostra for the Badejo development in Brazil's deepwater Campos Basin in 2008. The HT VIEC have been used with separators for processing the 12°API fluids of Siri reservoir within the Padejo field [148]. The pilot installation was in July 2009 and the start up operations in 2010 [147].

The **Low Water Content Coalescer (LOWACC or VIEC LW)** from Hamworthy Technology and Products AS) was developed by Vetco Aibel (former ABB Offshore Systems) to operate downstream the VIEC or HT VIEC inside high and low pressure vessels, and polish the crude oil to export specifications. VIEC LW applies dielectrophoretic forces to the mixture of oil and water [144]. These forces imposed by the non-uniform field gradient are 2 – 5 times higher than the gravity force. This way the droplets are guided to predetermined sections of the separator with the highest field strength and they are forced to merge [113]. While VIEC performs the bulk separation of water, VIEC LW complements the process by removing the remaining water down to export oil quality of 0.5% BS&W. VIEC LW in conjunction with VIEC can be applied in cases where there are intense emulsion separation problems, when there is serious possibility for hydrate formation and also in subsea separation processes because they reduce the water content in oil early enough in the separation train. [144]. According to Hamworthy, the VIEC LW insulated electrodes eliminate the risk of short circuiting and reach BS&W of export quality (0.5%) after the first and/or second stage separation without the need of traditional electrostatic treaters. Moreover, VIEC LW can reduce demulsifier's dosage and rate, increase liquid production, reduce the need for fluid heating, increase the ability to separate heavy oil, improve level control and reduce weight and footprint of topside separation equipment [149] [150]. VIEC and VIEC LW electrostatic internals enable one-step separation and heavy oil

separation and both can be used in high and low pressure vessels with separation difficulties or heavy emulsions. [149].

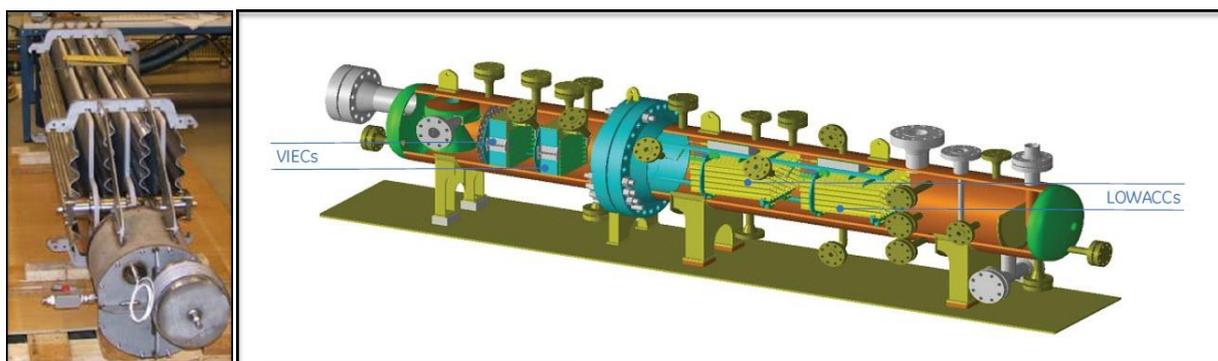


Figure 80: LOWACC electrostatic coalescer element (left), 3D CAD model of test separator with 2 VIEC and 2 LOWACC electrostatic coalescers at Vetco Aibel 2004, *Novel Coalescer Technology in First-Stage Separator Enables Single Stage Separation and Heavy Oil Separation*, T. A. Fjeldly, E. B. Hansen, P. J. Nilsen, Aibel, SPE 118891, 2008, [144].

A new version of VIEC was presented by Wärtsilä Oil & Gas Systems AS in 2014. In this case the VIEC assembly consists of single vertical electrodes cast in a special insulating material as it can be clearly seen in figure 81

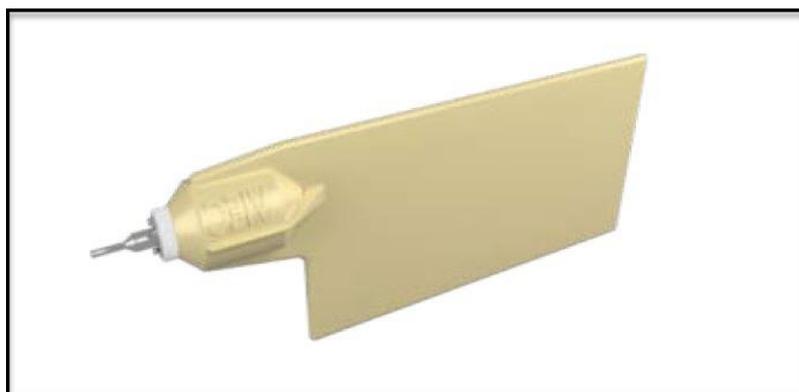


Figure 81: A single VIEC element that consist of an active electrode and an internal transformer inside a single continuous mold, *Overcoming Separation Challenges by Use of Electrocoalescence Technology*, Morad Anarzuguioui and Per Christian Jacobsen, Wärtsilä Oil & Gas Systems AS, SPE-171483-MS, 2014, [151].

The main mechanisms to prevent arcing are the insulated electrodes and the fact that the high voltage is confined in the volume of each electrode and the conversion from low (220V) to high voltages is conducted by an internal transformer. In this version of VIEC the modules differ from the ones already presented, since the vertically-placed VIEC elements are installed in one or several walls of modules, inside the production three-phase separator (Figure 82). The modules are placed in the form of “cassettes” and the elements are installed in a height appropriate for treating the water-in-oil emulsion.

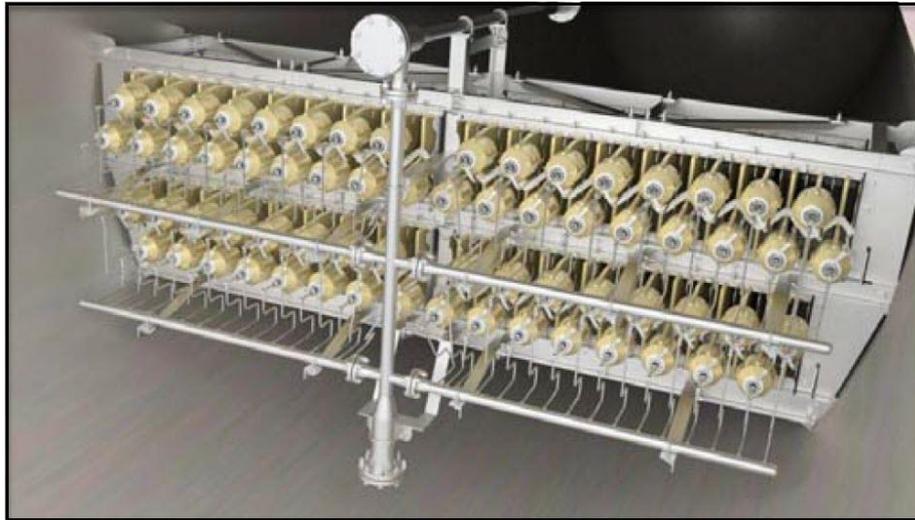


Figure 82: VIEC wall inside a separator, Efficient and Flexible Crude Oil processing technology for Oil Separation facilities, MORAD AMARZGUIOUI & PER CHRISTIAN JACOBSEN, Wärtsilä Oil & Gas Systems AS, Norwegian oil and gas partners, 2014, [143].

A VIEC installation in a three-phase production separator is shown in the following figure. VIEC can be also installed as retrofit to an existing three-phase separator.

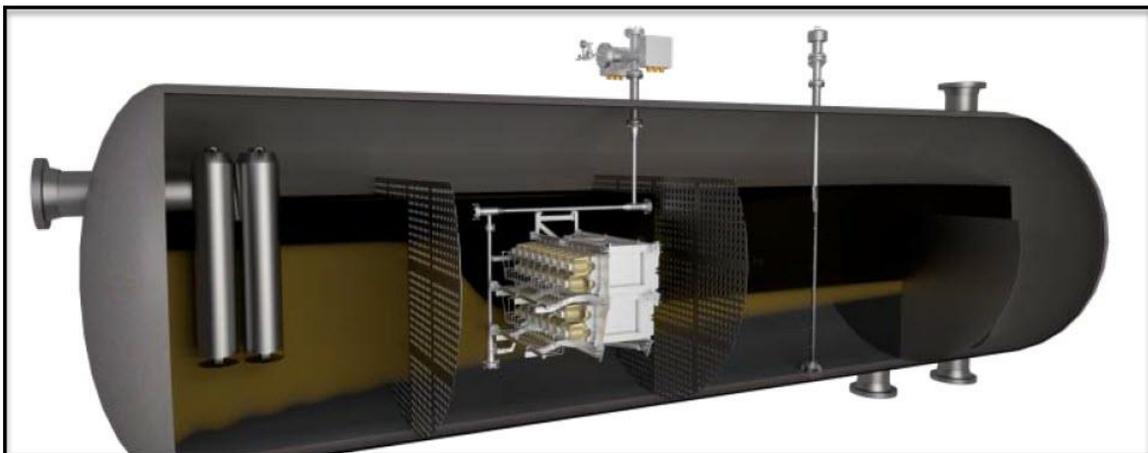


Figure 83: Three-phase separator with VIEC system installed, Efficient and Flexible Crude Oil processing technology for Oil Separation facilities, MORAD AMARZGUIOUI & PER CHRISTIAN JACOBSEN, Wärtsilä Oil & Gas Systems AS, Norwegian oil and gas partners, 2014, [143].

A conventional separation train includes a first and a second stage 3-phase gravity separator. The outlet from the second stage separator is introduced to one stage (dehydration) or two stages (dehydration and desalting) scheme of conventional electrocoalescers. The choice between one or two coalescing stages depends on the BS&W and salinity specifications [151]. Conventional electrocoalescers have very strict limitation regarding the water content they can handle, which is typically between 15 and 20%. If the first two stages of gravity separation fail to remove the designed quantity of emulsified water, then the electrocoalescers will also fail to produce at the required specifications. The main purpose of VIEC installation in a separation train is to remove most of the emulsified water from the first stage and/or the second stage three-phase separators before it reaches the electrocoalescer as it is demonstrated in the following schematic representation.

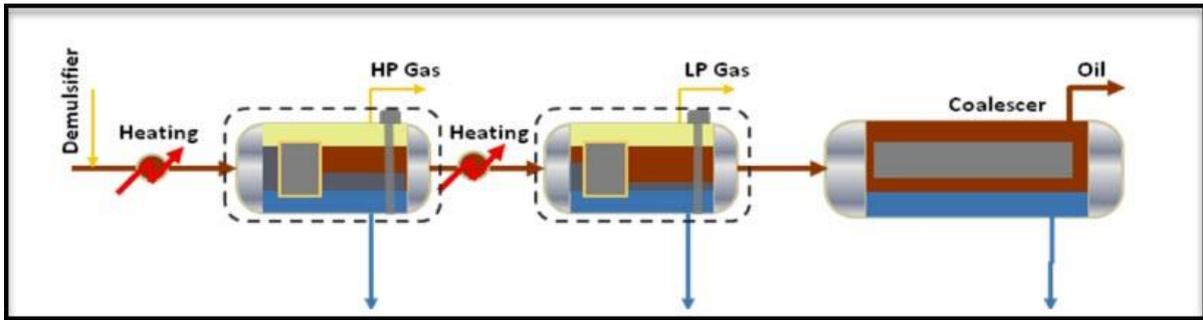


Figure 84: Schematic representation of a typical VIEC technology application in either first and/or second stage three-phase separators. VIEC electrocoalescers resolve the w/o emulsions and separate out as much water possible early in the process, *Overcoming Separation Challenges by Use of Electrocoalescence Technology*, Morad Anarzguioui and Per Christian Jacobsen, Wartsilä Oil & Gas Systems AS, SPE-171483-MS, 2014, [151].

Since VIEC technology is tolerant to high water cut and the presence of gas, it is commonly installed either in the first and/or in second stage separators. This allows the early resolution of the emulsions, it increases the effectiveness of the downstream separation train and it safeguards the compliance of the feed of the electrocoalescers to the water concentration limitation.

In the following figure, the reduction in the thickness of the emulsion layer with the application of VIEC is depicted, with a clear benefit for the level control in the separator vessels.

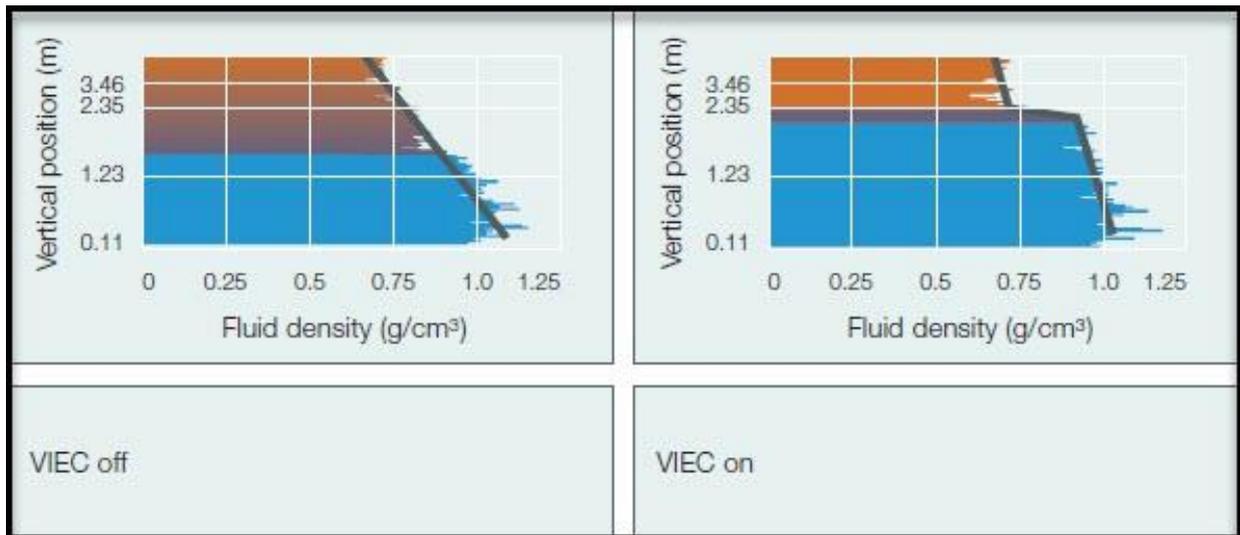


Figure 85: Screenshots showing level profiles without VIEC (left) and with VIEC (right). Comparison between the left and the right image shows significant improvement in interface development, *Innovative Engineering, Vessel Internal Electrostatic Coalescer (VIEC), Novel oil-water separation technology*, ABB review, Wojciech Piasecki, Marek Florkowski, Marek Fulczyk, Jakub Sipowicz, Hans Kristian Sundt, 2004, [152], *Advanced Electrostatic Internals in the 1st Stage Separator Enhance Oil/Water Separation and Reduce Chemical Consumption on the Troll C Platform*, Erik A. Wolff/ABB Offshore Systems, Trond L. Knutsen/Norsk Hydro, Wojciech Piasecki/ABB Corporate Research, Peder Hansson/ABB Offshore Systems, Pål J Nilsen/ABB Offshore Systems, OTC 16321, 2004, [145].

For medium to light crudes the VIEC technology installation may reduce the necessary dehydration or desalting stages of the process [151].

The increased coalescing rates and growth of the water droplets, induced by VIEC technology, leads to the efficient treatment even in the case of high viscosity emulsions and the decrease in the required

treatment temperature. Since water has a heat capacity as twice as the one of crude oil, the application of VIEC in the first separation stage where the water content is high can lead to significant energy savings [147].

VIEC has been installed so far to process oils between 12 and 40 API and has been tested for an oil range from 11 to 40 API.

The case studies in appendix C do not only include the actual implementation of VIEC technology in production facilities but also the tests that were conducted to reassure different operators that VIEC technology can yield the benefits they required.

The tests on VIEC technology aiming to predict its separation performance with accuracy in each case are conducted in three steps. The first step includes the crude oil characterization by quantifying its viscosity, density and dielectric parameters as a function of temperature. The purpose is to evaluate potential emulsion stability and separation efficiency. The second step includes small scale separation batch tests in which synthetic emulsions are created by mixing dry oil with produced water with a high-shear stator-rotor mixer (Figure 86).

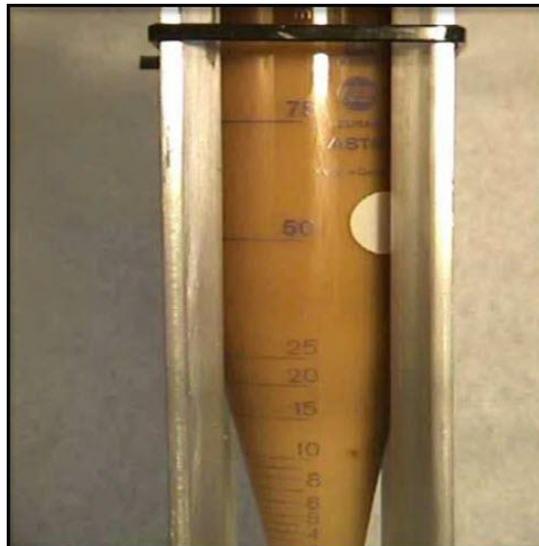


Figure 86: Example of a batch test, Electrostatic Destabilization of water-in-oil Emulsions: Use of laboratory scale experiments to debottleneck and optimize separation processes, Erik Bjørklund, Toulouse Workshop, 2012, [153].

The purpose is to evaluate the separability of the emulsions in varying temperatures, water cuts, demulsifier types and dosages and electrostatic treatment time. However, batch tests are static and cannot be directly extrapolated in full-size separator conditions because they do not include the dynamic effects of turbulence and shear that are met during production. The third step includes flow loop tests that are based on the two previous steps and are conducted under representative flowing conditions [143]. These tests are performed both with and without VIEC application in order to quantify the separation efficiency of the technology. The following figure shows the flow loop rigs in Hamworthy's laboratory.



Figure 87: Laboratory multiphase platform, *Electrostatic Destabilization of water-in-oil Emulsions: Use of laboratory scale experiments to debottleneck and optimize separation processes*, Erik Bjørklund, Toulouse Workshop, 2012, [153].

Compact Electrostatic Coalescer CEC™

One more electrocoalescer for heavy oil dehydration processes is the Compact Electrostatic Coalescer (CEC).

The initial research of the Compact Electrostatic Coalescer (CEC) was conducted at the Southampton University. Based on this research, Statoil patented the idea and issued a license agreement with Kværner Process Systems (KPS) (later Aker Solutions and currently Fjords Processing) to develop and commercialize CEC. Therefore, Fjords Processing owns the exclusive commercial rights

The CEC is an in-line vertical vessel that can be placed upstream an existing gravity separator to enhance its performance. It is considered a relatively small and lightweight electrocoalescer since a typical unit that operates with flow rates of 130000BOPD has a height of 5.5m, a diameter of 1.2m and approximate dry weight of 7t [115]. The basic operating principle of CEC is the application of strong AC electrostatic field on the water-in-oil emulsion which flows vertically in the vessel in turbulent conditions [125]. The CEC consists of annular concentric, fully insulated electrodes placed one inside the other. Voltages of the order of thousand volts create a strong electrostatic field which forces the water droplets to merge within a matter of seconds. The result is water droplets increased by ten times in diameter than before entering the CEC vessel.

CEC is qualified for crude oil of 14 to 40 API. It is tolerant to any water cut but the high performance is between 2 – 40% water cut. The CEC can operate up to 160°C with high viscosity crudes showing efficiency in breaking heavy crude oil emulsions and emulsions created from ESP pumps [154]. An important benefit of CEC is the very low energy consumption and the undisturbed production during maintenance. The production can readily bypass the CEC, through pre-installed lines, directly to the gravity separator and continue at reduced flow rate [115]. Additionally CEC has a very quick-respond power modulation and an

automatic control that protects from short circuiting. In the following figure a CEC unit and a schematic of its operation are presented.

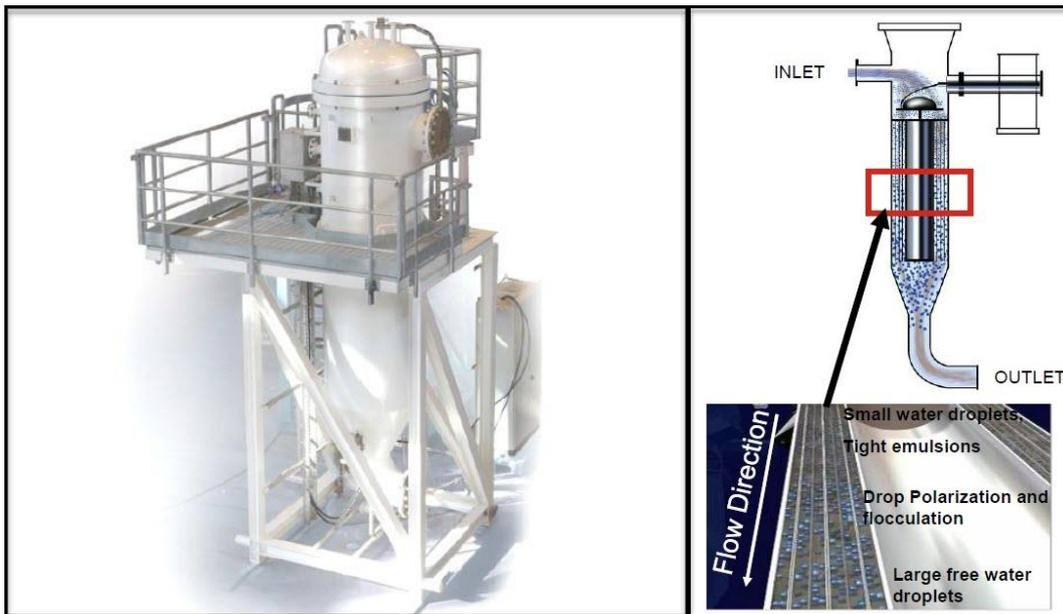


Figure 88: Actual CEC unit (left), schematic of CEC unit (right), *Offshore processing challenges with heavy oils and use of robust high performance technologies, 2nd Tekna Conference on Heavy Oil Technology for Offshore Applications, Mika Tienhaara, Aker Process Systems Norway, 2011, [154]. (published with permission of Fjords Processing)*

The CEC can operate with flow rates from 3 to 220 MBPD and it is also tolerant to vessel motions (level fluctuations) and water slugs but it is limited to a maximum gas volume content of <10%. Taking into account the 2-40% water cut where the device demonstrates the higher performance, the typical installation for CEC is after the second stage separator (Figure 89).

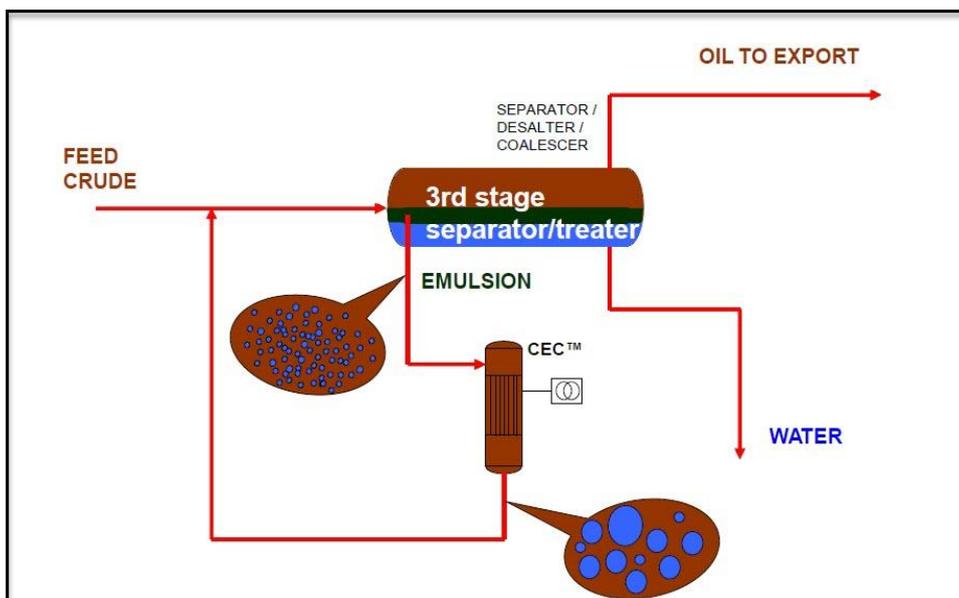


Figure 89: Typical configuration of CEC for emulsion breaking between second and third stage separator, *Offshore processing challenges with heavy oils and use of robust high performance technologies, 2nd Tekna Conference on Heavy Oil Technology for Offshore Applications, Mika Tienhaara, Aker Process Systems Norway, 2011, [154]. (published with permission of Fjords Processing)*

Since this unit is designed to contain scarce amount of gas it should be placed below the lowest liquid level of the upstream separator so the relative elevation will not permit any gas flashing from the oil [113]. Due to its robustness the CEC device was qualified for subsea use from the Norwegian Research Council and Statoil.

In the following table the CEC operational envelope is summarized

Parameter	Unit	Best operation	Operational envelope
Liquid flow rate	% of design	2 - 100	0-150
Watercut	Vol%	5 - 30	0 - 100
Water droplet size	microns	All	All
Op. Pressure	barg	< 30	< 150
Op. Temperature	°C	< 100	< 160
Oil viscosity	cP	< 10	< 30
Oil density	API	19 - 40	14 - 40
Water salinity	%	All	All
Gas volume fraction	%	< 10%	-
	For GVF > 10%, degassing functionality will be implemented		
Sand/particles	Not significant parameter (low erosion and accumulation tendencies)		

Table 11: Operational envelope of CEC device, (published with permission of Fjords Processing)

The CEC technology has been subjected to qualifications testing with high viscosity crudes in a test skid that was designed for 17°API. The test results for a 14°API with no demulsifier added were:

- Test 1: At 70°C and crude kinematic viscosity of approximately 280cSt, the inlet water was between 51 to 56% and the separator outlet water was 45%. With the implementation of CEC the inlet water was 34 – 45% (reduced) and the water in the separator outlet 16 – 22%
- Test 2: At 70°C and crude kinematic viscosity of approximately 520cSt, the inlet water was approximately 20% and the separator outlet water was again 20%. With the implementation of CEC the inlet water was 21 – 25% (reduced) and the water in the separator outlet 11 – 14%.

It is clear that in the second test the separation is not possible without the implementation of CEC whereas in both test the CEC results in emulsion breaking and approximately 50% water removal.

3D Horizontal Flow Electrostatic Coalescer

Another electrocoalescing vessel that is commercialized by Fjords Processing is 3D Horizontal Flow Electrostatic treater. This treater follows the advanced flow path technology that utilizes horizontal flow instead of vertical introduction of the fluid which is used in conventional electrocoalescers. Fjords Processing has developed the horizontal flow electrostatic treater especially for heavy oil treatment and so far it has demonstrated very good separation efficiency. This configuration is based on the Combined Electrostatic Coalescer and Oil Storage Tank (EC/OST) design [155]. The horizontal flow direction

minimizes water carry-over with respect to the upflow conventional electrostatic treaters, where the flow direction of the bulk phase is upwards and the coalesced water droplets should flow countercurrent to the vertical oil flow. In the case of heavy oils the countercurrent flow creates additional problems to separation of the phases. Consequently, in heavy oil dehydration applications horizontal flow is preferable to upflow or countercurrent flow. The horizontal flow configuration is also beneficial to crudes that exhibit foaming tendency.

The cyclonic inlet device of the horizontal flow electrostatic treater prevents foaming and ensures a very good liquid/gas bulk separation. A subsequent series of mounted externally adjustable louver plates create a baffle section that can be constantly adjusted to match the flow conditions, with slight pressure drop, uniform flow distribution and additional surface for the water droplets to coalesce. Then electrostatic grids are placed so that they can prevent short circuiting caused by production upsets and high inlet water cuts by favorably adjusting the distance between the charged and the ground electrode rather than reducing the grid voltage with a voltage transformer. The adjustable spacing between the grids reduces the field density and eliminates short circuiting. The grids have extensions on the bottom and sides which create additional electrostatic field between the vessel wall and the water phase. The combination of adjustable louver baffles and grids renders the electrostatic coalescer effective in a wide range of changes in operating conditions such as changes in flow rates, inlet water cuts and emulsion levels. Matrix packing is also used in the Horizontal Flow Electrocoalescer. In order to resist plugging during heavy oil treatment these matrix packs have a downwardly sloped design with no static zones that allows solids and asphaltenes to be removed without being trapped in the vessel [155].

The 3D Horizontal Flow Coalescer operates in a range of 2 to 35% water cut and produces lower than 0.5% BS&W crude oils of over 14°API (Figure 90).

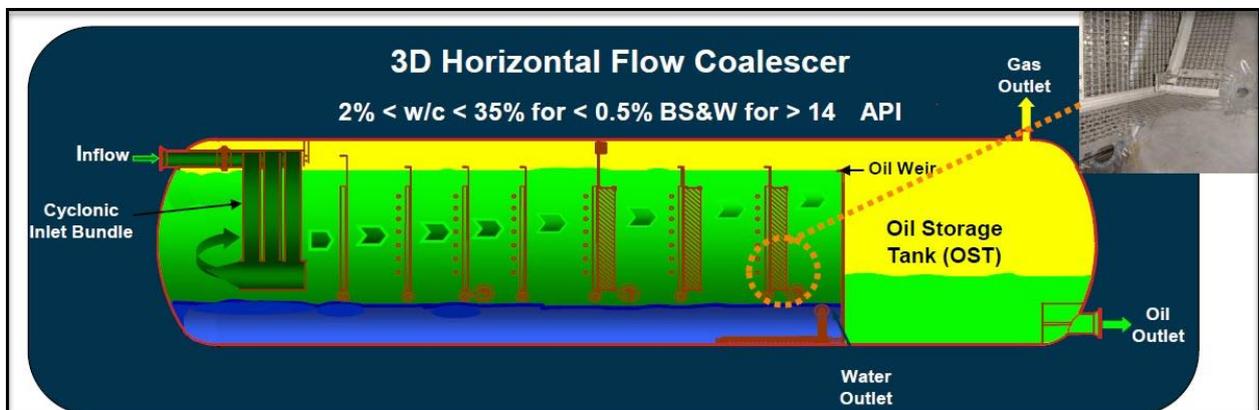


Figure 90: Schematic representation of the 3D Horizontal Flow Coalescer. On the upper right corner a photograph of the electrostatic grids with the extensions on the bottom and sides is provided. Offshore processing challenges with heavy oils and use of robust high performance technologies, 2nd Tekna Conference on Heavy Oil Technology for Offshore Applications, Mika Tienhaara, Aker Process Systems Norway, 2011, [154]. (published with permission of Fjords Processing)

The horizontal flow electrostatic treater can handle efficiently water, oil and gas and thus a separate upstream degasser is not required. The design ensures that 100% of the water-in-oil emulsion is exposed to the electrostatic field so results in smaller size or increased capacity compared to conventional upflow electrocoalescers. Furthermore, the mechanical internals that are adjusted in the 3D Horizontal Flow Coalescer can also be retrofit in a gravity separator. In case that heating is required the Horizontal Flow electrocoalescer can be equipped with internal fire tubes (Figure 91).

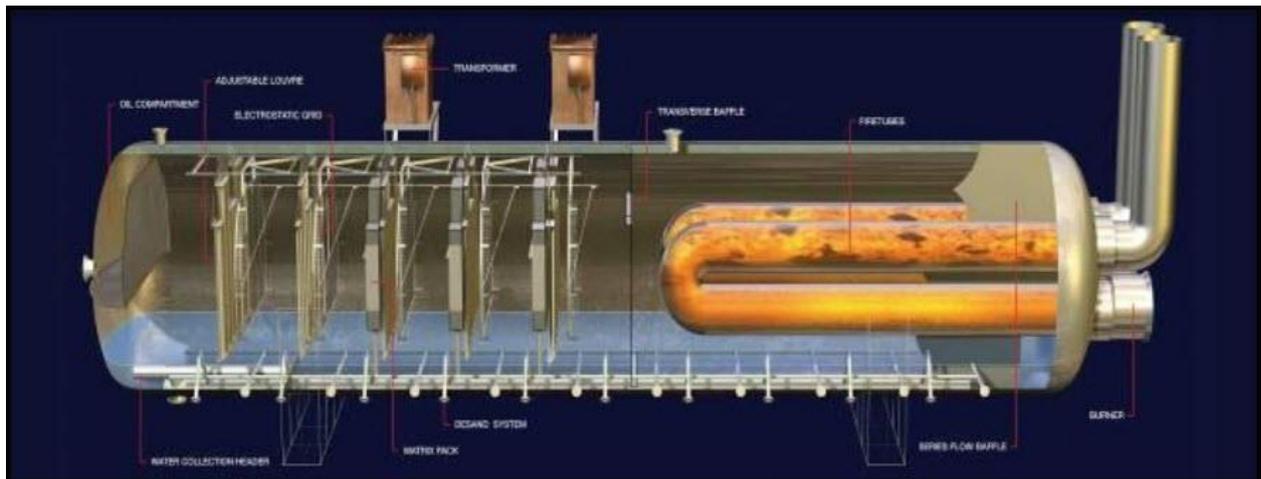


Figure 91: 3D Horizontal Flow Coalescer with internal fire tubes, (published with permission of Fjords Processing)

The following tables include installations of the 3D Horizontal Flow Electrostatic treater in SAGD and CHOPS processing facilities of high water cuts and also Offshore, FPSO and Onshore installations.

Client/Contractor	Project	°API	Application
Husky	Tucker Lake	16° API	SAGD onshore
Nexen	Long Lake South	16° API	SAGD onshore
Occidental	Alturitas	12° API	Onshore
Statoil	Grane	18.5-20° API	Offshore - fixed platform
Petrobras	P34	17° API	FPSO
OGX	Waimea	17° API	FPSO
Husky	Sunrise	16° API	SAGD onshore
PDSVA	PLC Refinery Conversion	16° API	Refinery Desalting
EM	Hebron	18° API	Offshore
OSUM	Taiga	16° API	SAGD onshore
Japan	Canada Oilsands	16° API	SAGD onshore
Brion Energy	McKay River	16° API	Onshore

Table 12: Recent references of 3D Horizontal Flow Electrostatic Treater installations, (published with permission of Fjords Processing)

Client	Project	°API	Inlet	Outlet
Suncor	Firebag	20	30%	<0.5
Norcen	Energy	14	50%	<0.4
Wascana	Plover Lake	12.5	30%	<0.5
Deer Creek	Energy	12.4	71%	<0.5

Table 13: 3D Horizontal Flow Electrostatic treater in SAGD and CHOPS processing facilities of high water cuts, (published with permission of Fjords Processing)

In line electrostatic Coalescer

In 2014 ExxonMobile Upstream Research Company (EMURC) working in a program of subsea separation technology, developed and performed laboratory-scale tests on an InLine electrocoalescer supplied by FMC Technologies. The InLine electrocoalescer was tested with medium and heavy crude oil in varying operating temperatures and a range of oil viscosities [109]. Currently this InLine electrostatic coalescer is being commercialized by FMC Technologies.

The inline electrocoalescence device, just like any other inline device, enhances the efficiency of the downstream separator. The inline electrocoalescer includes insulated electrodes in a ceramic housing. An intense electrical field is formed by a high voltage AC current applied to the electrodes. The dispersed droplets get polarized and collide in a turbulent flow regime. The polarization of the droplets with a short range attraction force enables the break of the interfacial film between the droplets. Additionally, due to the rapid change in the direction of the AC field the droplets deform because they constantly follow the direction of the field at a relatively high frequency [114]. The turbulent flow regime in the inline electrocoalescer increases the collision rate between the droplets and thus their chances to coalesce.

The test results showed that the inline electrocoalescer device is able to process medium and heavy oils achieving high coalescing performance. Because of its compact design it is suitable for subsea and offshore facilities where space and weight restrictions exist and has the potential to improve medium and heavy oil dehydration processes. Compared with electrocoalescers of similar flow capacity the inline electrocoalescer consumes much less power. It is a unit that must be liquid-filled during operations and it is important that the piping downstream of the inline electrocoalescer will not break up the coalesced droplets before they enter the separator.

The inline electrocoalescer can also increase the performance of existing separation equipment, reduce demulsifier addition and energy consumption [109]. A schematic of the inline electrocoalescer is given in the following figure.

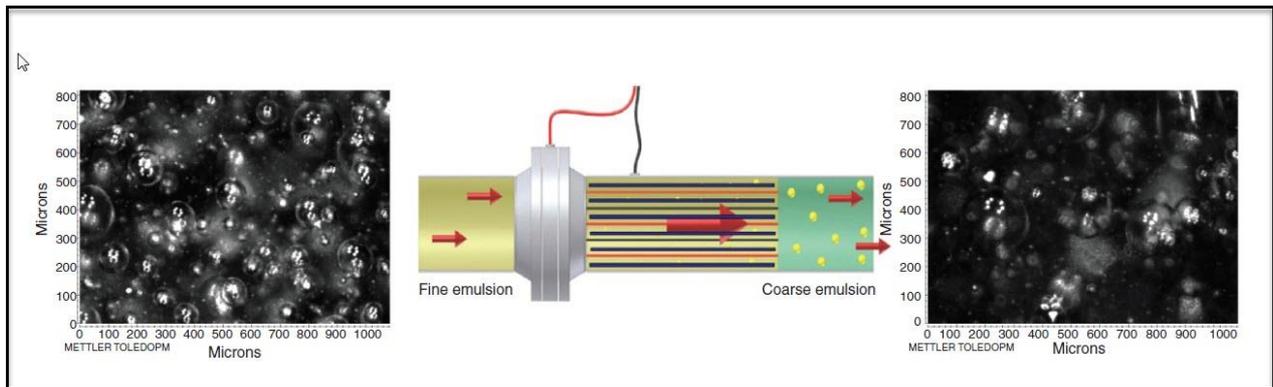


Figure 92: Schematic of the inline electrocoalescer device. On the left is the emulsion entering the device with dispersed water droplets and on the right the emulsion exiting the device with enlarged water droplets. Performance Testing of an Inline Electrocoalescer Device With Medium and Heavy Crudes E.J. Grave and M.D. Olson, ExxonMobil Upstream Research Company; and A.E. Menchaca, R.W. Westra, and M.R. Akdim, FMC Technologies, SPE 174090, 2015, [109].

INLINE SEPARATION TECHNOLOGIES

Inline Separation technology has been implemented during the last decade in many installations and involves separation by applying cyclonic forces in pipe segments instead of large vessels. This technology involves the separation of gas and liquid, oil and water and sand from liquid or multiphase streams. It is already used extensively in retrofit applications to increase the performance and capacity of existing production systems.

Besides the improvement in performance of already existing downstream equipment, inline technology has improved in such a way that total production systems can be developed based on use of inline separators. This can allow for more compact and cost efficient field developments and it will also enable new applications, such as heavy oil and deepwater subsea applications, which are not feasible to develop with conventional technology [156].

FMC Technologies in cooperation with Statoil has developed and commercialized a portfolio of inline separation technologies that covers the whole range of hydrocarbon production. Consequently some of these technologies have direct implementation in heavy oil production and make both onshore and offshore heavy oil projects economically viable. This is achieved through enhanced phase separation from generation of high G forces, reduced residence times resulting in reduction in size and weight of the equipment, reduced operating temperatures and energy consumption, reduced chemical consumption and improved oil and water quality.

– **InLine DeGasser (Gas-Liquid Separation)**: The InLine DeGasser is used to separate gas from the liquid stream and it can be used from moderate to medium gas volume fractions from 0 to 60%. It is a very compact centrifugal inline device and is suitable for newbuilt topside and subsea systems and also for de-bottlenecking existing equipment. In new installations it can degass heavy oil and in retrofit projects it can pre-treat the feed stream to the existing separator in order to reduce the effects of mousse. The InLine DeGasser has two separation stages; in the first a gas-liquid bulk separation is taking place but some liquid still manages to escape with the gas. In the second stage (scrubber) the entrained liquid is separated from the gas. The size of the degasser depends on the section where the second stage

separation is performed. This section can either have the dimension of the pipe or even larger, depending on the application and performance requirements [100]. In the following figures an inline degasser installation and a schematic illustration of the device are presented.

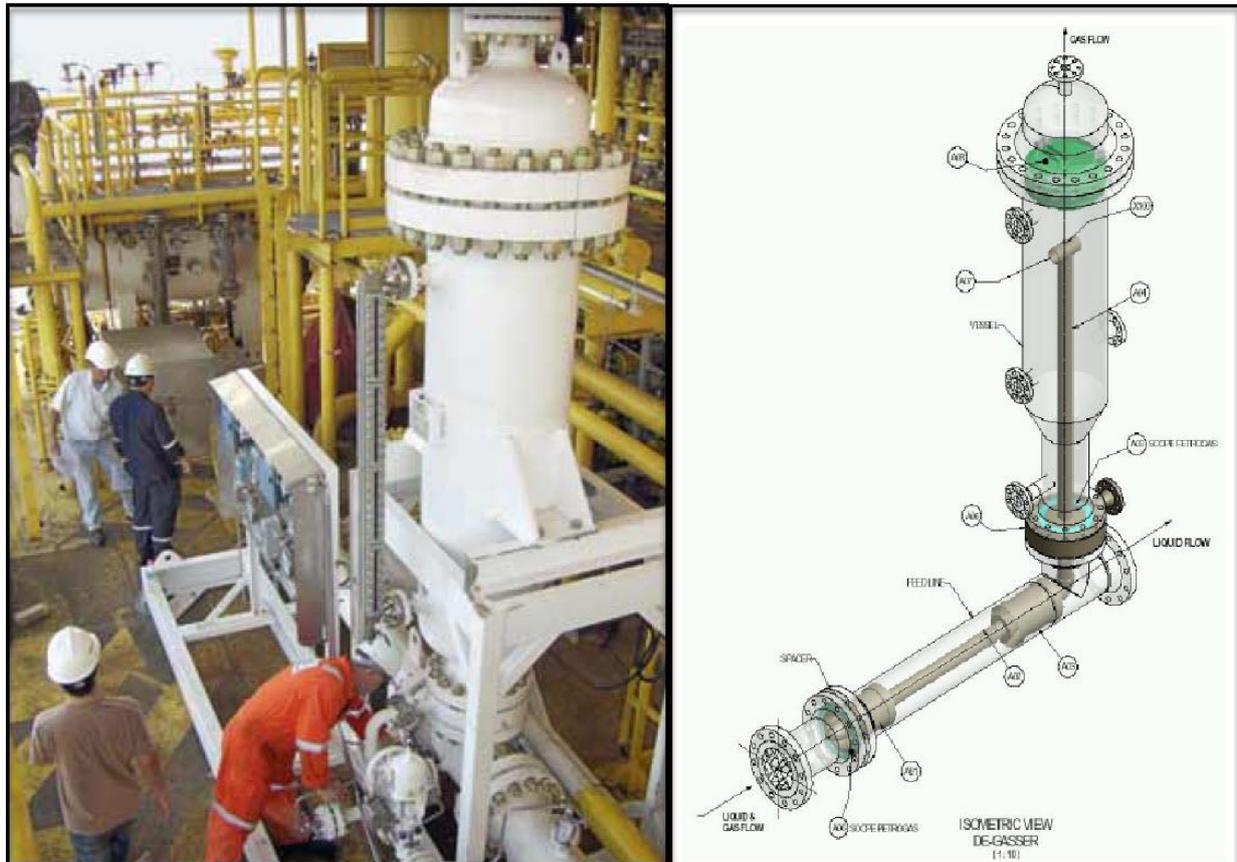


Figure 93: Installation of an InLine DeGasser on the Jasmine field in Thailand (left) and a schematic representation (right), *New Separation Technology for Heavy Oil Applications* Marit Storvik, FMC Technologies, Rune Fantoft, FMC CDS Separation Technologies, [157].

– **InLine DeWaterer (Liquid-Liquid separation):** The InLine DeWaterer is a compact centrifugal inline separator, which removes hydrocarbon liquid from water-dominated streams thus it can be applied when the inlet stream is water continuous. Thanks to its ability to handle up to 30 % - 50 % of gas and water cuts of 50 % and higher, it offers an effective solution for mature fields, new heavy oil topside and subsea systems as well as for debottlenecking and upgrading existing production facilities. This technology is considerably smaller in size than a conventional horizontal separator as it can be seen in the figure.

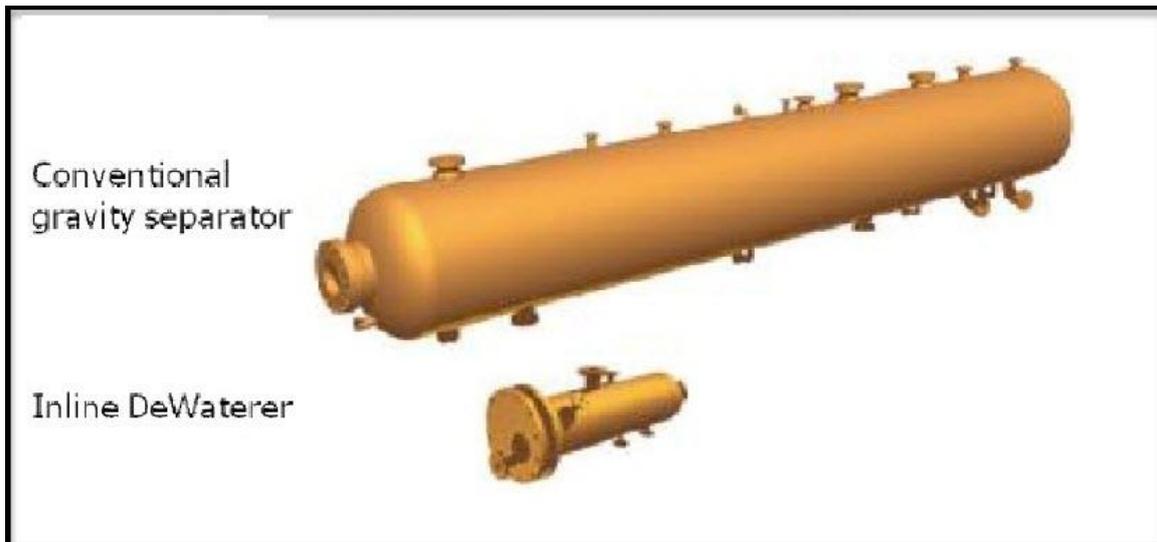


Figure 94: Comparison of the InLine DeWaterer with a conventional separator of similar performance and capacity, *Revolutionizing Offshore Production by InLine Separation Technology*, R. Fantoft, R. Akdim, R. Mikkelsen, T. Abdalla, R. Westra, and E. de Haas, *FMC CDS Separation Systems, SPE 135492, 2010*, [156].

The InLine DeWaterer is a gas tolerant device with no significant change in performance for gas volume fractions up to 50% in the inlet. The typical oil in water concentration in the outlet is 100 – 1000ppm and the pressure drop from the inlet to water and oil outlets 1 – 1.5bar [158].

The InLine DeWaterer is already used in an integrated InLine system for Brazilian company Petrobras in Pazflor field. The integrated system consists of a CDS-Gasunie cyclone, InLine DeWaterers and optimized hydrocyclones. The main challenges of the field were the low energy reservoir, the 17 - 22°API oil and emulsion stability, the high water production that demanded large amounts of methanol to prevent hydrates and the presence of sand.

– **InLine DeSander (Solid-Liquid/multiphase stream)**: The InLine DeSander is a compact centrifugal inline separator, which removes solids from liquid or multiphase streams. It can be installed at the inlet of the production system, upstream and downstream of primary separators, upstream of the hydrocyclones or anywhere in the process where solids separation is required. The InLine DeSander has already been applied in several applications in the oil industry one of which is the multiphase and water de-sanding for the subsea heavy oil application at the Petrobras Marlim field in Brazil. The InLine DeSander requires no power consumption and it is a device with no moving parts. The operating principle is relatively simple. The inlet flow enters the separator axially and meets an axial swirl element which generates high rotational velocities. The generated G-forces lead the solids to the outer surface and then downwards to the solids outlet. The solids exit the outlet in the form of particles or in the form of slurry. This separation of solids from the stream is done in low pressure drop. In the following figure the operating principle of the InLine DeSander shown in a schematic representation [100].

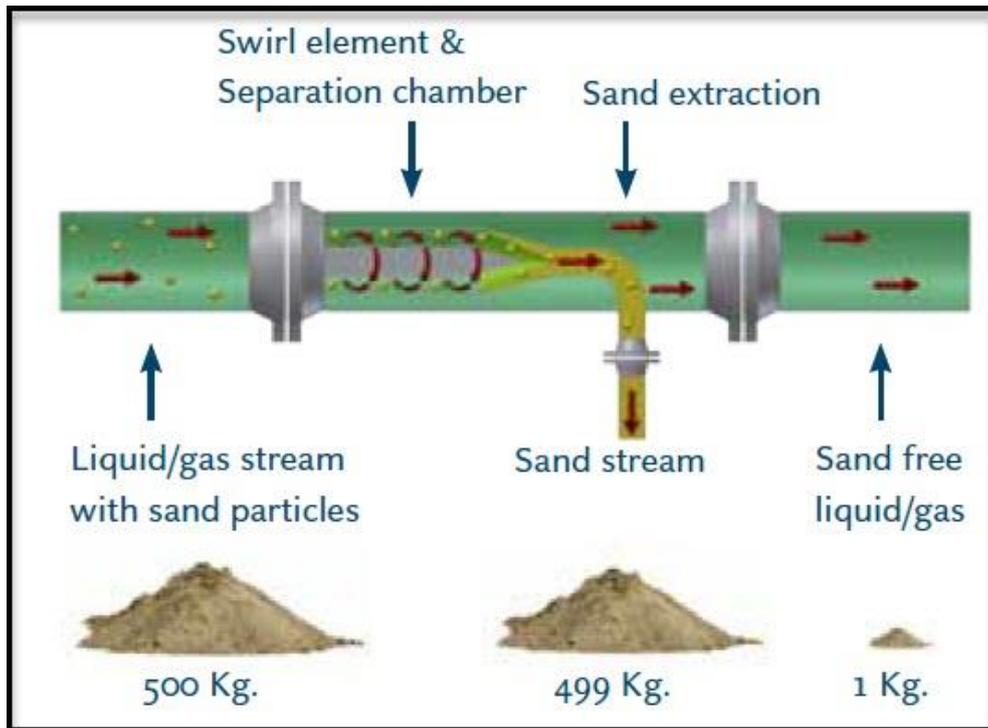


Figure 95: Schematic representation of the operating principle of the InLine DeSander, Revolutionizing Offshore Production by InLine Separation Technology, R. Fantoft, R. Akdim, R. Mikkelsen, T. Abdalla, R. Westra, and E. de Haas, FMC CDS Separation Systems, SPE 135492, 2010, [156].

HEAVY AND EXTRA HEAVY OIL TRANSPORTATION

INTRODUCTION

Globally, the demand for crude oil has increased from 60 million barrels per day to 84 million barrels per day during a period of 20 years as it was estimated in 2010 [159]. In the same time, the expected decline of conventional hydrocarbon sources comprising from medium and light crude oils has created great interest in the exploitation of heavy, extra heavy oil and bitumen sources despite the profound difficulties of their production [160]. In the past, the high costs in recovery, processing, transportation and refining of this type of hydrocarbons compared with their low market value rendered their exploitation uneconomic. Today, due to intense scientific research and remarkable technological developments, this exploitation is feasible.

The technical challenges and operational difficulties concerning the recovery and production of heavy, extra heavy oils and bitumen have already been discussed extensively in this work and they are mostly attributed to their high viscosity ($10^3 - 10^6$ cP), low API gravity, low mobility and flowability, asphaltene and wax deposition as well as sulfur and salt content. The same challenges and difficulties are also met during their transportation through pipelines.

Pipelines are the most common, least expensive, environmentally friendly and most effective means of continuous transportation of crude oils and derived products from the field to the refinery [2]. It is noteworthy that Canada and Venezuela, the two major countries presently exploiting heavy crude oil reserves, are using pipelines to transport 95% of the heavy crude oil they produce. This form of transportation, even though is widely used, is also challenging. Problems such as asphaltene precipitation due to depressurization below asphaltene precipitation onset pressure may occur when transporting heavy crude oil [161]. However, big pressure drops are expected when transporting heavy crude oils via pipelines due to their high viscosity. To counteract this, high pump power is needed which renders pipelining of heavy crudes energy intensive. Additionally, clogging of the pipeline wall due to asphaltene deposition reduces the accessible cross sectional area with consequent decrease in flow rate and increase in pressure drop [2] [161].

In order to mitigate the problems and transport heavy crude oils and bitumens economically, the pressure drop in the pipelines should be lower. Due to the heavy crude's high viscosity, the conventional pipeline transportation, which is used in the case of medium and light crudes, is not adequate. Therefore, a reduction in their viscosity can facilitate the transportation with pipelines [162]. The methods used to facilitate heavy crude oil and bitumen transportation are presented in figure 96.

There are three main categories of methods used in heavy oil pipeline transportation.

(a) Viscosity reduction, that can be achieved by dilution with lighter hydrocarbons or solvent, heating the oil and the pipeline, emulsification through the creation of oil-in-water emulsion and pour point reduction by using pour point depressants (PPD).

(b) Friction reduction, that can be achieved by lubricating the inner surface of the pipeline through the combined use of drag reducing additives and core annular flow.

(c) In-situ/partial upgrading of the produced oil to an oil with improved viscosity, higher API gravity and reduced content in pollutants such as asphaltenes, sulfur and heavy metals.

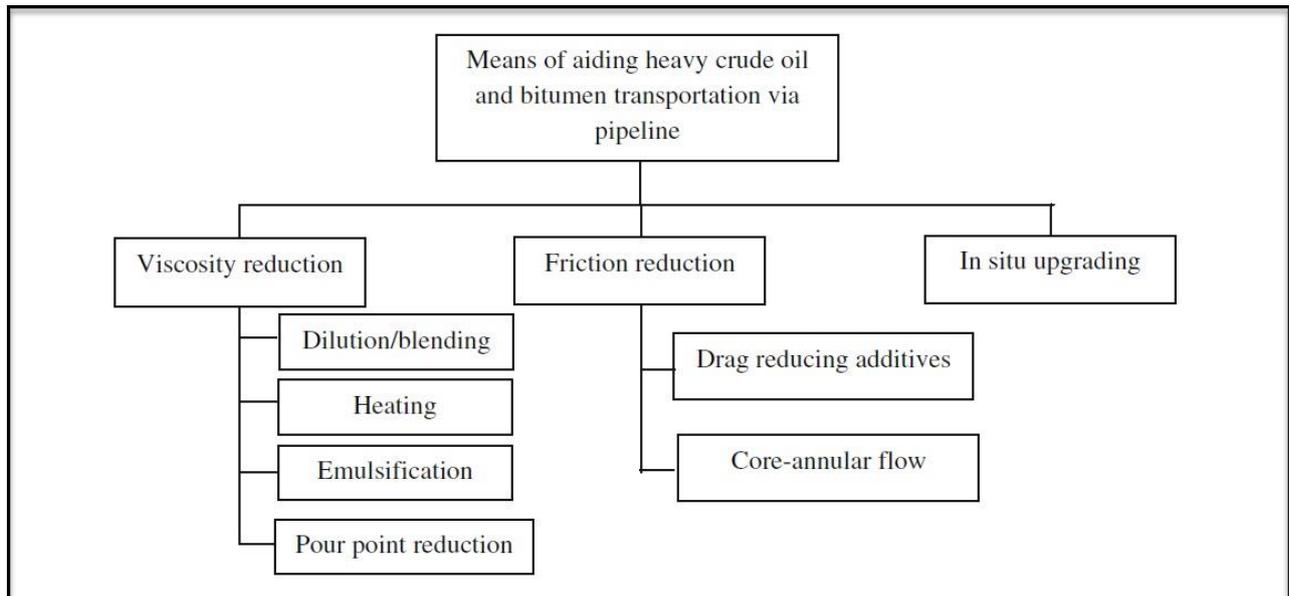


Figure 96: Methods that enhance heavy crude oil and bitumen transportation through pipelines, A review of technologies for transporting heavy crude oil and bitumen via pipelines, Abarasi Hart, J Petrol Explor Prod Technol (2014) 4:327) 4., [160].

The upgrading method modifies the relative proportion of the oil hydrocarbons without changing its refining characteristics, by hydroprocessing. The process breaks the carbon-carbon bonds, producing smaller molecules of paraffins and olefins and resulting in reduced viscosity of the crude [1] [7]. Upgrading processes can take place in-situ, i.e. in the reservoir during EOR techniques, on-site as a part of crude oil processing on the field, or/and in the refinery. The EOR upgrading methods were analysed in previous chapters, so only the partial upgrading techniques that are used on-site will be further analysed, since this category falls within the scope of this work.

VISCOSITY REDUCTION

Dilution

The oldest and most commonly used method to reduce the viscosity of heavy and extra heavy crude oils is dilution. This method has been used by the Petroleum Industry since the 1930's and it consists of blending the heavy and extra heavy crude with diluents that are less viscous. It is based on the fact that the lower the viscosity of the diluents, the lower the viscosity of the blend thus, the result is a blended mixture of lower viscosity than the initial crude [163].

The diluents usually are liquid hydrocarbons that may include condensates from natural gas production, naphtha, kerosene. The potential use of organic solvents such as alcohol, methyl tert-butyl ether, tert-amyl methyl ether has also been studied [164]. These solvents are used in the refinery to improve the octane number of gasoline, but they have also shown promising results in reducing the viscosity of heavy hydrocarbons [163]. The viscosity of the blended mixture, besides the type of the diluent, depends on the

operating temperature and the dilution rate. The amount of the required diluent for a heavy crude oil is expressed by its ratio in the mixture and it usually ranges from 20 to 30%. This quantity is usually enough to avoid pressure drops and high transportation temperatures. Also the addition of diluents in the crude may facilitate other processes such as dehydration and desalting. On the other hand, the dilution of the heavy crude results in additional investments in pumping and pipelines, since the volume of the transported liquid increases. The solvent is usually reused, so it is separated from the mixture downstream, processed, and returns to the production site from a separate pipeline [2].

Dilution of extra heavy crude oil with **light natural condensate** (C₅₊) yields a blended mixture with substantially lower density and lower viscosity than the initial crude. Natural gas condensate is a mixture of hydrocarbon liquids with low density and moderate viscosity and is a by-product of natural gas processing. Canada and Venezuela have been using natural gas condensate to dilute heavy crude oil and bitumen to enhance its transportation through pipelines from the oil fields to the refinery. Even though the blended mixture of heavy oil or bitumen with the condensate has improved density and viscosity, a risk of asphaltene instability exists. Due to the fact that the asphaltenes contained in the crude are insoluble in n-pentane and n-heptane, the dilution with light gas condensate which is rich in paraffins may lead to asphaltene aggregation and precipitation during their transportation and storage [165] [166]. The limitations in the use of condensate as a solvent derive also from several other factors. Initially, the availability of the gas condensate as a solvent depends on the natural gas demand [167]. However, the growing production of heavy crude oil and bitumen requires more condensate as a solvent than it is available.

The use of **light crude oil** as a solvent is less efficient than the natural gas condensate in lowering heavy crude oil and bitumen viscosity [168]. Additionally, the problems of compatibility during mixing and the availability of the diluent are still present. Kerosene has been found to yield good results concerning the viscosity and the density of the mixture when it is used as solvent. A correlation of the expected viscosity of the resulting mixture based on the viscosities and the volume fractions of the heavy crude and the diluents has been developed since 1933 and it is the following [169]:

$$\log \mu = \left(\frac{\alpha V_o}{\alpha V_o + V_d} \right) \log \mu_o + \left(1 - \frac{\alpha V_o}{\alpha V_o + V_d} \right) \log \mu_d \quad (13)$$

where V_o and V_d, are the volume fraction of the heavy crude oil and diluents, μ_o and μ_d are the viscosity of the heavy crude oil or bitumen and the diluents, respectively, and α is an empirical constant ranging from 0 to 1. The empirical formula for the determination of constant α depends on the viscosity ratio of heavy hydrocarbons to diluents (light hydrocarbons) and their densities and it is given by the expression:

$$a = \frac{17,04(\rho_o - \rho_d)^{0,5237} \rho_o^{3,2745} \rho_d^{1,6316}}{\ln \frac{\mu_o}{\mu_d}} \quad (14)$$

The use of light hydrocarbons is an expensive solution and the required large quantities are not always available. Therefore the recycling and reuse of diluents is a common practice. For instance, in the Orinoco projects in Venezuela the diluent, which is used to transport the extra heavy crude from the production facilities to the special upgrader located on the coast, is separated and it is sent back to the production site for reuse [7]. In Figure 97 a schematic representation of an SAGD facility in Canada is shown where the diluent is also recycled from the upgrader back to the production facilities.

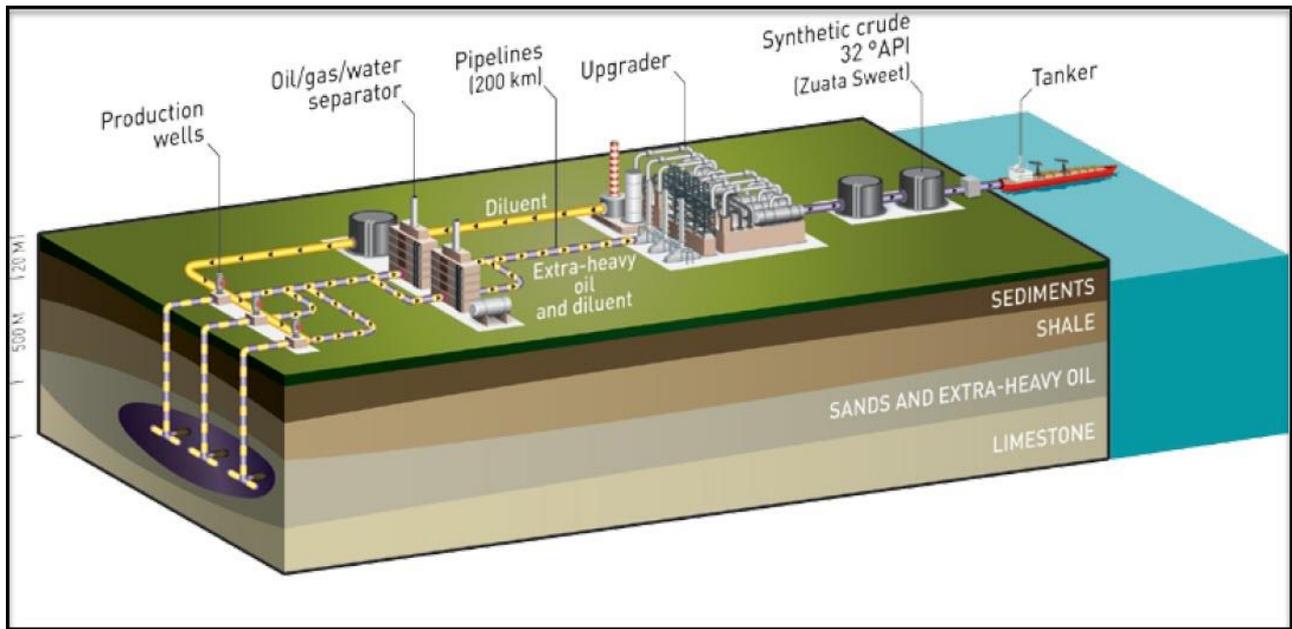


Figure 97: Orinoco belt Petrocedeño facilities, Crude oil of an average 8.3 °API gravity is diluted with naphtha. After separation, it is then transported to the upgrader, which yields a commercial-grade light synthetic crude of 32 °API gravity. The diluent from the upgrader is returned to the production field. Total, Strategic Sectors, Extra-Heavy Oils & Oil Sands, the challenges of development, Exploration and Production, 2012.

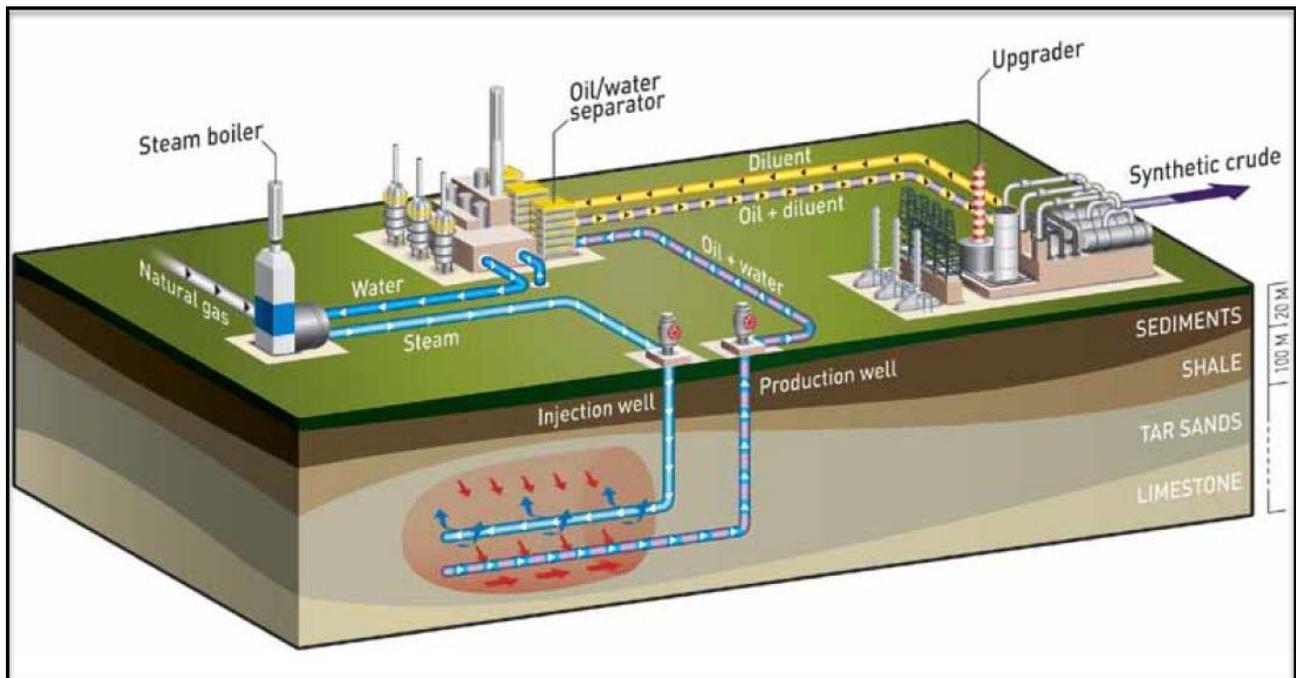


Figure 98: Suront project, Canada, SAGD heavy oil production facilities. The oil and diluent are blended and sent to the upgrader, from which the diluent is sent back to the production facilities to be reused, Total, Strategic Sectors, Extra-Heavy Oils & Oil Sands, the challenges of development, Exploration and Production, 2012

On the other hand, the bitumen extracted from the oil-sands in Canada is also mixed with diluent but the bitumen/diluent mixture (DilBit) is sold as a single product and so the diluent is not recycled. The availability and price of the light gas condensate used in the production of DilBit renders this method viable. The development layout of Joslyn North Mine in Athabasca is shown in Figure 99 where the aforementioned procedure is applied.

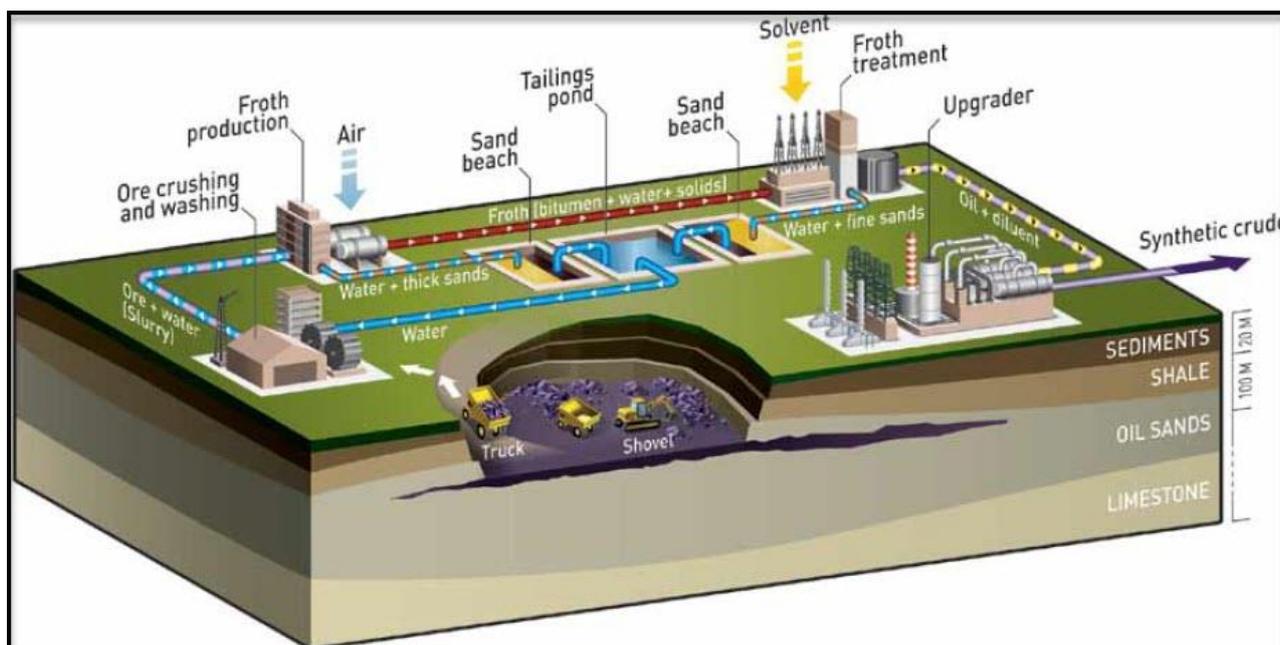


Figure 99: The Joslyn North Mine, the bitumen mixed with the diluent produces DilBit, an exporting bitumen blend. The diluent in this case is not reused. Total, Strategic Sectors, Extra-Heavy Oils & Oil Sands, the challenges of development, Exploration and Production, 2012.

The reused diluent usually requires a separate pipeline from the separation facilities which is costly. However, the re-use of the diluent consumes less solvent than the DilBit method so it is less prone to solvent's availability and price issues.

Another diluent that is commonly employed is **naphtha**. Naphtha is a petroleum fraction with high API gravity and blended with other organic solvents can reduce the amount of diluents needed to lower the viscosity of the heavy crude to pipeline transportation specifications. The efficiency of naphtha with organic solvents blend is attributed to the increasing polarity of the hydrogen bonding of the solvents. A high polarity solvent causes large viscosity reduction of the diluted heavy crude oil and exhibits very good compatibility with the asphaltenes [163].

One solvent that was proposed as more efficient than the light distillates is dimethyl-ether (DME). If DME is used under pressure it can reduce the viscosity of the heavy crude as well as reduce the pressure drop in the pipeline during transportation. One more benefit of DME is that it can be easily recovered in the refinery.

Several other dilution methods have been proposed in order to achieve viscosity reduction of the heavy crude oil. Some of the methods include dilution and partial upgrading during which the solvent is produced through the separation, distillation and thermal cracking of the heavy hydrocarbon feed [170]. An alternative approach for transporting extra heavy crude oil through pipelines was developed in 2006. According to this method, breaking the oil colloidal system by inducing asphaltene precipitation and then re-introducing the asphaltenes in the oil in a non-colloidal form reduces significantly the viscosity of the heavy crude [171].

Heating

The second most common method is to preserve the elevated temperature at which the oil reaches at the wellhead by insulating and burying the pipeline to conserve heat. However, the insulation is not always sufficient and the heat losses during production may be compensated by heating the oil as well. Preheating the heavy crude is always followed by subsequent reheating in the pumping stations [2].

The enhancement of the flowability of crude oil at elevated temperatures has been extensively studied. In a study in 2006, it was concluded that the effect of the increased temperature is more intense in medium than in heavy crudes like Canadian Athabasca, Venezuelan oil sands and Mexican Ku-Maloob_Zaap and Ayatsil-Tekel crude oils. This study has also indicated that preheating alone is not enough to maintain the flowability of the oil and periodical reheating is required [172].

Applying heat involves a lot of technical considerations such as the rapid expansions and contractions of the pipelines, a big number of pumping/heating stations, accurate estimation of the heat losses along the pipeline, high corrosion rates in the internal wall surface due to the high temperatures, changes in the rheological properties that may be the cause for flow instabilities and possible changes in the emulsified structure of the oil. Additionally, heating can be an expensive method for viscosity reduction of the heavy crude, especially when this method is applied in cold environments [173] [7] [1].

The need for direct heating along the pumping stations can be compensated by generating a shear heating effect. In this case the temperature of the oil can be controlled by the appropriate combination of pipeline diameter, distance between the stations, operating pressure range and the viscosity specification of the transported oil. The proposed method is applied in pipelines at least 250km long with the optimum length being 500km and the transported fluid must be heavy oil diluted with a liquid hydrocarbon having at most five carbon atoms and vapor pressure higher than the atmospheric.. The pressure drop should be at least 1250psia (86bar) between the stations because the objective is to use the shear rate for heating the oil. In this case, the temperature increases through friction inside the pumps and through energy losses at high velocities inside the pipelines. The resulting temperature increase can be half degree Celcius for every 15-30 km, depending on the design of the pipelines. The shear friction is directly dependent on the viscosity of the crude oil. The higher the viscosity the higher the shear friction and the more heat is generated. The design must take into consideration emergency shut down conditions and this is the reason why diluted heavy oil must be used. This method is not applicable in medium and light oils since their viscosity is too low and consequently the shear heat effect is very small [174].

In subsea environments, the cooling effect of the surrounding water is unfavorable for transporting heavy and extra heavy oils through pipelines. These oils tend to become thick and viscous in subsea temperatures so the insulation of the pipelines is effective only for distances less than a mile and for high velocities of the transported fluid. Passive insulation is not enough for long subsea distances and although many alternatives have been proposed, such as the use of expensive subsea booster pumps or auxiliary heating along the pipelines, the majority of the solutions is uneconomic. However two promising alternatives that include electrical heating exist: The single heated electrically insulated pipeline (SHIP) and the pipe-in-pipe subsea pipeline [175]. In the first configuration, electrical current flows along the pipe. The power and voltage requirements for direct electrical heating of the pipeline and power transmission are within the conventional AC power limits and they are already available on platforms' power plants at the standard 60-Hertz of frequency, so there is no need for special purpose generators for pipeline lengths up to 65 km. In the second configuration the oil flows through an inner electrically

insulated pipe which is surrounded concentrically by an electrically conductive outer pipe. The drawback of this configuration is that it is costly to implement and to operate due to its design complexity.

Emulsification of the heavy crude oil in water

Emulsification is a process that usually occurs during petroleum production and pipelining. The most common types are the water-in-oil emulsions, which have already been discussed in this work, and the complex oil-in-water-in-oil emulsions. Both types can result in viscosity which is higher than the viscosity of the non emulsified crude and they are difficult to break during dehydration and desalting processes.

On the other hand oil-in-water emulsions are easier to break and have relatively lower viscosity than the heavy crude. The formation of oil-in-water emulsions is therefore an alternative technique in heavy oil transportation because the reduced viscosity of the o/w emulsion enhances the flowability of the heavy crude through the pipelines [176]. In this technique the oil droplets are stabilized in a water continuous phase with the help of surfactants. [177]. The mechanism of the surfactant monolayer is to sit on the oil water interface and prevent the growth of the droplets and phase separation into distinct oil and water phases. The surfactants gather on the interface of the oil and water emulsion with the hydrophilic side (polar region) in contact with water and the hydrophobic side (non-polar region) in contact with the oil [178]. It is a possible solution for locations where hydrocarbon diluents or lighter crude oils are not available while brine, formation water or sea water are available for emulsification. There are cases where the o/w emulsions are created on purpose to facilitate the flowability of extra heavy crudes due to viscosity reduction [179].

The oil-in-water emulsions do not form spontaneously and they need an energy input in order to be created. This energy input comes from shaking, stirring or some other kind of static/dynamic mixing process with the use of devices such as dispersing machines, rotor-stator mixers, colloid mills, high pressure homogenisers applying shearing stresses, emulsification by membrane and ultrasonic waves [159]. If the oil-in-water emulsion stability is successfully achieved then in case of resolution due to emergency shut-downs, the reemulsification of the oil does not pose a problem.

The type of the emulsion depends on the surfactants. In the case of o/w emulsions the surfactant should be soluble in water (the continuous phase), otherwise the heavy oil-in-water emulsions cannot be formed. The surfactant should be initially dissolved in water or brine and then mixed with the oil in order to migrate to the oil water interface. The use of a rotor/stator mixer is essential to the emulsification process. However, the dynamic mixers were found to create dispersed oil droplets with a mean diameter size of less than 10 μ m with an increased viscosity value for the emulsion. Conversely, the creation of a stable oil-in-water emulsion with the help of static mixer consisting of baffle elements yields an efficient dispersion of 27 μ m mean diameter size oil droplets in water and a reduced emulsion viscosity [180] [181] [182]. Another o/w emulsification technology involves the use of an eccentric cylinder mixer that allows low laminar flow in order to reduce the oil viscosity. This geometry achieves very efficient mixing even in highly viscous oils like Zuata and Athabasca crude oils [183].

The use of surfactants with low molecular weight which reduce the interfacial tension is essential for the stabilization of the oil-in-water emulsions. Additionally, in some cases, chemical stabilizers with high molecular weight are needed to avoid phase separation.

The rheology of the oil in water emulsion depends mainly on the volume of the dispersed oil and the drop size distribution [184]. Likewise, the drop size distribution depends on the mixing energy and pressure

and the type of the surfactant defined by the properties of the polar hydrophilic head and the non-polar hydrophobic tail. The most commonly used surfactants are the non-ionic such as Triton X-114. Non-ionic surfactants represent a good choice due to their ability to withstand the salinity of the produced water, they are relatively cheap, their emulsions are easy to separate and they do not yield organic residues that affect the oil properties [185] [186]. Also bioemulsifiers deriving from different strains of the *Acinetobacter* bacteria have been extensively studied and they show a very good performance as oil-in-water emulsifiers both in fresh and sea water [187].

The resulting emulsion must contain as much dispersed oil as possible and as less continuous water as possible. Even though in laboratory tests the required viscosity (approx. 400cP) for transportation at ambient temperatures can be obtained with only 25 – 30% w/w water content, in practice water cuts of less than 30% lead to high viscosities and inversion of the emulsion from oil-in-water to water-in-oil.

Especially in the case of transportation in permafrost regions, insulated large diameter pipes are used at working temperatures below 273.15°K. The oil-in-brine emulsion contains 30-60% water with dissolved salts at sufficient amounts to prevent freezing. At these conditions the pipes can be buried without causing thawing of the permanently frozen soil (permafrost) [188]. For the creation of the o/w emulsion it is suggested to use a mix of common surfactants. Also the use of the xanthan biopolymer can guarantee the emulsion stability. The brine should contain dissolved salts and freezing point depressants - like ethylene glycol - at concentrations that allow the o/w emulsion to be stable at 253.15°K or less [2].

The emulsification process is conducted in three stages that must be followed. The first stage is the formation of the o/w emulsion, the second stage is the transportation of the emulsion through the pipeline infrastructure and the third is the separation of the oil phase from the water phase. The third stage is very important because the surfactants should allow a simple and efficient rupture of the o/w emulsion before crude oil refining. Additionally, the separated water should be treated properly in order to comply with the current environmental and industrial regulations for discharge or recycling. A number of techniques have been developed concerning the o/w emulsion separation that include thermal demulsification, electro-demulsification, chemical demulsification, freeze-thaw method, pH modification, addition of solvents and demulsification by membranes [189] [190].

There have been a lot of studies and several advances during the past few decades concerning the formation of oil-in-water emulsions, their stability and their properties. Oil-in-water emulsions are very complex systems that are difficult to be stabilized. There are cases especially with extra heavy oils that the formation of a stable oil-in-water emulsion cannot be accomplished. Factors such as the HLB (hydrophilic-lipophilic balance) values of the surfactants, the multiple interactions oil-water-surfactant and the possible molecular rearrangements at the oil/water interface should be taken into account but even then it is still difficult to predict the phase and flow behavior of the system [2].

The most widely known application of the oil-in-water emulsion for transportation was the ORIMULSION® process that was developed by PDVSA (Petróleos de Venezuela) and it is based on the research conducted by the Research and Development Subsidiary, Intevep S.A., jointly with British Petroleum [191]. The aim of the research was to reduce the cost involving the Cerro Negro bitumen transportation through pipeline. Cerro Negro is part of the Orinoco Oil Belt in Venezuela and the pipeline transportation of its 8.5°API bitumen was initially achieved through blending with lighter crude. However, in the late 80's the bitumen's dilution with lighter crude oil became economically unattractive in contrast to the ORIMULSION® alternative.

The commercialization of ORIMULSION® in 1988 was undertaken by Bitumenes Orinoco S.A. (BITOR S.A.) In ORIMULSION® proprietary process the bitumen is mixed with water and a surfactant chemical in order to produce a stable emulsion which can be transported by pipeline and by ship in a similar way as the fuel-oil. More specifically, ORIMULSION® consists of 70% natural Cerro Negro bitumen of 8.5°API dispersed in 30% fresh water with the addition of less than 1% alcohol-based surfactants (emulsifiers) that allow the bitumen droplets to remain suspended and maintain the emulsion stability. One additional advantage of ORIMULSION® was that it could be directly used as a fuel for the generation of electricity in existing thermal power plants designed for the combustion of fuel oil and/or coal, using conventional equipment with only minor modifications. The CO₂ emissions from power plants using ORIMULSION® was 20% lower than the ones from coal-fired power plants and the amounts of ash were 50 times less using ORIMULSION® than using coal.

The ORIMULSION® product was clearly successful and up to the late 90's there was an increasing demand for the new fuel which resulted in the increasing capacity of the manufacturing and production facilities in order to keep up with this demand. However, the low value of the emulsified or diluted non-conventional hydrocarbons in conjunction with the steady increase of oil prices and consumption from 1990 to 1997, the new technologies for production of non-conventional crude oils, the temporal reduction in royalty payments and the application of the normal corporate income tax rate instead of the oil rate, made more attractive the development of upgrading facilities to convert the Orinoco Belt hydrocarbons from approximately 7-9° API to lighter, sweeter upgraded crude with 14-30° API.

Consequently, the in-situ upgrading of extra heavy oil and bitumen in order to produce exportable blends or syncrude became more profitable than ORIMULSION® and PDVSA announced the closure of its filial BITOR in 2003 [192].

The exploitation of the Orinoco Belt has been achieved through three very diverse technological methods, namely, Dilution, ORIMULSION® and Upgrading. ORIMULSION® and Upgrading were both used to monetize the Orinoco Belt but in entirely different ways; The Upgrading option was targeted to the transportation sector as the synthetic oil could be used as feedstock for refineries, while ORIMULSION® was targeted to the electric sector as part of the fuel mix for electricity generation.

Other studies include the flow rate tests performed with 13° API oil from a field in Sghanjiasi (China). The pressure loss reduction was around 80% for emulsion with water content from 60 to 80% [193]. It is noteworthy that high reduction in pressure loss was achieved with highly viscous oils at low temperatures.

Another technology that incorporated emulsification, characterization of transport properties and evaluation of storage and recovery conditions by de-emulsification was TRANSOIL® which was developed through a joint project between BP Canada and Alberta Energy Company. The emulsions produced by the TRANSOIL® technology were pumped in a continuous mode for six days with a rate of 80m³/day and subsequently stored for another six days without exhibiting any sign of degradation. The specifications for heavy oil BS&W were 0.5% and the oil content specification for the separated water after demulsification was less than 100 ppm [194].

Heavy oil emulsification was tested in Sicily and Adriatic Sea in transport and production applications. An aqueous phase was injected straight into the well without any change to the existing system. The reservoir production when the water phase got in touch with the oil phase contained 70% of emulsified oil resulted in viscosity that was 30 – 50 times lower than the one obtained by dilution. Also, the oil productivity was four times greater compared to conventional production by diesel dilution [1].

Pour point reduction

The pour point of the oil is the temperature at which the oil loses its flow properties and ceases to flow. In the case of heavy oils, high pour points usually occur with high paraffin (wax) content. The paraffins have also the tendency to precipitate when temperature decreases. The result is an accumulation of continuously growing wax precipitates up to the point where the oil can no longer flow [195].

The most attractive alternative to prevent wax deposition is the use of polymer compounds that are acting as pour point depressants. Pour Point Depressants, also known as wax crystal modifiers, are chemical additives that affect the nucleation and solubility of waxes as well as the shape of wax crystals and thus they can help in depressing the pour point of the crude oil, its viscosity and yield stress [196] [2]. The addition of copolymers such as polyacrylates, polymethacrylate, ethylene-co-vinyl acetate [197], long-chain fatty acid amides, poly-n-alkyl acrylates methacrylate [198] and many more additives prevents deposition and stabilizes transportation [199]. Viscosity measurements at temperatures below the wax appearance temperature, showed that there is strong influence of the copolymer in the reduction of viscosity [197].

The typical polymer concentration is 20-50% in hydrocarbon solvents (ex.toluene) and the commonly employed injection rates are 100-1000ppm. The injection must be conducted while the crude oil is still hot before the wax crystals are formed [196].

A study over the use of 12 different commercial depressants on a waxy North Sea crude oil showed that they all achieved a viscosity reduction from 1000 to 10cP for temperatures between 283 to 293°K. These substances are capable of building up in the wax crystals and change their growth rate and surface. Consequently the wax crystals cannot grow in size so easily and their adherence on the metal surfaces of the pipelines is reduced [198]. Pour point depressants which contain oil-soluble long-chain alkyl group and a polar part in their structure prevent the formation of a wax crystal network. The long-chain alkyl group is inserted in the wax crystal and the polar part crops out on the wax surface preventing the increase of the crystal size [200].

FRICION REDUCTION

Drag reducing additives

The purpose of these additives is to delay the onset of the turbulent eddies and consequently allow higher flow rates with the same pressure drop. [201]. The addition of the drag reducer, just by a few parts per million, helps to reduce turbulence [202]. When the drag reducer is added it dampens the turbulent fluctuation in the near wall region of the pipeline while the flow of heavy oil towards the core of the pipeline is laminar or slightly turbulent [203].

Drag reducing additives can be divided in three main groups: surfactants, fibers and polymers. The surfactants can reduce the surface tension of the liquid whereas fibers and polymers adjust their molecular orientation towards the direction of the flow in order to suppress the turbulent eddies and thus, reduce friction. The formation of polymer films in the crude oil act as a lubricant with a significant drag reduction. In this case the solubility of the drag reducer in the crude oil is an important factor [204]. When the drag reducer is a polymer, the properties that affect its performance are the high molecular weight, the resistance in shear degradation, the quick solubility in the fluid and its stability against heat,

light, chemical and biological agents. Ultra-high molecular weight polymers that are composed of long chained hydrocarbons are a new kind of drag reducing additive. These hydrocarbons create an intermediate layer between the transported fluid and the inner wall of the pipeline that reduces the energy loss caused by turbulence. Unfortunately, the commercial polymeric additives, mostly homopolymers and copolymers of alpha-olefins, are not yet efficient in heavy oils of low API gravity and high asphaltene content.

On the other hand high molecular weight acrylate-based polymers exhibit a drag reduction percentage ranging from 28 to 36% especially for heavy oils which is a much better performance compared to commercial drag reducers [205]. The soluble drag reducers create films or layers in the crude oil matrix which cause the oil to slip easily achieving higher flow rates at constant pumping pressure. A similar performance is observed when drag reducers/pentanol mixtures are used in extra-heavy oils of 10°API [204]. The problems encountered by the use of these additives include the difficulty of dissolving the polymeric material in the hydrocarbon stream as well as their susceptibility to shear degradation when they are dissolved in the hydrocarbon stream. The latter problem results in their limited use even though the high molecular weight polymers are considered to be the most efficient drag reducers [206].

Surfactants are less efficient in drag reduction than polymers but they are more resistant in shear degradation. Their advantage is that they modify the orientation of their large-scale rod-like micelle structures that promote drag reduction and this orientation is rapidly recoverable even after the flow is disrupted [207]. Drag reduction increases with additive concentration up to a point, beyond which further addition of drag reducer does not have any effect [208].

Mixtures of polymers and surfactants can have a synergistic effect over drag reduction [209]. This is achieved initially by the formation of a polymer film around surfactant's micelle which results in molecular aggregates. These aggregates, under flow conditions, tend to orientate with the flow and, consequently, when the flow rate reaches a certain value (critical point) in which the surfactant induced drag reduction disappears, the drag reduction caused by the polymer-micellar solution remains.

In conclusion, the major problems of using drag reducing additives include the tendency of the additive to separate when it is stored, the difficulty of dissolving the additive into the heavy crude, the resistance in shear degradation and the determination of the additive dosage that will maintain constant pressure drop [160].

Core-annular flow

Another way to facilitate heavy oil pipeline transportation is reducing the pressure drop by developing core annular flow. The main idea of this method is to surround the core of the heavy oil in the pipeline with a lubricating thin film of water (or solvent) near the pipe wall and maintain the pumping pressure close to the one needed for pumping the water (or solvent) alone [210] (Figure 100).

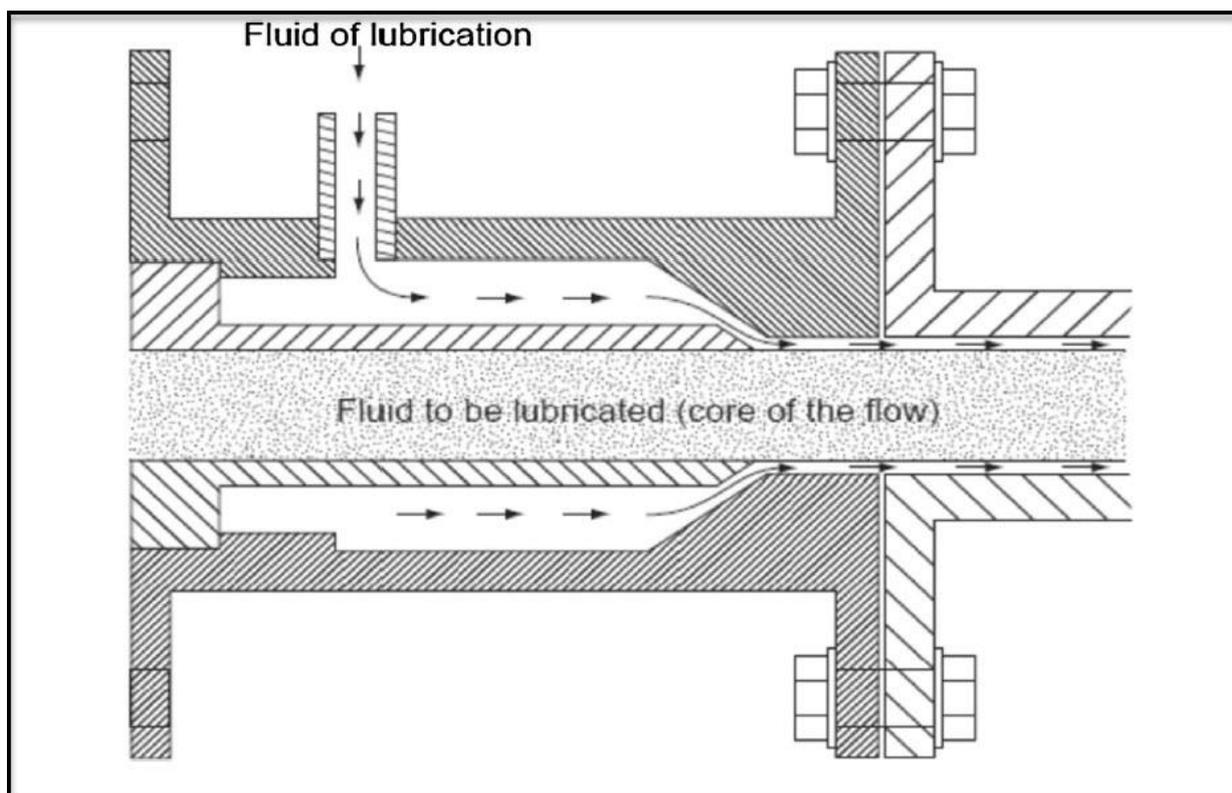


Figure 100: Schematic illustration of the core annular flow configuration, Bensakhria, A., Peysson, Y., Antonini, G., 2004. *Experimental study of the pipeline lubrication for heavy oil transport*. *Oil Gas Sci. Technol.* 59, 523–533, [211].

There are cases where the pressure drop reduction with water lubrication was around 90% compared with that without water lubrication [211]. The required volumetric percentage of water in core annular flow configuration ranges from 10 – 30% [212].

The idea of water lubrication was first proposed in 1904 [213]. Since then the first commercial pipeline that was constructed for core annular flow transportation of heavy oil operated in 1970's by Shell near Bakersfield, California [214]. That was a 39-kilometer pipeline that operated with 30% volumetric water with 24000 BOPD flow rate. Core annular flow method for pipeline lubrication was also implemented in Lake Maracaibo (Venezuela, 35 kilometers long pipeline) and in pipelines of Syncrude Canada Ltd [191].

This pattern of water flowing at the annulus while the heavy oil is the core of the flow is stable if the liquids have similar densities and they are immiscible [215]. In practice, core annular flow conditions are very rare [211]. Unless the densities of the fluids are closely matched, it is more likely that waves are created in the oil-water interface leading to a wavy core annular flow [216]. If the density difference of oil and water is large the oil core will move radially due to buoyancy force and consequently it will be pushed to the wall of the pipeline as it is shown in the following figure.

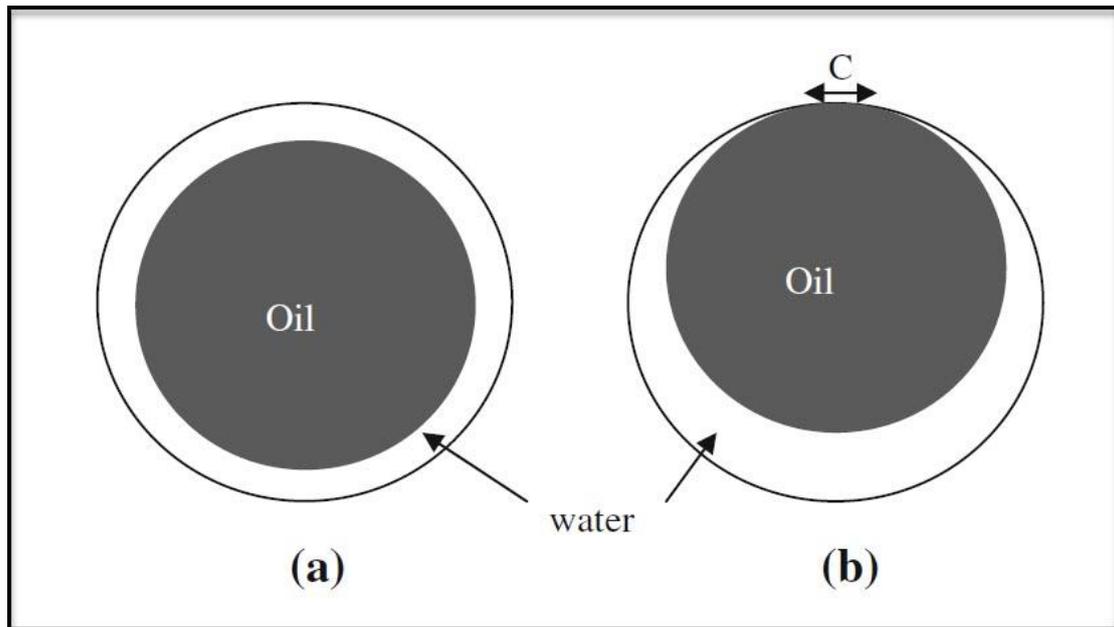


Figure 101: Radial position of the oil core: (a) perfect core-annular flow and (b) with density difference. C is the contact perimeter between the oil phase (core) and the pipe wall, *A review of technologies for transporting heavy crude oil and bitumen via pipelines*, Abarasi Hart, *J Petrol Explor Prod Technol* (2014) 4:327–336, [160].

The hydrodynamic instability of the core annular flow originates from the capillary forces and the difference between the slip velocities of the fluids. The capillary instability is a result of surface tension and density difference between the two liquids which can lead in the breakage of the core flow into slugs. In that case, increasing the velocity of the fluids may enhance core stability whereas lower velocity leads to the stratification of the liquid phases [1].

Commercial application of core annular flow for heavy oil transportation via pipelines encounters many difficulties. The fact that there are strict flow and phase restrictions prevents sharing the pipeline with fluids that are not in the core annular flow regime and lead to the exclusive use of the pipeline by a single production fluid. The tendency of heavy oil to accumulate at the walls of the pipeline is an additional problem. The adherence of oil is gradual and may end up in the blockage of a pipeline. Incrustations in heavy oil pipelines can also be formed because of the reversal of wettability that can be caused by asphaltenes and naphthenic acids. Such effects can be reduced or even avoided by adding sodium meta-silicate in the water phase [217].

Additional problems are the difficulty to maintain the stability of the system over long distances, fouling and corrosion of the pipe walls and difficulties in restarting the flow after unscheduled shut ins [2]. The residence of stagnant fluids during shut in leads to stratification of the phases and the restoration of the core flow regime afterwards is difficult [7]. In order to solve the problem of restoring the core flow regime after a shutdown period a method was proposed that involved different flow onset for each phase and different pressure build up. First the low viscosity water phase is pumped at a gradually increasing rate into the pipeline until it reaches a critical velocity which is needed to develop an annular flow at steady state condition. Then the heavy oil phase is pumped into the pipe and the flow gradually increases but at a lower rate than the water [218].

IN-SITU/PARTIAL UPGRADING

The general idea behind In-Situ and Partial upgrading processes is the optimization of the heavy and extra heavy crude oil properties. The conventional refineries are designed to process light and medium crudes and they are not adequate for processing feedstock consisted exclusively of heavy and extra heavy crude oils. For that reason, upgrading the heavy crudes, such as increasing their API gravity and reducing the viscosity and the content of sulfur, nitrogen and minerals, can significantly enhance transportation and create a feed appropriate for refineries [219]. An additional purpose for upgrading heavy oils is to improve the quality of refinery residues and increase their market value. Lastly, the increasing demand in fuel oil is another reason for processing the residue from the atmospheric or vacuum distillation columns [1].

The upgrading processes aim to increase the hydrogen-to-carbon ratio of the heavy crudes. A number of technologies have been developed which are based either on carbon rejection (non-catalytic) or on hydrogen addition (catalytic) [220] [221].

A schematic representation of the catalytic and non-catalytic routes of heavy and extra-heavy oil upgrading are shown in the following figure.

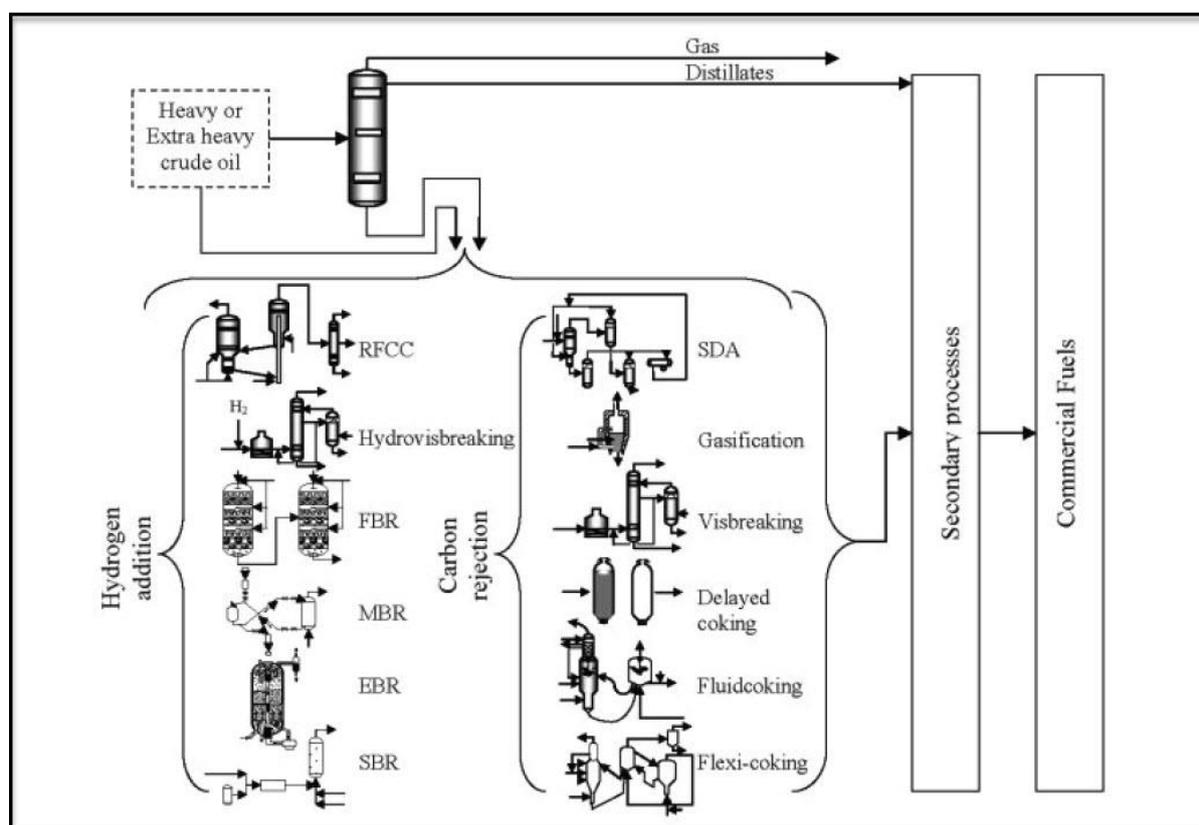


Figure 102: Hydrogen Addition (Catalytic) and Carbon Rejection (non-catalytic) processes for upgrading Heavy oil, Extra Heavy oil and Residua, Future Technology In Heavy Oil Processing, Jorge Ancheyta, Mohan S. Rana, Petroleum Engineering-Downstream, Encyclopedia of Life Support Systems (EOLSS), [221].

Carbon rejection is a type of upgrading in which heavy hydrocarbons and fuels are heated under pressure (Thermal Processes). The method has been applied in the industry since 1913. The heavy hydrocarbons and fuels are heated in large drums until some of the oil components reach their thermal fracture point and yield lighter products with lower boiling points. At the same time other oil components

react in the opposite direction, i.e. forming larger molecules and thus producing significant amounts of coke [219] [220]. This procedure is also known as thermal cracking or coking process and involves hydrogen atoms transferred from the heavier oil molecules to the lighter ones resulting in coke formation. Hydrogen atoms are also transferred from the feed residue to the lighter products at high temperatures, in a process which also results in coke formation.

Carbon rejection processes are Non-catalytic and include the technologies of gasification, delayed coking, fluid coking, flexicoking and visbreaking [222].

- **Gasification (GF)** involves the complete cracking of the residue into gaseous products and solids at temperatures over 1000°C and it produces syngas, carbon black and ash [223]. The integrated gasification combined cycle (IGCC) is an alternative process for heavy residue conversion and power generation with minimum effect on the environment (low SO_x and NO_x). Gasification is mainly used in Europe where coking of residues is not that extensive.

- **Delayed coking (DC)** is mostly preferred from industries due to its flexibility to convert any kind of residue and it involves complete rejection of metals and solid carbon while providing partial conversion to liquid products such as naphtha and diesel. It is more expensive than Solvent Deasphalting (SDA is described later in this chapter), but less expensive than other thermal processes and its major disadvantage is the high coke formation and the low yield of liquid products (Figure 103) [224].

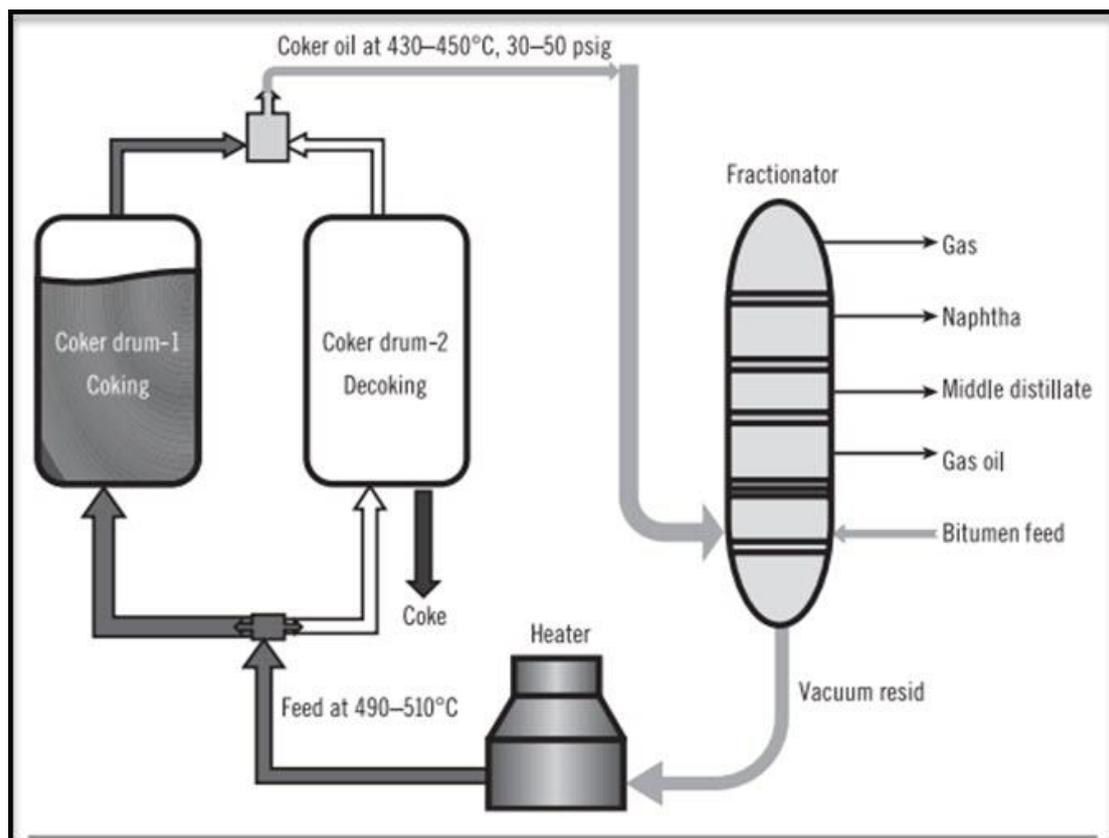


Figure 103: Schematic of delayed-coking technology, Oil sands, heavy oil, and bitumen : from recovery to refinery, Dwijen K. Banerjee, PennWell, 2012, [225].

Delayed coking is a mature technology offered by Foster Wheeler SYDEC (Selective Yield Delayed Coking) [219].

Solvent Deasphalting (SDA) is a Non-catalytic upgrading process that does not belong to Carbon rejection processes. In this case the residue has high asphaltene content which is reduced by extraction. During SDA process a paraffinic solvent (propane, butane, pentane or n-heptane) is added to the feed. The non asphaltene components of the oil are dissolved, while the asphaltenes are insoluble and precipitate. This low cost process is flexible and yields a low contaminant deasphalted oil (DAO) with a low metal content (Ni, V) and rich in paraffins [222].

Hydrogen addition technologies yield more products of higher quality than the carbon rejection processes. The hydrogen is usually extracted from natural gas and is added to the feed in order to reduce the carbon-to-hydrogen ratio [219]. Hydrogen addition technologies are Catalytic processes that have high investment cost and demand for substantial volumes of hydrogen gas [220]. These processes include residue fluid catalytic cracking (RFCC) and hydroprocessing which is further divided in hydrotreating (HDT) and hydrocracking (HCR) processes. Hydrotreating processes are basically hydrodemetallization (HDM), hydrodesulfurization (HDS) and hydrodenitrogenation (HDN) and hydrocracking processes are classified according to the type of reactor in fixed-bed, moving-bed, ebullated-bed and slurry-bed processes [222].

Hydrocracking is the first of the two stages of hydrogen addition process. In this stage heavy oils are heated with hydrogen in high pressures in order to promote molecule rupture and form smaller and simpler chemical structures. Hydrotreating is the second stage where hydrogen is added to promote hydrogenation and remove impurities without breaking the structures [1]. The main hydrogen addition processes are presented briefly below.

- **Residue Fluid Catalytic Cracking (RFCC)** is an established technique that converts the heavier fractions of the crude into a high octane gasoline blending component. This technique offers high gasoline and lower gas yield than hydroprocessing and thermal processes. The catalyst play an important role in the process efficiency. The RFCC technique also works well with other heavy oil conversion processes such as delayed coking and visbreaking [220].

- **Hydroprocessing** (Hydroconversion processes) is the combination of hydrotreating and hydrocracking processes where the feed is treated at low temperatures but at high hydrogen pressure. Some of the hydroconversion processes are non-catalytic like hydrovisbreaking. Hydrovisbreaking is based on visbreaking and involves treatment with hydrogen at mild conditions This process includes three reactors: (i) visbreaking, (ii) demetallization and (iii) hydrocracking and the procedure is mild thermal processing with hydrogen in the first reactor, followed by contaminant removal (particularly metals) in the second reactor and hydrocracking in the last reactor. Besides hydrovisbreaking another extension of visbreaking process is Aquaconversion that involves catalyst in slurry mode and was developed in 1996 by PDVSA Intevap (research and technology center of Venezuela's Petroleum Organization). An alliance between UOP, Foster Wheeler-USA and PDVSA Intevap was formed in order to commercialize the process in which the catalyst can be used as a support or mixed directly with the feedstock. The presence of the oil soluble catalyst and water steam prevents coke formation and sediment deposition that occur during visbreaking (extensive description of this process is given later in the chapter) [222].

- **Fixed-bed catalytic process (FBR)** includes a reactor in which the catalyst is stationary (the catalyst bed is fixed). Reactors can have an upflow or downflow mode of operation depending on the flow direction of the feed. Hydroprocessing in fixed-bed reactors is a well established technique of simultaneous or sequential hydrogenation of hydrocarbon in the presence of sulfided catalyst by reacting with hydrogen. The basic disadvantage of fixed bed catalyst is the poisoning of the catalyst over time. This effect can be minimized with the use of a guard-bed reactor in order to prevent the metal and coke deposition in the downstream reactors [222] [227].

- Moving-bed catalytic process.** The moving-bed catalytic process was developed to overcome the difficulties in the operation of the fixed-bed catalytic technology when concentrations in metals are over 100ppm in the feed. In a moving-bed process, the fresh catalyst is added at the top, and the spent catalyst is withdrawn at the bottom of the reactor. Special equipment is required for the system to operate at high pressure. As the feed becomes heavier and the metal concentration increases beyond 200 ppm, it becomes more difficult to operate the process at higher conversion levels [225]. The main moving bed processes are OCR (Online Catalyst Replacement) developed by Chevron, Hycon process developed by Shell and Hyvahl process developed by IFP.

- Ebullated-bed (EBR) catalytic process.** In the ebullated-bed process the catalyst in the reactor is not fixed and the fine catalyst particles are added from the top of the reactor. The hydrocarbon feed and H₂ are flowing upwards through a catalyst bed which is kept in suspension by the pressure of the fluid feedstock. The catalysts are made of pellets or grains that have a size of up to 1mm in order to facilitate suspension. The upward expansion of the catalyst bed is controlled by the upward flow of the oil. The expanding bed allows the catalyst to be recovered online, without disturbing the process conditions, while the catalyst is regenerated and recycled back; the catalyst is collected from the bottom of the reactor. Fresh catalyst is added to the top of the reactor and spent catalyst is withdrawn from the bottom of the reactor (Figure 105) [228]. Ebullating-bed reactors can upgrade feedstocks with high content of asphaltenes, metals and Sulfur. There are two ebullated-bed processes currently in use: the H-Oil process of Axens/IFP and the LC-Fining process of Chevron Lummus Global [220]. There is also another ebullated-bed process, called T-Star that was designed to treat heavy gas oil; originally developed by Texaco but now licensed by Axens/IFP. T-Star is fundamentally similar to H-Oil [225].

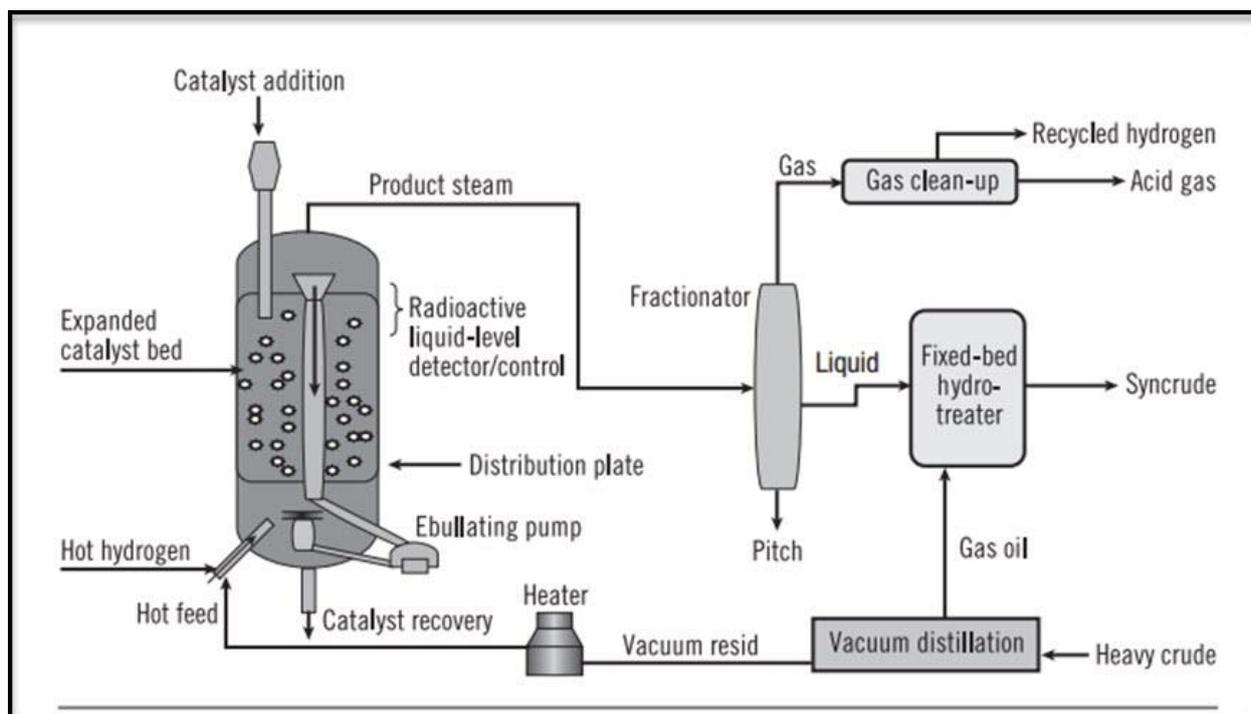


Figure 105: Schematic of ebullated-bed reactor design and process, Oil sands, heavy oil, and bitumen : from recovery to refinery, Dwijen K. Banerjee, PennWell, 2012, [225].

- Slurry-bed processes (SPR)** utilize a disposable low cost catalyst. Its cost is lower than the cost of the fresh hydroprocessing catalysts and that of the spent catalysts. The slurry catalysts that are extensively investigated are typically Molybdenum-based prepared from oil-soluble or water-soluble precursors.

These catalysts possess large areas and have particle sizes of a micron scale that consist of a metal sulfide within an asphaltenic matrix. They can efficiently upgrade low value heavy crudes or residues to products of increased API gravity, lower viscosities and reduced content in contaminants. Transition metal-based catalysts deriving from Vanadium, tungsten, chromium and iron are also considered as potential slurry catalysts (Figure 106) [229].

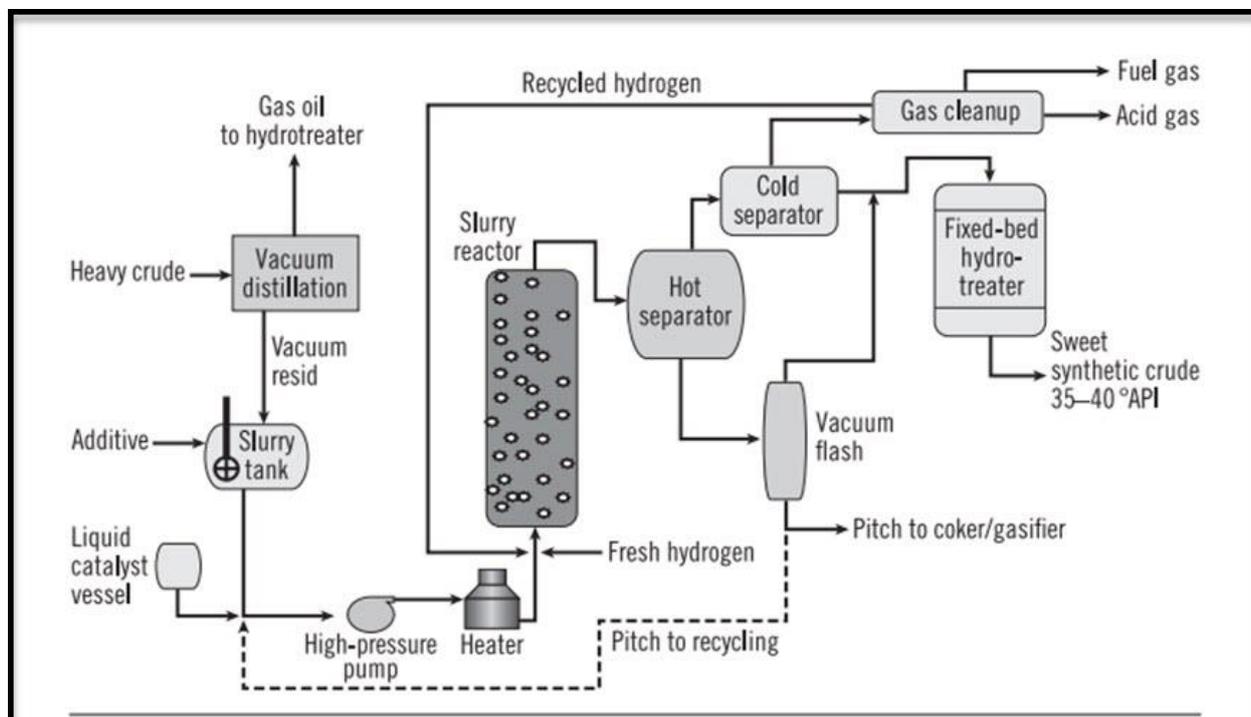


Figure 106: Schematic of slurry-phase upgrading process, Oil sands, heavy oil, and bitumen : from recovery to refinery, Dwijen K. Banerjee, PennWell, 2012, [225]

Hydrodenitrogenation (HDN), hydrodesulfurization (HDS) and hydrodemetallization (HDM) are very expensive processes due to the excessive deactivation of catalysts caused by metal and carbon deposition. On the other hand carbon rejection processes produce large amounts of coke and undesirable products and have low liquid yield. The aforementioned processes for upgrading heavy, extra heavy oils and residues are evaluated by their liquid yield (naptha, distillate and gas oil), the heteroatom removal efficiency (HDS, HDM, HDN) and the quality of the feedstock and degree of conversion [221].

Generally, low-metal content feedstocks are better upgraded by RFCC process whereas high-metal content feedstocks are better upgraded by carbon rejection or hydrogen addition processes. The products from coking base upgraded oil require further hydrotreatment in contrast to the hydroprocessing products. The factors that should be taken into consideration in an upgrading design are the price of crude oil, the level of impurities of the feed, the target of upgraded oil quality, and the process scheme of the refinery where the upgraded oil will be sent [220].

In order to achieve the maximum quality of upgraded products, more the one processes can be chosen. Combining various upgrading technologies is a way to take advantage of the benefits of each approach in an integrated process scheme that may yield higher quality of products than using one single upgrading process. Various integrations of upgrading processes are reported in the literature, which include deasphalting, gasification, delayed coking, RFCC, ebullated-bed process, slurry-bed process and fixed-bed hydrotreating [220].

It has to be noted that since asphaltenes and resins are the main components of heavy and extra heavy oils they significantly affect the choice of the appropriate upgrading method or combination of methods. Two feedstocks of the same API gravity, but of different concentration in asphaltenes and resins will lead to different upgrading processes [1].

In the following table the global commercial processing capacity for upgrading residua is being presented in Mbbbl/day

Technology	United States	Europe	Canada/Mexico/Venezuela	Japan	Rest of the world	World total	% capacity
<i>Carbon rejection</i>							
Cracking/visbreaking	44	2260	331	24	1635	4293	25.85
Coking	2245	673	951	66	1169	5104	30.73
Total	2289	2933	1282	90	2804	9397	56.58
<i>Hydrogen addition</i>							
Fixed-bed, HDT	499	149	30	591	1042	2312	13.92
Ebullated-bed, HDC	102	79	244	23	49	497	2.99
Slurry-phase, HDC	-	-	4	-	-	4	0.02
Residue FCC	831	681	281	318	1832	3942	23.73
Total	1432	909	559	932	2923	6755	40.66
<i>Others</i>							
Deasphalting	283	46	39	16	75	458	2.76
Total	4002	3889	1879	1037	5801	16,609	100.00

Table 14: Worldwide commercial residue processing capacity (Mbbbl/d), Combined process schemes for upgrading of heavy petroleum L.C. Castaneda, J.A.D. Munoz, J. Ancheyta, Fuel 100 (2012) 110–127, [220].

In Situ Upgrading

In Situ upgrading refers to the upgrading of the oil in the reservoir in order to avoid difficulties during heavy oil production, surface processing and transportation. The high capital and operating expenditures of a surface upgrader can be significantly reduced by implementing in situ oil upgrading [230]. In situ upgrading is incorporated with Enhanced Oil Recovery and more specifically with thermal recovery methods such as ISC, SAGD, CSS and THAI or, if a catalyst bed is placed in the reservoir, like the one with the THAI-CAPRI method [160].

In situ upgrading can dramatically reduce the viscosity of heavy oil by thermal cracking reactions and facilitate oil flow and production. It can also improve the oil quality by reducing or removing asphaltenes and resins. Asphaltenes may contain iron, nickel, and vanadium, which are poisons for the catalysts of the refineries. The excess carbon in the form of coke can be left in the reservoir during in situ upgrading. In this way the upgraded oil is able to flow easily to the wellbore, resulting in a higher recovery factor, lifted to the surface and reduce or eliminate the need for diluent in a pipeline transportation scheme [231].

Heavy oil molecules can be broken into lighter hydrocarbon molecules at high temperatures when the vibrational energy exceeds the chemical energy in the C-C bond in the absence of oxygen. In the refinery, steam cracking and thermal cracking occurs at temperatures above 800°C and, even though such high temperatures are not usual in the reservoir, steam and thermal cracking can also occur but at much lower rates.

Steam injection and in situ combustion are ways to heat the reservoir. Steam injection pressures are relatively low because heavy oil reservoirs are usually shallow and thus the maximum steam temperature is limited. For this reason, steam injection is not efficient for significant upgrading in a short time scale. In situ combustion can achieve much higher temperatures than steam injection and that allows for

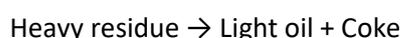
significant upgrading. The upgrading efficiency can also be enhanced by the addition of a catalyst, such as iron, even at the low temperatures of steam injection [232].

Toe-to-Heel Air Injection is a thermal EOR technique (presented analytically earlier in this thesis) that incorporates In situ combustion method with a horizontal producing well. In this case upgraded oil is produced because of thermal cracking in the combustion zone (Mobile Oil Zone, MOZ). The components that react in the combustion zone are water (steam), oil, carbon monoxide and small amounts of unconsumed oxygen. The temperatures at the combustion front are 500-600°C or even higher (peak combustion temperatures of 700 - 800°C have been recorded [233] whereas the pressure is equal to the reservoir pressure, i.e. 30 – 50 bars. Under these conditions the reservoir plays the role of the reactor. The oil upgrading occurs by carbon rejection reactions and it is beneficial because it leads to increased oil recovery, easier transportation and a higher value products with lower sulfur and heavy metal content [234].

The carbon rejection reactions occurring in the combustion zone during THAI method can create the precursor conditions for catalytic upgrading in which a catalyst is wrapped around the THAI horizontal production well in a process known as CAPRI (Catalytic Upgrading Process In Situ). The use of a catalytic bed was also proposed in the conventional ISC method [235] with a vertical production well using a NiMo hydrotreating catalyst but this method faces two important problems: the first is that external heating of the catalyst bed is required and the second is that severe coking can foul the catalyst bed since the combustion front is not constantly in touch with the production well and thus the oil that flows over the catalyst is relatively cold.

In the case of THAI method the combustion zone (MOZ) is constantly in touch with the production well and it provides favourable operating temperatures. The cracking reactions happening in the MOZ along with water (steam) and combustion gases as reactants come in contact with the catalyst layer placed around the production well. In this way the already upgraded THAI oil from the carbon rejection reactions (thermal cracking) is further upgraded by contacting the hydroconversion or hydrotreating catalyst in the annular layer surrounding the perforated horizontal production well. The source of in situ hydrogen is considered to derive from gasification and/or water gas shift reactions [236]. The THAI-CAPRI in situ upgrading mechanism is described by the following equations [237]:

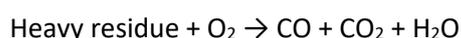
Thermal cracking (Pyrolysis):



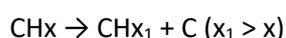
Oxidation of coke (high temperature oxidation):



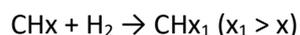
Oxidation of heavy residue:



Carbon rejection:



Hydrogen addition:

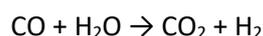


The water gas shift reaction is described by the following simple equations:

Gasification of hydrocarbon:



Water-gas shift:



NiMo and CoMo hydrotreating catalysts have been experimentally studied on heavy crudes from Athabasca oil sands and Wolf Lake heavy oil. API upgrading of 8 and 10.5 degrees respectively have been achieved, together with significant reductions in the content of sulfur and heavy metals, such as Ni and V [238] [237].

However, the THAI-CAPRI method involves significant issues that limit its efficiency. The first issue is that heavy and extra heavy oils have usually a sulfur content of over 5% and high content of heavy metals (Ni, V) that can deactivate the catalyst. The second issue is the coke produced during pyrolysis which can cover the surface of the catalyst layer.

Electrical heating is also an alternative technique for recovering and upgrading heavy oils. Shell has worked on this field since 1980 developing an in situ technique for heavy oil recovery aiming in improved product quality and minimum environmental impact. The method utilizes subsurface heating to convert the kerogen of shale oils into cleaner transportation fuels and gas. In order to minimize a potential pollution of the groundwater, a series of holes is drilled outside the oil zone through which a chilling fluid is circulated in a closed loop. Consequently the rock and the groundwater freeze and create a "freeze wall" which is maintained throughout the project. Afterwards, producer holes are drilled together with holes to house the electric heaters that heat the hydrocarbons to between 288 - 399°C for a period of years. Thermal cracking splits kerogen molecules into smaller molecules which are slowly upgraded by in situ hydrogenation. The upgraded oil and gas are processed at the surface with conventional techniques. Besides the advantages of minimum by-products formation, low water usage, no surface or underground mining and possible recovery from deep deposits, the technique faces the problem of high electricity requirements and long payback time [233].

ExxonMobil has also developed a similar technique that is called Electrofrac and involves electrical heating of an electrically conductive material injected in the fractures of the oil shale and forms a heating element. ExxonMobile is still in research to fully evaluate the operability of this technique.

Partial Upgrading On Site

The upgrading of heavy oil, extra heavy oil and bitumen can also take place partially on site (on the oil field) before the upgraded oil is transferred to the refinery for further treatment. That requires the installation of upgrading facilities integrated with surface facilities near the production wells.

There is a number of partial upgrading technologies that can be applied in the oil field but only some of them have been commercialized. In Appendix D some of the partial upgrading technologies are presented. The first one is Aquaconversion, a technology applied since the early 2000s. Additionally, some technologies that have strong commercial potential over the next few years (based on a study conducted by Kline & Company, Inc. in 2014, [239] are described in detail whereas the others are only mentioned in brief.

CONCLUSIONS

According to recent estimations, the volume of recoverable unconventional oil reserves that consist of heavy oil, extra heavy oil, and bitumen is at least equal to remaining conventional oil reserves. Approximately the 1/4 to 1/5 of this quantity is attributed to heavy oils. Most of the world's heavy oil, extra heavy oil and natural bitumen reserves are located mainly in Canada, in Alberta and Saskatchewan provinces and Eastern Venezuela in Orinoco Oil Belt. Additionally, large amounts of unconventional reserves also exist in the Middle East and Russia.

The geological characteristics, basin types and tectonic activity render the Mediterranean an area prone to oil exploration. Most of the oil fields in Albania were discovered in the Ionian zone which also extends in Western Greece. Geological analogues between Albania and Italy have led to significant discoveries for both sides. In this concept, similar geological traits between the oil producing areas of Albania and NW Greece increase the possibility of an oil discovery in Western Greece.

The prediction of the asphaltene precipitation onset is very important during heavy oil production in order to avoid problems related to asphaltene stability. Besides the asphaltene depositions, further issues that have to be addressed are scale and wax deposition, the formation of gas hydrates in HP, LT environments, naphthenate deposits and naphthenic acids corrosion phenomena.

The tendency of the heavy and extra heavy oils to create stable water-in-oil emulsions during production render the oil dehydration process very challenging. The prediction of the emulsion inversion point is also very important due to the big reduction in the emulsion apparent viscosity when the oil continuous phase passes from oil to water in high water cuts.

The foamy oil phenomenon has a positive impact in heavy oil production because the "solution gas drive" is connected to increases production rates. However, this phenomenon also reduces the capacity of oil and gas separators because it leads to long retention times, liquid carry over and gas carry under.

The formation of a slug flow pattern is very probable to occur during multiphase flow through pipelines especially when high viscosity oils are produced. In fact this probability increases as the liquid viscosity increases for relatively low liquid and gas flow rates. When slug flow appears in horizontal and near horizontal pipelines is called hydrodynamic or terrain induced slug respectively. Slugs can also be created by flow rate changes (turn up slugs) or during pigging. The slug flow characteristics should be studied in order to avoid damages to surface facilities, flooding of separators and high pressure drops.

Many methods have been proposed for heavy and extra heavy oil recovery. Some of them have not yet been applied commercially or are still under study, whereas others are widely implemented. From secondary recovery methods the most common technique is CHOPS. In high viscosity oils, however, the implementation of EOR techniques can seldom be avoided. The thermal EOR methods have been so far the most successful during their long period of utilization. Among the thermal EOR methods the steam methods such as Cyclic Steam Stimulation, Steam Flooding and SAGD are more preferable compared to the In Situ Combustion because they are easier to be controlled. On the other hand, emerging technologies like THAI and CAPRI produce smaller environmental footprint and require less energy compared to SAGD which involves high capital costs and extended steam generation. Additionally, the non-thermal EOR techniques are not widely implemented in heavy and extra heavy oil recovery and especially VAPEX has not yet proved its commercial viability. Future trends seem to converge more on the

integration of EOR methods with in-situ upgrading in order to produce high quality hydrocarbons without the need of diluents or upgraders on the surface.

The disk-stack centrifuges are mostly used in the processing of bitumen froth, whereas in heavy and extra heavy oil processing the emulsion is resolved with a combination of heating, chemical demulsifiers and electrostatic forces. Dilution is mostly used for crude oils with API gravity less than 14. Numerous improvements have been made in the electrostatic treatment since it is considered as the most energy efficient and environmentally friendly path for heavy and extra heavy oil dehydration that reduces the dosage of chemical demulsifiers. These improvements include the application of electrocoalescence in first stage separators and three phase flow conditions. Further research over electrocoalescence is directed to subsea applications with compact and inline solutions such as the in line electrocoalescer.

Pipeline transportation of heavy and extra heavy oils is difficult due to their flow resistance. This resistance has its origin in the composition and for this reason minor modifications in the oil microstructure can drastically improve the flow properties of heavy oils. Partial upgrading on site has received a wide attention and a lot of research but so far the field operations need substantial improvement. Heating and dilution are the most commonly used methods for transporting heavy and extra heavy oils through pipelines. However, the disadvantages of the high heating costs and the low availability in diluents and light oils push the industrial research for alternative methods. Lubricated pipeline methods such as o/w emulsification and core annular flow, are attractive alternatives because they reduce cleaning and maintenance costs and increase the flow rate and energy savings by decreasing the friction losses. In spite of the benefits, these methods must be further evaluated, concerning the emulsifier costs and their hydrodynamic stability for o/w emulsions and core flow respectively.

APPENDIX A

MECHANICAL SEPARATION

Before moving further in this chapter it is useful to clarify the principles under which the separators work. In the term mechanical separation 3 mechanisms are included:

Gravity settling involves the way the dispersed phase, either gas bubbles or liquid droplets, will settle at a certain velocity called terminal or settling velocity. Settling velocity results from equating the gravity force acting on the bubble or droplet with the drag force caused by its motion in the continuous phase [47]. For example, the action of these forces on a liquid droplet in gas continuous phase are illustrated in Figure A1.

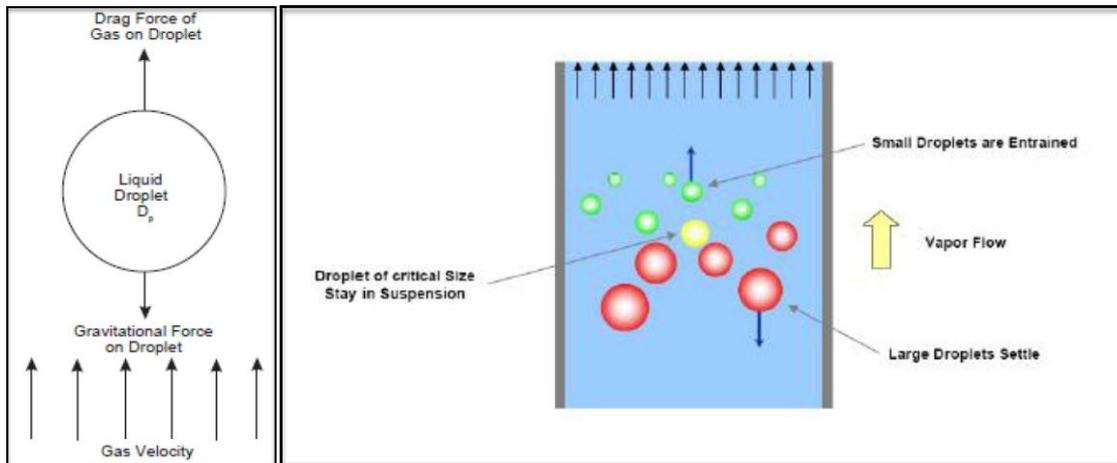


Figure A1: Forces (drag and gravitation) on a liquid droplet in gas stream (left) gravity settling of droplets in gas phase(right), Performance of internals in three phase separators.

When two liquids are immiscible, as in the case of oil and water, they can form an emulsion in which the dispersed liquid forms droplets in the continuous phase. When a water droplet falls in oil continuous phase three forces act on it, F_b , F_d and F_g where F_b is buoyancy force due to fluid displacement from the water droplet, F_d is drag force due to the drag of the oil continuous phase on the water droplet and F_g is the force of gravity (mg) [240] (Figure A2).

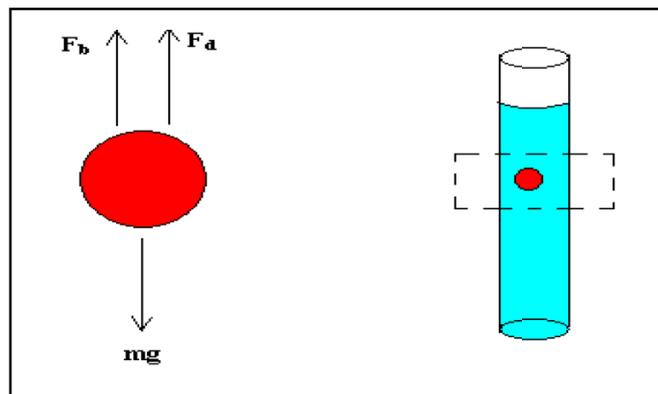


Figure A2: Forces on droplet in stagnant fluid, Vessel Internal Electrostatic Coalescer Technology (VIEC), Ali AlQahtani, QATAR PETROLEUM, SPE 156087, 2012, [240].

In order for the water droplet to reach its terminal (settling) velocity, the sum of the drag and the buoyancy force must be equal to the gravity force, which is shown in the following equation:

$$F_b + F_d = F_g = mg \quad (15)$$

Respectively, the same forces act when an oil droplet rises in water continuous phase in order to reach the oil/water interface. Whether the droplets will rise or fall by their buoyancy is dictated by the density differences of the two liquids. Terminal settling velocity is constant because there are no net forces acting on the droplet [241]. This mechanism of separation by gravity is expressed by Stokes' equation for the terminal settling velocity:

$$V_s = \frac{g(\rho_H - \rho_L)d^2}{18\mu} \quad (16)$$

where, V_s is the settling velocity, g is the gravitational acceleration, ρ_H is the density of the heavy liquid, ρ_L is the density of the light liquid, d is the droplet diameter and μ is the continuous phase viscosity [242]. One can easily understand from Stokes' equation that the time needed for effective separation is increased by higher viscosity of the continuous phase, lower density difference between the phases and smaller droplet diameter [243].

Stokes' equation is valid under the assumptions that the droplets are spherical and that they move in a stagnant liquid. However this is rarely the case when the fluids enter in a separation vessel. In most typical cases the inlet stream is described by turbulent flow (rather than laminar) which can deform the droplets, increase the drag, hinder the settling rate and create eddy currents that will carry the droplets in the wrong direction [242]. The flow regime is described by the flow Reynolds number which is defined as

$$Re_c = \frac{\rho_c V_c d}{\mu_c} \quad (17)$$

where, ρ_c is the density of the continuous phase, μ_c is the dynamic viscosity of the continuous phase, V_c is the continuous phase velocity and d is a characteristic length that in the case of droplets moving in the continuous phase it represents the diameter of the droplet [47] [244].

It is therefore important to sustain a laminar flow regime in the vessels (that happens when $Re < 2300$), [245] and that can be mechanically achieved with the addition of certain internal devices such as plate packs, that enhance gravity settling in the way shown in figure A3.

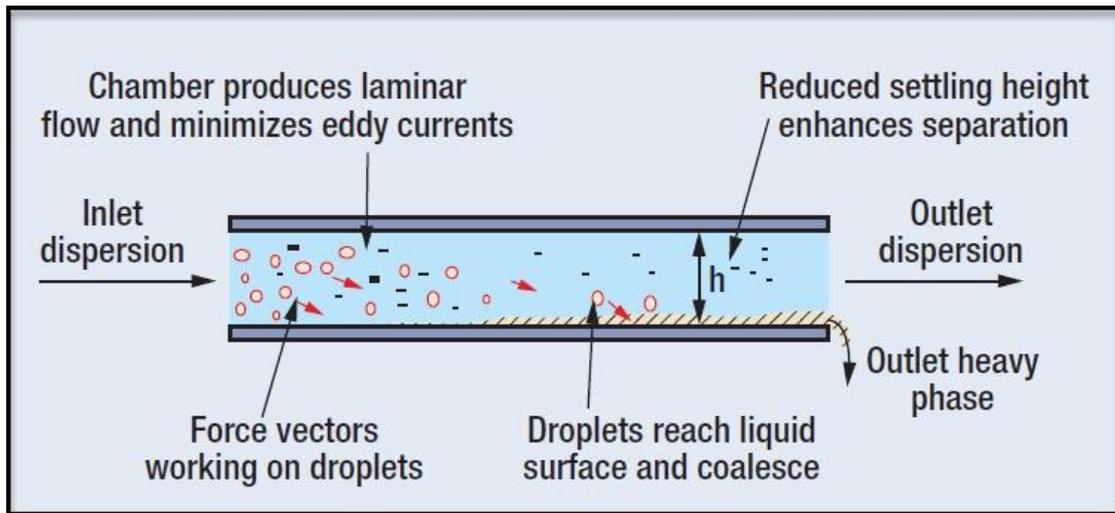


Figure A3: Separation with enhanced gravity settling, *Hydrocarbon Processing, Process and Plant Optimization, Rethink your liquid-liquid separations*, R. Cusack, Koch-Glitsch, LP, Wichita, Kansas, June 2009, [242].

Stokes equation for settling velocity can also be implemented for the settling of small particles or sediment movement under the force of gravity in viscous fluids [240]. In that case, equation (2) can be written as

$$V_s = \frac{g(\rho_s - \rho_l)d_p^2}{18\mu} \quad (18)$$

where, ρ_s is the density of the solid, ρ_l is the density of the liquid continuous phase in which the solid particles settle, μ is the fluid viscosity and d_p is the diameter of the solid particle [246].

One more important parameter is retention time which is defined as the amount of time a liquid remains in the separator at the given separator pressure. The retention time assures that equilibrium between liquid and gas has been reached and for that a certain liquid storage is required. Thus, retention time in a separator is given by dividing the liquid volume inside the vessel by the liquid flow rate [247] as it is shown in the following expression

$$t_r = \frac{V_L}{Q_L} \quad (19)$$

where, t_r is retention time, V_L is the liquid volume and Q_L is the liquid flow rate. If the two liquids are calculated separately a certain amount of oil storage is required to assure that the oil reaches equilibrium and that flashed gas is liberated and an additional amount of storage is required to assure that the free water has time to coalesce into droplet sizes sufficient to fall in accordance with Stokes' equation. Based on settling theory, the design of horizontal separators is determined by a relation between the vessel's length and its diameter. For vertical vessels the settling theory results in a relation for the vessel diameter [47]. The common retention times that are used range from 3 to 30 minutes but they can increase as the oil density or viscosity increases [52]. 2-phase gas/liquid separators are rarely designed to remove droplets smaller than 250 μm [248]. Respectively, 3-phase separators are usually designed to settle water droplets larger than 500 μm and oil droplets larger than 200 μm . In heavy crude oil systems it is sometimes necessary the separator to be designed for settling water droplets larger than 1000 μm . That can result in oil containing as much as 20% to 30% water before it is further processed [52]. In heavy viscous oils, where

the density difference between the oil and water is very small and the oil viscosity is high, the gravity separation of small water droplets from the oil phase is extremely difficult [97].

Centrifugation is a mechanical separation process that is used both in gas/liquid and liquid/liquid separation as well as removing solid particles from the hydrocarbon mixtures. Especially in the case of liquid/liquid separation, when water must be removed from a heavy oil continuous phase, the very small density differences between the oil and water and the high viscosities of the oil continuous phase that can reach over 5000 mPa.s, result in a very slow separation [249]. In addition, heavy crude oils normally contain more solids than lighter crudes, which is appended to the high viscosity of the former that pulls away more solids from the formation during extraction. Centrifugation is more effective than simple gravitational separation due to the increased separation forces.

Centrifugation involves the application of a high centrifugal force by rotation of the liquid, thus creating acceleration by rotation. In Stokes' equation the acceleration caused by gravity is then replaced by centrifugal force:

$$V_s = \frac{d^2(\rho_d - \rho_c)r\omega^2}{18\mu} \quad (20)$$

where, V_s is the settling velocity, ρ_d is the density of the dispersed phase, ρ_c is the density of the continuous phase, d is the diameter of the dispersed phase (either droplets or solid particles), μ is the continuous phase viscosity, r is the centrifugal radius and ω is the angular velocity [131]. This method is based on the density difference between the phases and on the viscosity of the continuous phase, but it is also highly dependent on the intensity of the centrifugal force and the size distribution of the dispersed phase [132]. By replacing gravitational acceleration, g , with a centrifugal force field, the applied g -force can be increased several thousand times and remove droplets and particles up to 10 μ m which could not be achieved by simple gravitational forces.

Centrifugation is implemented through cyclones at the inlet of the separators and at mist extractors. There are also centrifugal inline devices that are installed upstream of the separation vessels to ensure more efficient separation. In cases where the density difference is very small and the continuous phase viscosity high, centrifugal separators are preferred from conventional gravity settlers that must be very large and need long retention times to achieve separation. Especially in offshore heavy oil facilities where the dehydration of the heavy crude is even more challenging due to limited space on the platforms, centrifugal separators are preferred from the conventional ones [131].

Coalescence is the procedure during which two droplets collide and form one larger droplet. As it can be concluded from Stokes' equation, the settling velocity of the droplet is proportional to the square of its diameter [131].

In a liquid-liquid mixture, when two immiscible liquids like water and oil form an emulsion they are not separated by a single interface but one phase is dispersed in the other. The interfacial area between the dispersed droplets and the bulk phase is much larger than when the phases are separated in two continuous phases, thus, the energy of the system is increased w.r.t. the perfectly separated configuration [250] (Figure A4).

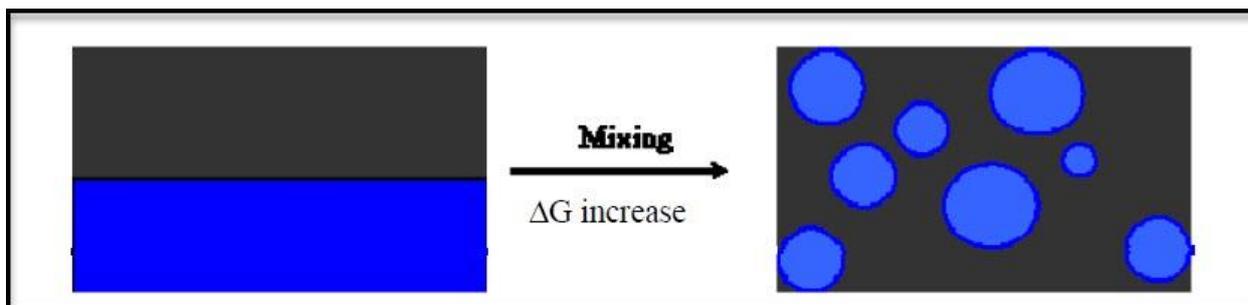


Figure A4: Droplet formation when oil (dark) and water (blue) are mixed. The energy of the system increases (ΔG is positive) as the interfacial area between the two phases is increased.

The increase in the energy of the emulsion compared to the non-emulsified components can be considered as a measure of the thermodynamic instability of the emulsion [251]. Because of their thermodynamic instability, emulsions will seek to minimize the surface area by separating back into the different phases, thus decreasing the interfacial area and the Gibbs free energy. For an emulsion to separate, the droplets must merge with each other. Processes that facilitate separation are sedimentation/creaming, flocculation and coalescence (Figure A5) [252] [253].

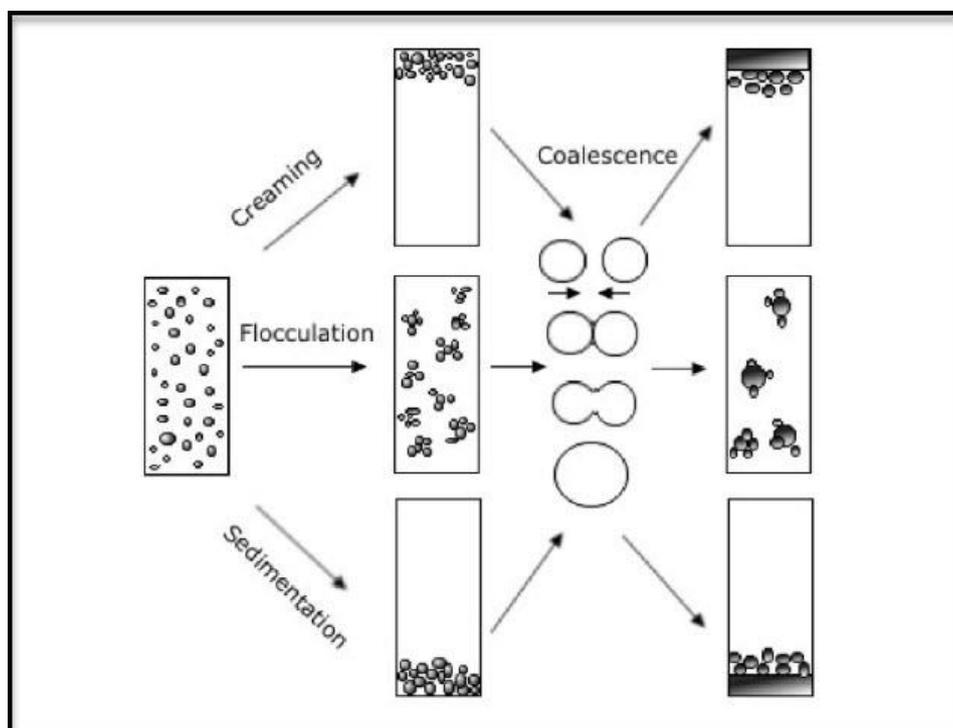


Figure A5: Oil droplets will cream and water droplets will sediment. Free water will exist at the bottom and free oil at the top, Aulfem, I. H. Influence of asphaltene aggregation and pressure on crude oil emulsion stability. Department of Chemical Engineering, Norwegian University of Science and Technology, Trondheim, 2002, [253].

Creaming and sedimentation create a denser droplet concentration, which results in the close packing of the droplets. When the droplets approach each other the coalescence occurs in two stages. The first stage involves the thinning or drainage of the film that separates the droplets in order to reduce the distance between the interfaces. The second stage involves the film rupture, when the film reaches its critical thickness [113].

Coalescence can be enhanced by inserting solid structures that act as nucleation sites in the liquid collection section of the vessel. This way, the droplets collide with the solid, adhere to its surfaces and consequently coalesce into larger droplets that settle [242].

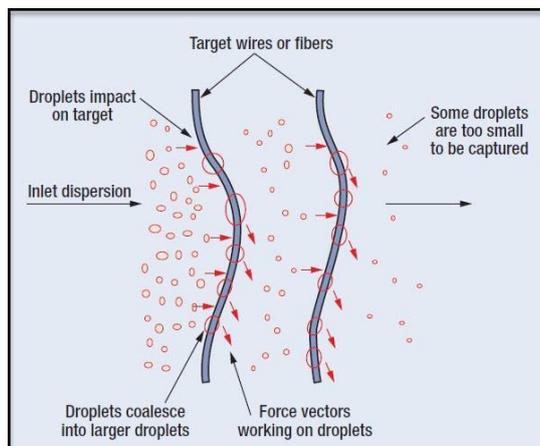


Figure A6: Coalescing process using target wires or fibers, *Hydrocarbon Processing, Process and Plant Optimization, Rethink your liquid-liquid separations*, R. Cusack, Koch-Glitsch, LP, Wichita, Kansas, June 2009, [242].

Such coalescing targets are plate/vane packs, matrix packs with corrugated metal sheet packing [254]. The materials that are used in coalescing devices to coalesce water droplets are hydrophilic materials, for example stainless steel or fiberglass [242]. That is because studies have shown that the packing material of the coalescing device must have the same wettability as the dispersed phase [255].

APPENDIX B

SEPARATOR INTERNALS FOR 2-PHASE AND 3-PHASE SEPARATORS

The interior of the separators contains several types of internal devices that are common in horizontal and vertical configuration. These internals serve different purposes depending on which part of the separator they are installed. There are four main sections in every separator and these are

- Inlet diverter section
- Liquid collection/coalescing section
- Gravity settling section and
- Mist extractor section

[108], [98], [109]. The internals that promote gas/liquid separation are common in 2-phase and 3-phase separators whereas the internals used for liquid/liquid separation are used only in 3-phase separators. In this chapter, the internals that are used in two and three-phase separators are presented.

The inlet diverter section is the part of the separator where the inlet stream is subjected to an initial bulk separation. The fluids enter the separator as a mixture of oil gas and water in turbulent flow with high velocity and therefore high momentum. The inlet diverter mounted in the separator interior changes suddenly the direction of the inlet stream and the density difference between the phases causes the liquid droplets to fall and the gas bubbles to rise. An effective initial separation of the phases can improve the overall separation efficiency. There are different kinds of inlet devices and based on their mode of operation they are separated in the following main categories: Diverter plate, Half pipe, Inlet vane distributor and Inlet cyclone.

- Diverter plates or baffles can have the form of a flat plate, a spherical dish, a cone or whatever can cause an abrupt change in flow direction when the fluids enter the separator [52]. Although the fluids enter the vessel with the same velocity their density is not the same and consequently their momentum is not the same [47]. The result of this is that the liquids with higher densities than the gas cannot change direction so easily, fall on the diverter and then are drained to the bottom of the vessel whereas the gas flows around the diverter. These inlets are preferred when gas load is low and the momentum of the fluid is not high. In case of high fluid momentum, diverter inlets can result in droplet shattering and foaming. In the Figure B1 below a plate inlet deflector and a tangential baffle are shown.

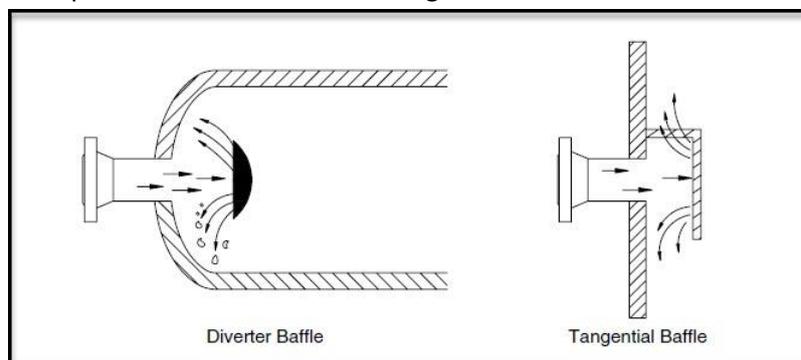


Figure B1: Two different types of diverter plates, Surface production operations, Ken Arnold, Maurice Stewart, 2008, [52].

- A half pipe inlet device is actually a cylinder horizontally oriented and enters either from the side or the top of the vessel (Figure B2). A bottom or a top part of the cylinder is removed lengthwise and in this manner the fluid is directed to the down part or the upper part of the vessel [256]. Half pipe inlets are an economical piece of equipment but they have the disadvantage that can cause short-circuiting or channeling of the fluid [47].

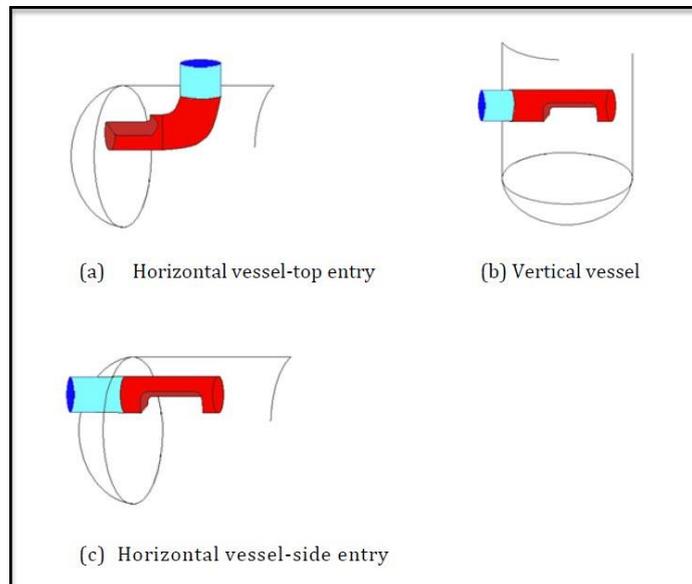


Figure B2: Half pipe configuration in horizontal and vertical vessels, Saeid, Rahimi, Three Phase separator –Gas internals, 2013, [256].

- Inlet vane distributors are more elaborate inlet devices than the aforementioned. The incoming fluid is separated to either side of the device which consists of an array of curved vanes that release the gas and liquid into the separator and distribute the phases. Inlet vane distributors are used when inlet momentum is roughly between 2500-8000 kg/ms² [257] and the simplest form of a vane inlet is the dual vane inlet, or else Bifurcator inlet (Figure B3).

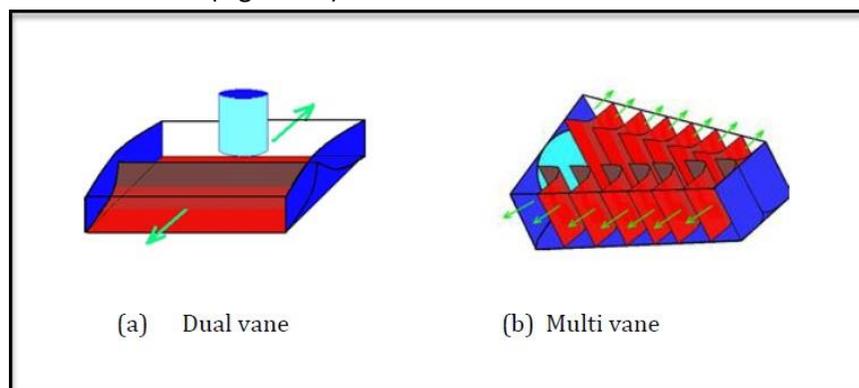


Figure B3: Schematic of dual vane and multi vane devices, Saeid, Rahimi, Three Phase separator –Gas internals, 2013, [256].

Inlet vane distributors achieve uniform flow distribution with low shear and pressure drop. They can be used when solids are produced or foaming is expected and they can handle liquid slugs effectively [47]. Droplet shattering is minimized and agitation is reduced compared to simple deflectors. When the gas flow is relatively higher than the liquid flow the multi vane device is preferred due to its excellent vapor distribution.

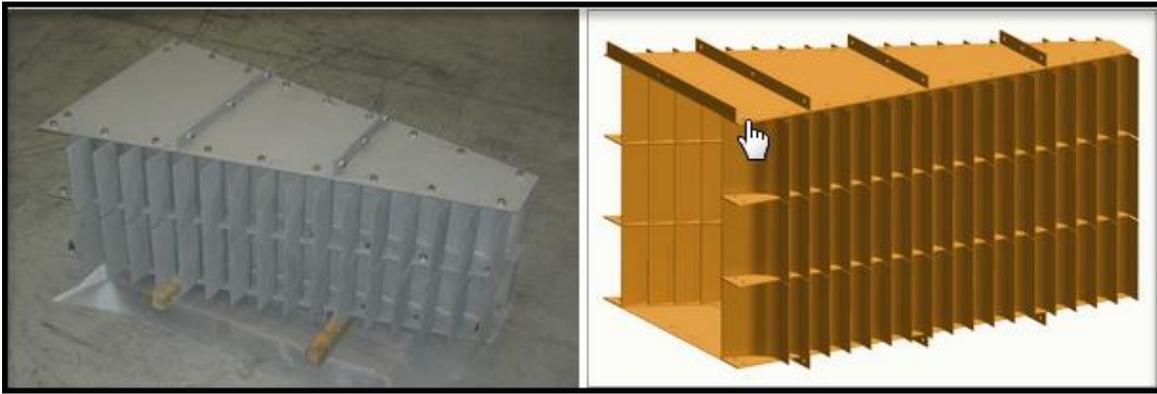


Figure B4: inlet vane distributor, Source, www.afptech.eu

- Inlet cyclones have the most complex design of all inlet devices but they also have the best performance. They take advantage of centrifugal forces to achieve phase separation and they are used when there high momentum dissipation is required, usually over $8000-9000 \text{ kg/ms}^2$, foam and emulsion reduction [47]. Solids can also be handled effectively by an inlet cyclone. The feed stream is brought into rotation by a spin device and the resulting centrifugal force moves the denser phase (liquids and solids if present) to the wall of the cyclone where it is drained out from the bottom of the device into the liquid compartment of the separator. The lighter gas phase is forced out from the top of the cyclone.



Figure B5: Inlet cyclone with turbine blades, internal vortex tube, gas blocking plate and liquid antispin blades (Photo courtesy of FMC Technologies), *Savvy Separator: Design of Cyclone Separators: Internals and Liquid Levels*, Robert Chin, *Padden Engineering*, 31 May 2015, *OGF Vol.4 Issue 3*, [258].

Cyclonic inlet devices can be used in low and high gas/oil ratios and they are the only type of inlet devices that is partly submerged in the liquid compartment of the separator. Their disadvantage is that they are sensitive in flow rate fluctuations and at low velocities they don't work properly. As a result, they are not chosen when the flow rates are not expected to be steady [52]. The design of the inlet cyclones has evolved from single or double to multi-cyclone arrangement.

In the following Table B1 the aforementioned devices are presented in summary [257]:

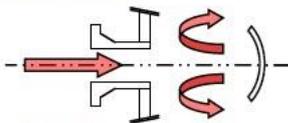
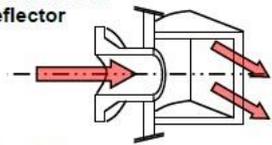
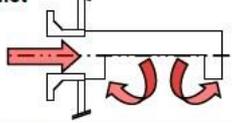
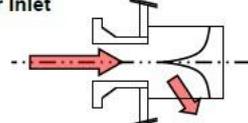
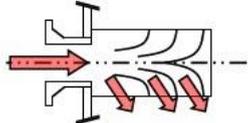
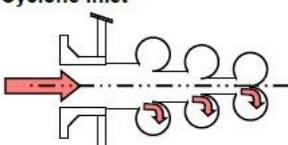
No Inlet Deflector	With generous vessel sizing it is acceptable to avoid the use of an inlet deflector or distributor if the incoming fluid momentum is low.
Plate Inlet Deflector 	Basic, economic momentum breaker used for simple 2 phase (liquid + light gas load) inlet streams with a horizontal entry nozzle in the vessel head. Avoid use where foam or oil/water mixtures are possible, or where the vessel is not generously sized.
Tangential or Box Inlet Deflector 	Standard, robust device used for many 2 phase (gas/liquid) inlet streams. Tangential arrangement on vertical vessels or Box arrangement on horizontal drums. Do not use where foam is predicted. A sloping baffle plate is often located below the device to prevent the swirling action from disturbing the liquid.
Half-Pipe Inlet 	Standard, economic distributor used for simple 2 phase (gas/liquid) inlet streams with a horizontal entry into a vertical vessel. Use where liquid load is very low in comparison. Not as efficient as other devices so needs larger clearances to other internals.
Bifurcator Inlet 	A robust, device used for many 2 or 3 phase (gas/oil/water) inlet streams with high liquid content into a vertical or horizontal entry nozzle in the vessel. Do not use where foam is predicted, or where gas load is high, unless the separator gas space is adequate.
Multi-Vane Inlet 	A more complex diffuser used for 2 or 3 phase inlet streams with a horizontal entry nozzle in the vessel. Use where gas load in relation to liquid load is high for the vessel size. Use caution where foam is predicted unless the separator gas space is adequate. Multi vanes enable more compact vertical layouts to be used.
Multi-Cyclone Inlet 	A sophisticated device used for 2, 3 or 4 phase (gas/liquid or gas/liquid/solids) inlet streams with a horizontal entry nozzle in the vessel. Use where foam is predicted, or where momentum is high. Use with caution on FPSO's as the liquid level (seal) needs to be reliable. Effective as sand separators.

Table B1: Summarization of the described inlet devices, Courtesy of HAT International

A comparison of the inlet devices is presented in the following table where it is obvious that the inlet cyclone has the best performance of all the other options.

Ability to	TYPE OF INLET DEVICE				
	None	Inlet Vane	Cyclone	Half pipe	Baffle
Momentum reduction	Poor	Good	Good	Good	Good
Bulk separation	Good	Good	Good	Average	Poor
Prevent re-entrainment	Good	Good	Average	Average	Average
Prevent liquid shatter	Good	Good	Good	Average	Poor
Low differential pressure	Good	Good	Average	Good	Good
Not create foam	Poor	Average	Good	Poor	Poor
Ensure good gas distribution	Poor	Good	Average /poor	Poor	Poor

Table B2: Comparison of inlet devices, NORSOK STANDARD P-100 Rev. 2, November 2001, [259].

The Liquid collection/Coalescing section is located at the bottom of the vessel and must provide the required retention time for the gas bubbles to rise and meet the gas/liquid interface. Furthermore, it must provide the required retention time for the oil droplets to rise through the water and settle in the oil continuous phase and the water droplets to fall and settle in the free water phase [52]. The degree of separation depends on the retention time which on its turn depends on the amount of liquid the vessel

can hold and the flow rate at which the liquids enter the separator. The internals that are placed in the liquid section of the vessel are the distribution baffles that ensure uniform distribution of the liquids and laminar flow conditions, the wave breakers and the coalescing devices to promote oil/water separation.

- The distribution baffles calm the inlet zone and redistribute the liquid flow in such way that turbulence is avoided and laminar flow is ensured [260]. They are also used for surge suppression. Distribution baffles can be either perforated or slotted (Figure B6), but in the case of heavy oil processing slotted baffles are preferred. That is because slotted baffles distribute flow equally and more efficiently compared to perforated baffles and the risk of fouling is greatly reduced [98].

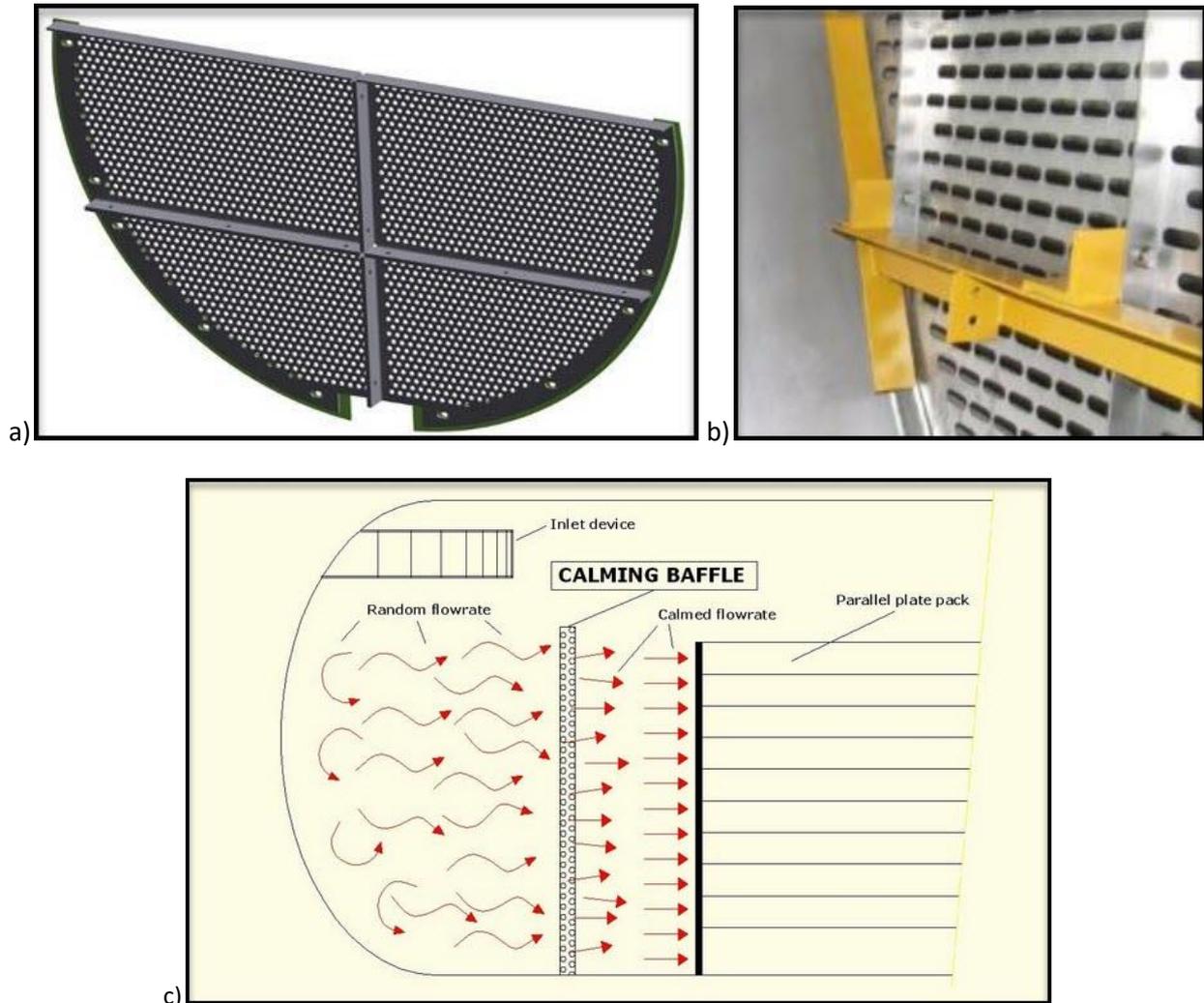


Figure B6: (a) perforated baffles (source www.afptech.eu), (b) slotted baffles (courtesy of Fjords Processing), (c) Effect of distribution baffles in a separator (source www.afptech.eu).

- The wave breakers are also perforated baffles that are located perpendicular to the flow in the liquid collection section of the vessel and they are used mostly in floating structures. Their purpose is to dampen any wave action of the incoming liquids caused from liquid surges, especially on floating platforms where internal waves may be the result of the motion of the platform (Surface production operations, Ken Arnold, Maurice Stewart, 2008). Wave breakers are necessary because the level of the liquids in the vessel must remain constant for the level controls and switches to work properly.

- Coalescing internals are needed in the liquid section of the vessel to aid in coalescing of droplets when gravitational settling alone is not effective. Gravitational settling becomes increasingly difficult as

the droplet size of the dispersed phase decreases. The settling process can be enhanced considerably by passing the dispersion through a suitable coalescer pack. Generally there are three kinds of coalescing internals.

a) Flat plate packs (or vane packs) that take advantage of the short settling distance between the plates and according to Stokes' equation they enhance gravity settling. Depending on the manufacturer, the spacing between the plates can be from 10 to 100mm and the inclination of the plates that are arranged in the axial direction of the flow can be from 45° to 60° [261]. The inclination of the plates is connected with the fouling tendencies of the stream and also ensures separation of the solids in order to avoid plugging [262]. Flat plate packs can remove down to 100µm droplets and they are suitable for heavy solid environments.

b) Matrix packs consist of corrugated sheet metal structure packing and they are mounted on the direction of the flow. They can remove down to 50µm droplets and they are effective in light fouling services [262]. In some cases they can be used as wave and foam breakers [263]. Generally in the case of heavy oil processing, the matrix packs are avoided due to the increased risk of fouling. They have a tendency to clog in the presence of rag layers (layers of unresolved emulsion and solids). A clogged matrix pack will lead to reduced performance and the necessary cleaning will result in lost production time [98]

c) Dual media knitted mesh pads are used in clean and non-fouling liquid services. They can remove droplets as small as 20µm. Separation is achieved by impact of droplets on a wire mesh of co-knitted oleophilic/oleophobic materials [261], [263].

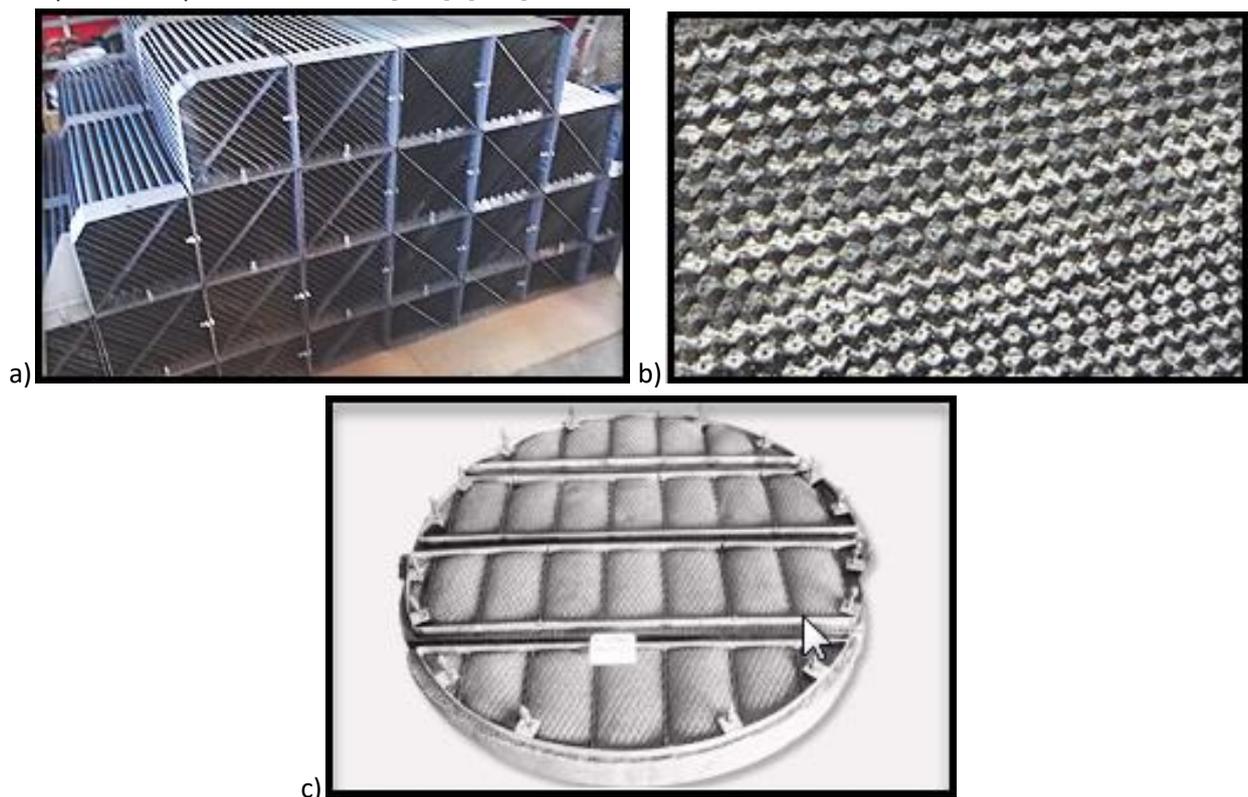


Figure B7: (a) plate packs, (b) matrix packs and (c) dual media coalescers, source: www.fenix.in/liquid-coalescer.html.

The gravity settling section is the part of the vessel where small liquid droplets entrained in the gas stream are separated by gravity and fall to the gas liquid interface. Gravity settling section is designed to remove droplets down to 100µm.

In the gravity settling section of the separators the internals that are mounted are flow straightening, foam-breaker vanes. These de-foaming internals are necessary because heavy oils have usually foaming tendencies that may reduce separation performance [47]. When a foamy crude is present, the retention time can be up to four times longer than the non-foamy case, thus the use of such internals will lead to separation improvements and reduction of the required de-foaming chemicals [52].



Figure B8: Defoaming internals, source: www.fenix.in/liquid-coalescer.html

The mist extractor section is used to remove the droplets less than 100µm that were not removed by gravity from the gas stream. In this section coalescing elements or cyclones are used to increase droplet size and enhance separation by gravity. The selection of a mist extractor is based on the droplet size that must be removed, the pressure drop tolerance, the presence of solids and the risk of clogging and the liquid capacity of the separator. Mist extractors' operations are usually based on a design velocity and depend on the type of the demister and the manufacturer. The designed velocity is given by:

$$V_d = k_d \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \quad (21)$$

where: V_d is the design maximum velocity, ρ_L is the density of the liquid ρ_G is the density of the gas and k_d is the design factor for demisting internals that depends on several factors such as liquid viscosity, surface tension, liquid loading and operating pressure. K values normally range between 0.09 m/s to 0.3 m/s [47]. Standards, like the NORSOK standard, give guidelines for the K-value in different devices.

- Wire mesh is the most common impingement type of mist extractors. The surface of the wire mesh on which the liquid droplets impinge is created by knitting wire together to a pad. The mesh pad is located right before the gas outlet of the separator. As the vapor that carries the liquid droplets passes freely through the knitted mesh the inertia of the liquid droplets does not let them change direction quickly, so they contact the wire surfaces, coalesce and finally drain as large droplets [47]. Due to their dense structure, mesh pads are not suitable for low viscosity liquids because they may get clogged. They are also not recommended when solids are present. A typical k value according to the NORSOK STANDARD for the efficient operation of a wire mesh demister is generally below 0.1 m/s. For mesh pads made from metal material the typical minimum droplet removal size is 10µm and for mesh pads made from plastic or fibre, the minimum droplet removal size is 3-5µm.



Figure B9: wire mesh mist extractors, left: source:www.fenix.in, right:Courtesy of ACS Industries, LP, Houston, Texas

- The vane type mist extractors are widely used in oil and gas separators. The design may vary but the most usual configuration is the one with single and double pocket vanes. The operating principle is to lead the gas stream through a number of parallel vane plates and force it to change direction several times. The mist droplets that are carried with the gas stream cannot follow the abrupt changes in direction and are collected on the vane plates. The coalesced liquid film is then drained through hooks (single pockets or slits (double pockets) as it is shown in Figure B10.

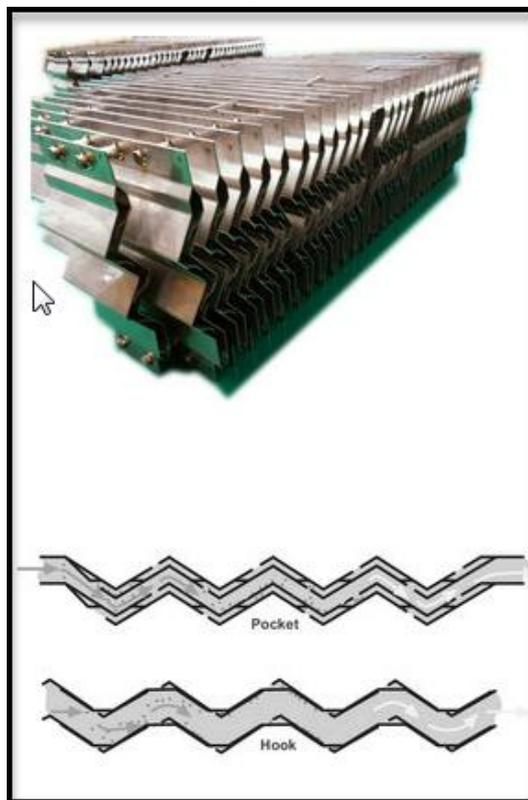


Figure B10: vane mist eliminator, single and double pocket configuration, source: www.saiptech.com

The main difference between the single and the double pocket design is that the coalesced liquid is not in contact with the gas flow so there is smaller possibility for re-entrainment than there is in the single pocket design. That allows the gas in the double pocket configuration to have more than double velocities than in the single pocket design [264]. The vane arrays can be mounted both horizontally and vertically but higher gas velocities can be handled if the vanes are installed in a horizontal gas flow, instead of vertical up-flow. That is because in the horizontal configuration the liquid can easily drain downwards due to gravity and move out of the way of the incoming gas, which also minimizes the chance of re-

entrainment [52]. In the following figure the horizontal and vertical double pocket configuration is demonstrated.

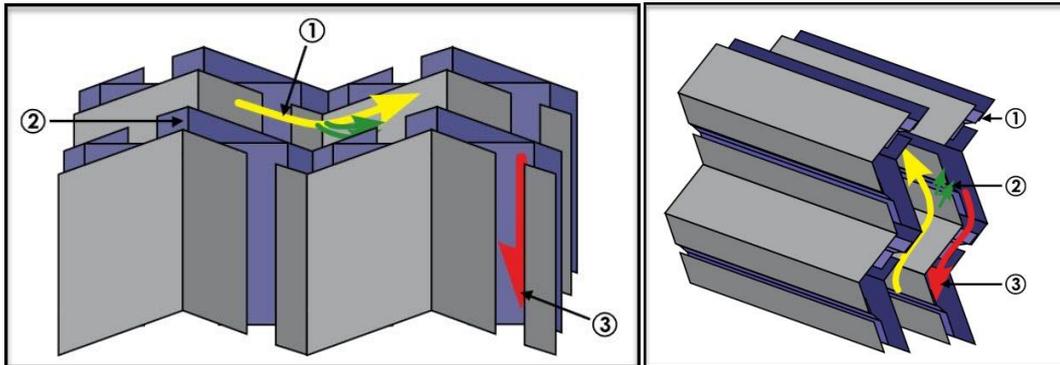


Figure B11: From the left to the right: Horizontal and vertical gas flow. 1. Double pocket hooks direct the collected liquid (green arrow) away from the main gas stream (yellow arrow). 2. The collected liquid (green arrow) flows into separate channels (purple). 3. The separate channels move the liquid away from the gas (red arrow). Koch-Glitsch-LP, Mist elimination. Koch-Otto York separations technology, www.koch-glitsch.com, [264].

Vane mist extractors are mostly preferred when the liquid entrainment has solid content or when there is high liquid loading. They are less efficient in removing very small droplets compared to wire mesh or micro fiber mist extractors and their standard design is generally for removing droplets larger than $40\mu\text{m}$. However there are designs that provide droplet removal down to $8\mu\text{m}$ in diameter.

- The cyclonic demisting device consists of multiple cyclone tubes are mounted on a deck or into a housing as it is shown in the following Figure B12

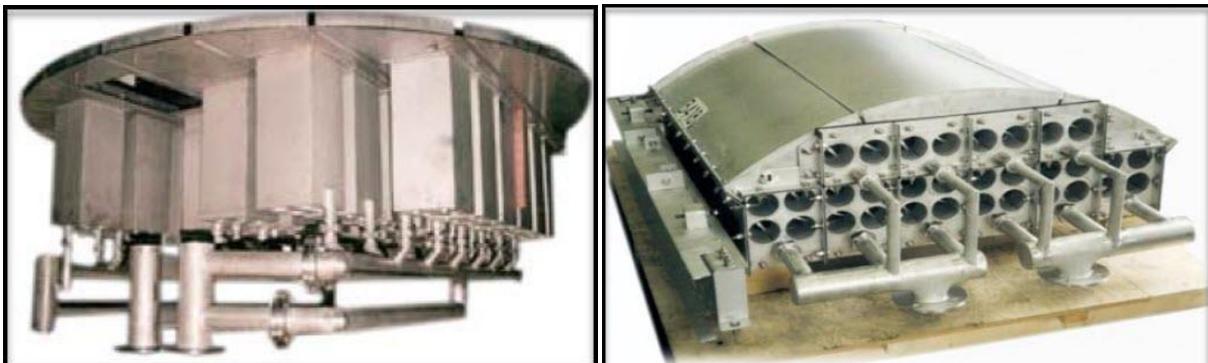


Figure B12: (left) Cyclone mist eliminator assembly with mounting deck and liquid drain piping for a vertical vessel, (right) Cyclone mist eliminator assembly including housing and liquid collection piping, Picture courtesy of CDS separation technology

Cyclone demisters can handle high gas and liquid capacities and offer efficient droplet removal down to $5\mu\text{m}$. Additionally, the cyclone demisters enhance foam breaking, they are not susceptible to fouling and offer high efficiency under high pressures [47]. Gas and mist enters the cyclone and goes through a swirl element. The high centrifugal forces cause the liquid droplets to move outwards and coalesce to a liquid film on the cylinder wall. The liquid is purged through slits, together with some gas, into a chamber where the phases are separated. The purged gas, with some remaining mist, is led to the low pressure zone of the cyclone where the remaining entrainment is removed. The main gas flow is discharged at the top of the cyclone while the liquid is drained at the bottom (Figure B13).

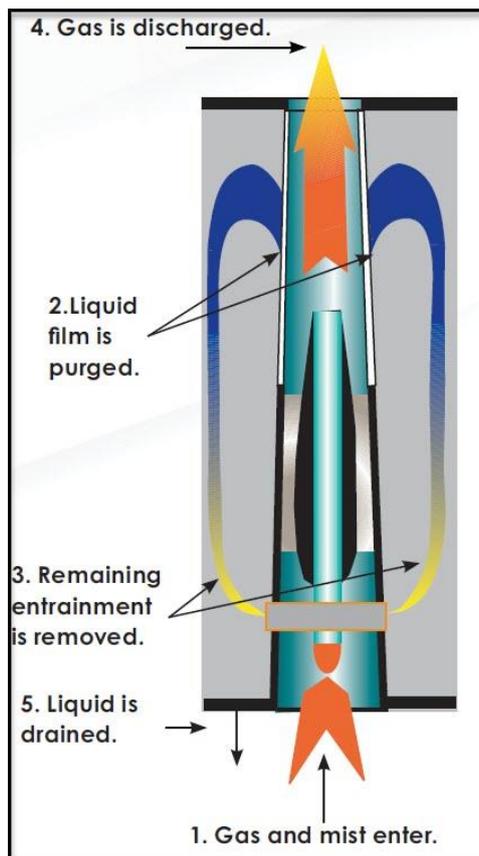


Figure B13: Cyclone mist eliminator, Koch-Glitsch-LP, Mist elimination. Koch-Otto York separations technology, www.koch-glitsch.com, [264].

Cyclones can be of horizontal or vertical configuration. Although they require careful design, they are expensive and sensitive to changes in the flow rate, their numerous advantages make them suitable for upgrading existing vessels or designing more compact new vessels.

The following table compares the different types of mist extractors according to their characteristics [264].

Table 1. Selection Guide - Common Mist Elimination Equipment				
	Knitted Mesh	Vane	Fiberbed	Cyclone
Cost	1	2-3	10	3-5
Gas Capacity	5	6-15	1	15-20
Liquid Capacity	5	10	1	10
Particle Size (micron)	3-10	10-40	< 0.1	7-10
Pressure Drop, WC	< 25 mm (1")	<10-90 mm (0.4" - 3.5")	50-500 mm (2" - 20")	200-240 mm (8" - 14")
Solid Handling	3	10	1	8
Relative scale based on 1 as the lowest. Others are scaled.				

Table B3: Selection guide from a mist extractor vendor, Koch-Glitsch-LP, Mist elimination. Koch-Otto York separations technology, www.koch-glitsch.com, [264].

Other Internal devices that are used in separators are:

- The vortex breakers that prevent the gas from exiting the vessel from the liquid outlet. This can happen when liquid vortices or whirlpools form near the separator's outlet. A vortex breaker is a cylinder with radially directed flat plates. These plates prevent any circular motion that may create a vortex as the liquid exits from the bottom of the vessel (Figure B14)

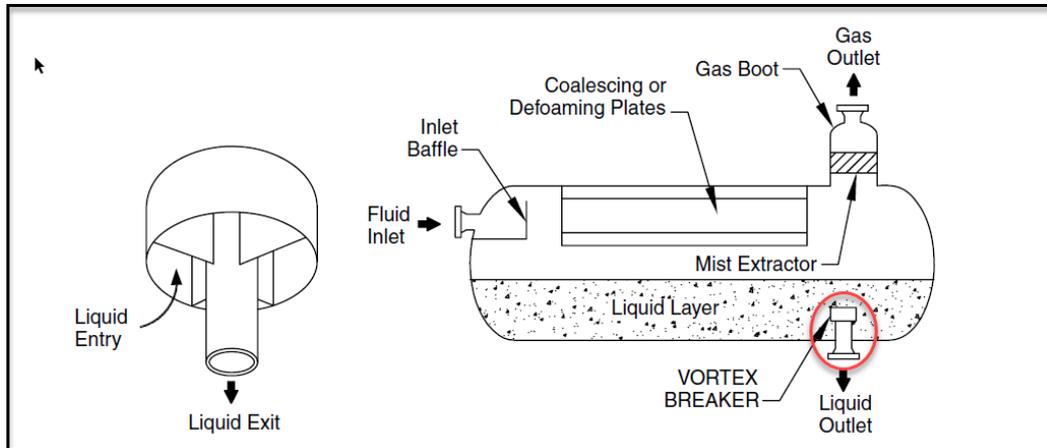


Figure B14: Schematic of a vortex breaker, *Surface production operations, Ken Arnold, Maurice Stewart, 2008, [52].*

- Solids removal systems are used in horizontal separators, where the accumulation of solids can create major problems. High viscosity and low API gravity crudes can carry significant amounts of formation solids [106]. The solids will build-up inside the vessels reducing the available space, or create solid deposits in the internals of the separator, which are difficult to be removed. Fine solids can stabilize the water-in-oil emulsion and result in the build up of an emulsion layer [105]. Special equipment is necessary when the solid loading is severe (Figure B15) such as:

- Sand jetting/mud wash systems
- Interface draw-off systems for draining the interface rag layer to a lower pressure system
- Mozley fluidizer system for draining the bottom of the vessel in a circular area between 24" – 32" diameter around an inserted nozzle.

The most common system is the sand jetting system. A series of mud wash drain valves that open up in a controlled way, discharge a mixture of high-pressure fluid with suspended solids. The nozzles are placed in such a manner that the jets can cover the entire bottom of the vessel. Clogging of the sand drains from the settled solids is prevented by sand pans at the outlets [52].

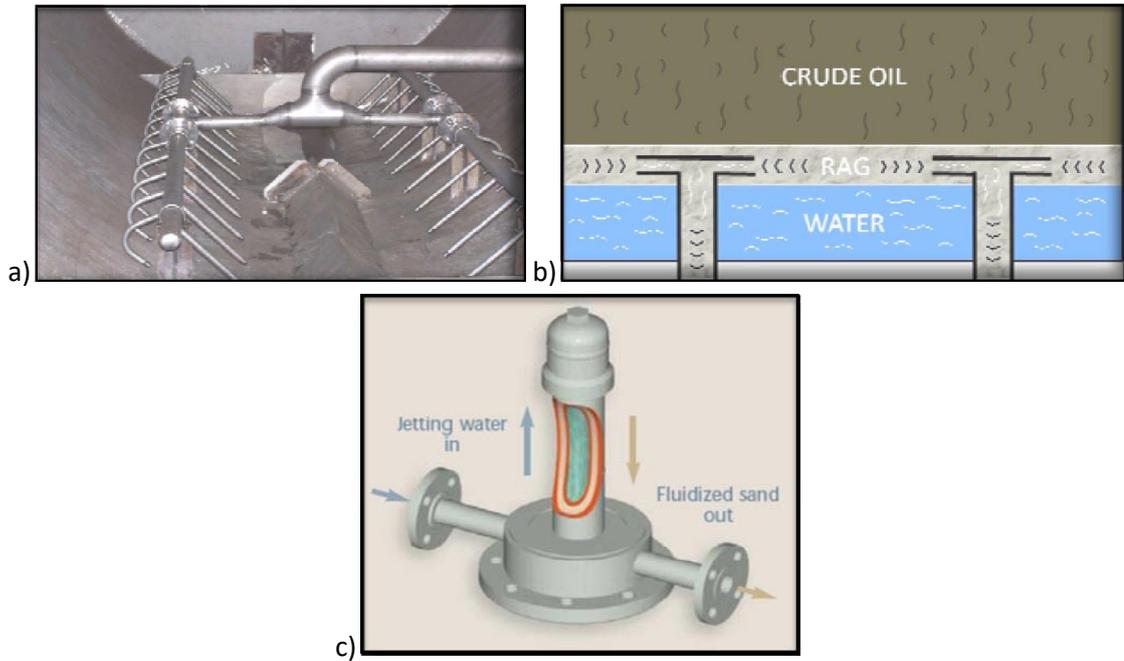


Figure B15: a) Sand jetting/mad wash systems, b) Interface draw-off systems, c) Mozley fluidizer system, *Effective Dehydration of Canadian Heavy Crude Oil and Dilbit*, Erik Sellman, SPE, Terry Murtagh and Benjamin Manuel, Cameron Process and Compression Systems, SPE 165464, 2013, [102].

An overall view of a three-phase separator equipped with the main internal devices is presented schematically in the following figure:

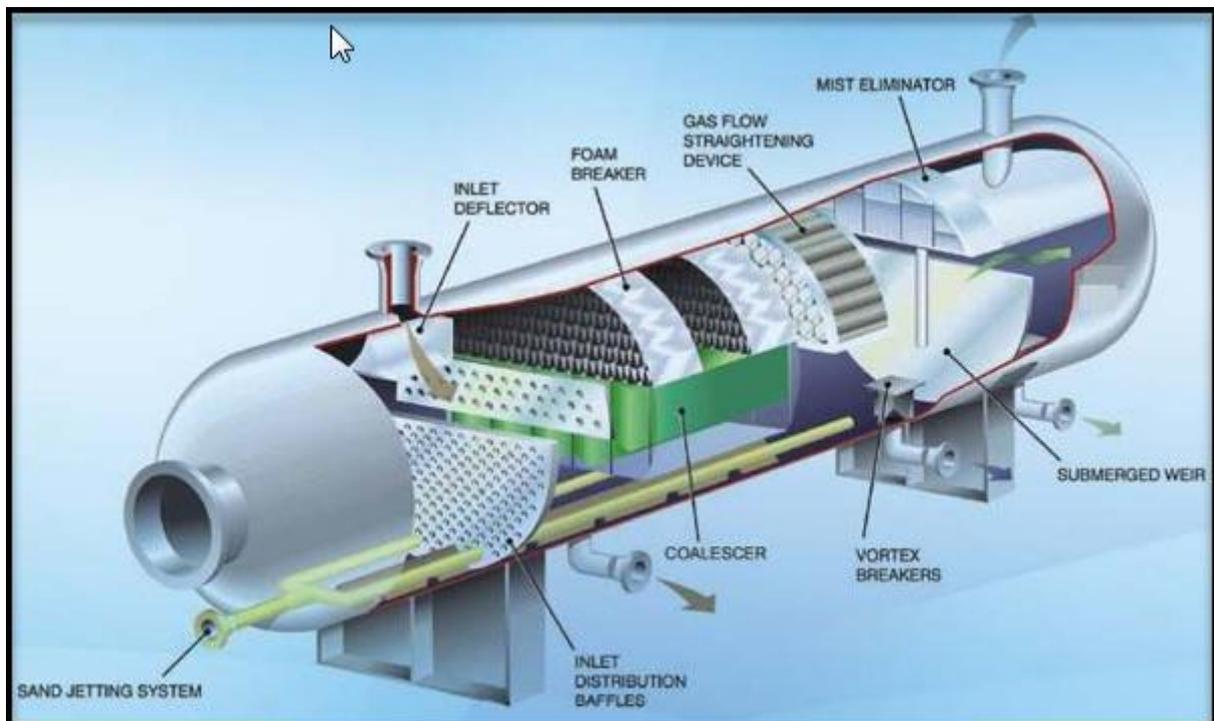


Figure B16: Main Separator Internals, source: Kirk Process Solutions Limited

APPENDIX C

CASE STUDIES

- **Dual Frequency® AC/DC electrostatic treating technology**

1. Mars Offshore Drilling and Production Platform

Mars platform is operating in the deepwater region of Mississippi Canyon area in the Gulf of Mexico. Production started in 1996 designed for 140000 BOPD and in 2003 Mars was producing 200000 BOPD with many process units, including the Bulk Oil Treater, operating above the design capacity. Consequently the platform faced dehydration problems and high rate of demulsifier was required to meet the oil discharge dehydration target of 1% BS&W. The electrostatic treater that was originally selected was a conventional AC treater with serpentine perforated pipe distributor and steel rods. During Katrina hurricane in the late 2005/early 2006 the platform was shut down due to severe damages. During the repair of the platform the Mars AC Treater was upgraded to composite grids, Dual Frequency® electrostatic treating technology and a HiFlo® distributor.

The Mars reservoir fluids are characterized by moderate API gravity and cannot be considered as heavy crudes. Nevertheless, the Mars crude oil has a high content in resins and asphaltenes that stabilized the water in oil emulsions. The fluid properties of four out of the 27 wells that were operating in 2005 are given in the table below.

Well No.	Reservoir	Depth	API gravity	Sat (v %)	Aro (v %)	Res (v %)	Asph (v %)	Atomic S (wt %)	Acid No.
A-1	Pink	13,036	17.3	21.0	61.6	14.0	3.4	2.7	4.4
A-12	Upper Green	16,910	23.5	24.3	55.6	14.9	5.2	2.6	0.6
A-2	Lower Yellow	17,200	28.2	31.5	51.2	15.0	2.3	2.1	0.7
A-4	Terra Cotta	18,476	22.1	25.2	50.2	13.4	11.2	2.8	1.0

Table C1: Fluid properties for the major producing Mars wells and Reservoirs, Field Implementation of New Electrostatic Treating Technology, John M. Walsh, Shell International, SPE, Gary Sams, Cameron, SPE, Joseph Lee, Cameron, SPE, OTC 23200, 2012, [108].

The low gravity of the Pink reservoir fluid and the high acid number can be clearly seen in Table C1 . When the wells from Pink reservoir were produced, the BS&W was even harder to be within specifications and greater dosages of demulsifiers were needed. The Pink reservoir fluids had also a destabilizing effect on the resins and asphaltenes of the other fluids and drove those components to the oil/water interface and stabilizing the oil/water emulsion. The low water content of the produced fluids, between 5 and 10%, made dehydration difficult due to the big spacing between the water droplets and the low coalescence rates.

The Mars process configuration consists of High Pressure and Intermediate Pressure primary separators, and a Low Pressure FWKO. The oil from the FWKO is led to the Bulk Oil Heat Exchanger, the Treater Degassing Separator and then to the Bulk Oil Treater. The water from the FWKO is led to a Hydrocyclone and then to the flotation unit. The described configuration is seen in Figure C1

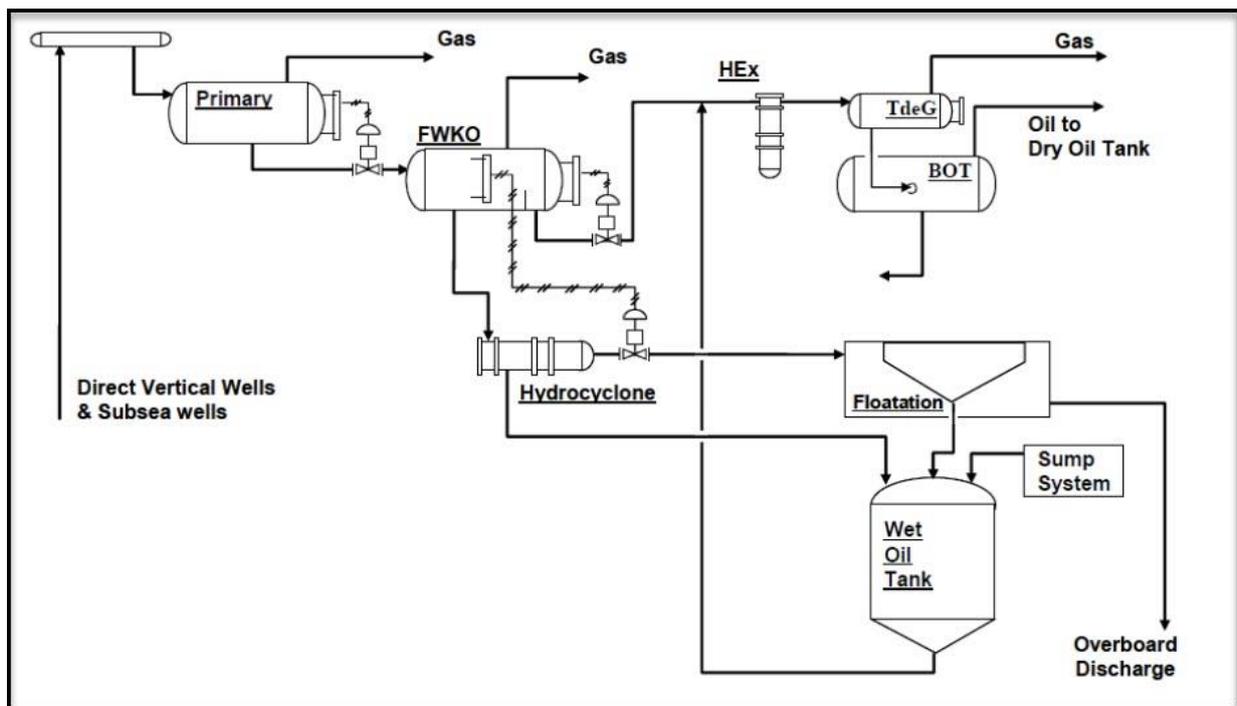


Figure C1: Mars process train, Field Implementation of New Electrostatic Treating Technology, John M. Walsh, Shell International, SPE, Gary Sams, Cameron, SPE, Joseph Lee, Cameron, SPE, OTC 23200, 2012, [108].

The FWKO was unable to remove sufficient amount of water leading higher water quantities than designed in the BOT. The result was the use of significant demulsifier injections in high rates. One additional problem was the oily water that was recycled from the Wet Oil Tank to the BOT. Due to pressure differences the water was pumped to reach the BOT and was subjected in severe shearing from the centrifugal pump creating a pad of a complex reverse (o/w/o) emulsion in the BOT. This pad had an intermediate viscosity between that of oil and water and reduced even more the effectiveness of the dehydration equipment.

In order to achieve dehydration improvement two options were studied before upgrading the electrostatic treater and these were heat addition and chemical optimization. Heating addition upstream of the FWKO and in the Heat exchanger were examined as two different options but they were both rejected as costly and inefficient respectively. Chemical optimization was studied as well but in comparison with mechanical/electrostatic solutions was evaluated to be much more expensive.

The original AC treater of the Mars platform was operating with an electric field of 1.13kV/cm which is considered to be a typical value for treating moderate gravity oil and the flux rate in the vessel during pick production was 383BPD/sq ft. This rate is considered very high for a conventional AC treater.

The potential for more reliable dehydration, increased capacity and reduced chemical usage led to the selection of Dual Frequency AC/DC electrostatic technology for the retrofit of Mars dehydrator. After retrofit was completed and by keeping all process parameters constant the demulsifier dosage was reduced almost about a half as the following table demonstrates.

Crude	API	Temp (F)	Inlet BSW (%)	Technology	Flow rate BPD	Outlet BSW (%)	Chemical rate (gal/1000 bbl oil)
Mars GOM	27	120	5	AC	148,000	0.3	2.0
Mars GOM	27	120	5	Dual Frequency	148,000	0.3	1.0 ~ 1.2

Table C2: Reduction in chemical consumption when the treater was converted from AC to Dual Frequency with all production parameters kept constant, Field Implementation of New Electrostatic Treating Technology, John M. Walsh, Shell International, SPE, Gary Sams, Cameron, SPE, Joseph Lee, Cameron, SPE, OTC 23200, 2012, [108].

Based on these results it is concluded that the Dual Frequency technology is much more efficient without increasing the process temperature or reducing the flow rate. This technology can handle higher flow rates in a smaller footprint while at the same time meets the specifications of export crude oil and it can be used in lower API gravity oils (see next case studies).

In numbers, after the implementation of DF technology, the savings in demulsifier costs were US\$ 600,000/year thus the payout of the electrostatic treater upgrade based on the demulsifier cost savings alone, is less than 2 years.

2. Reduced treater size and/or production increase for dehydrating 20.6 API, TAN 4 Crude Oil

Cameron Technology Center in Houston, Texas had carried process tests that verify significant improvement of dehydration capacity with the Dual Frequency technology [243]. More specifically, a 20.6 API, TAN 4 crude oil was used to demonstrate the increasing dehydration efficiency from the conventional AC technology to the Dual Polarity and Dual Frequency technologies. The treatment temperature was 221 °F and the dehydration from 15% to 0.5% BS&W [104]. Dehydration improvements were demonstrated both in terms of treater size reduction for a given flow rate and in terms of production increase for a given treater size.

For a 12' x 80' treater dehydrating API 20.6°, TAN 4 crude with an outlet specification of 0.5% BS&W using AC/DC Dual Polarity and Dual Frequency technologies instead of conventional AC technology the flux could be improved by 25% using Dual Polarity and 125% using Dual Frequency technologies. This is shown in Figure C2 and Table C3

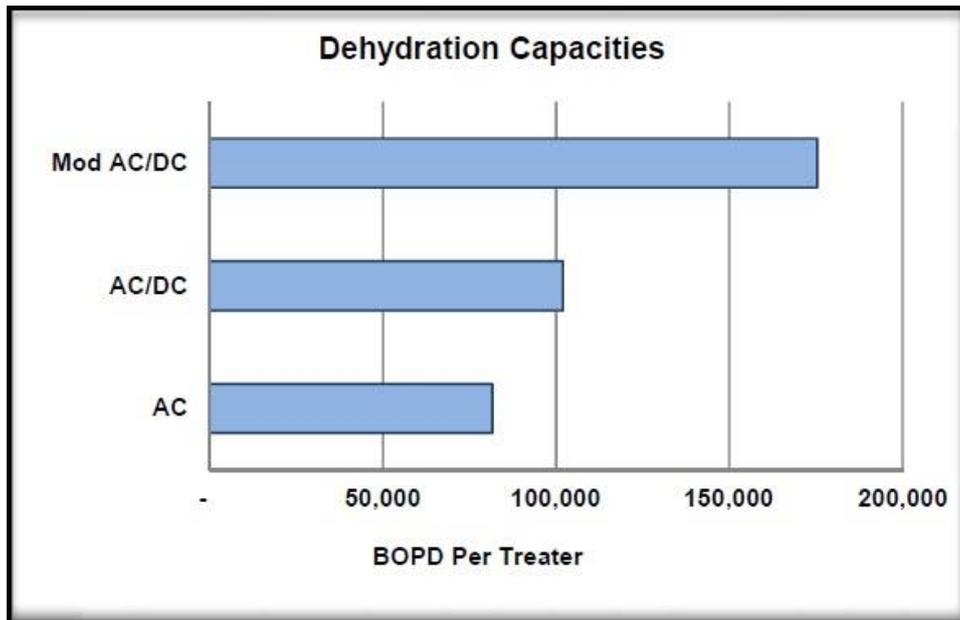


Figure C2: Treatment capacities for 12' x 80' treater using different electrostatic technologies, *Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters*, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, *Cameron Process Systems*, SPE 156299, 2012, [106].

Technology	Treatment Capacity BOPD	Capacity Increase BOPD	Capacity Increase %
AC	81,600	-	-
DP	102,000	20,400	25%
DF	175,500	93,900	125%

Table C3: Dehydration improvement in terms of flow rate, *Improved Dehydration and Desalting of Mature Oil Fields*, Erik Sellman, SPE, Gary Sams and S. Pavan Kumar Mandewalkar, SPE, *Cameron Process Systems*, SPE 164289, 2013, [103].

The flux versus outlet BS&W (%) from the technology upgrade can be also seen in the following diagram where AC, Dual Polarity, Modulated Dual Polarity and Dual Frequency technologies are projected.

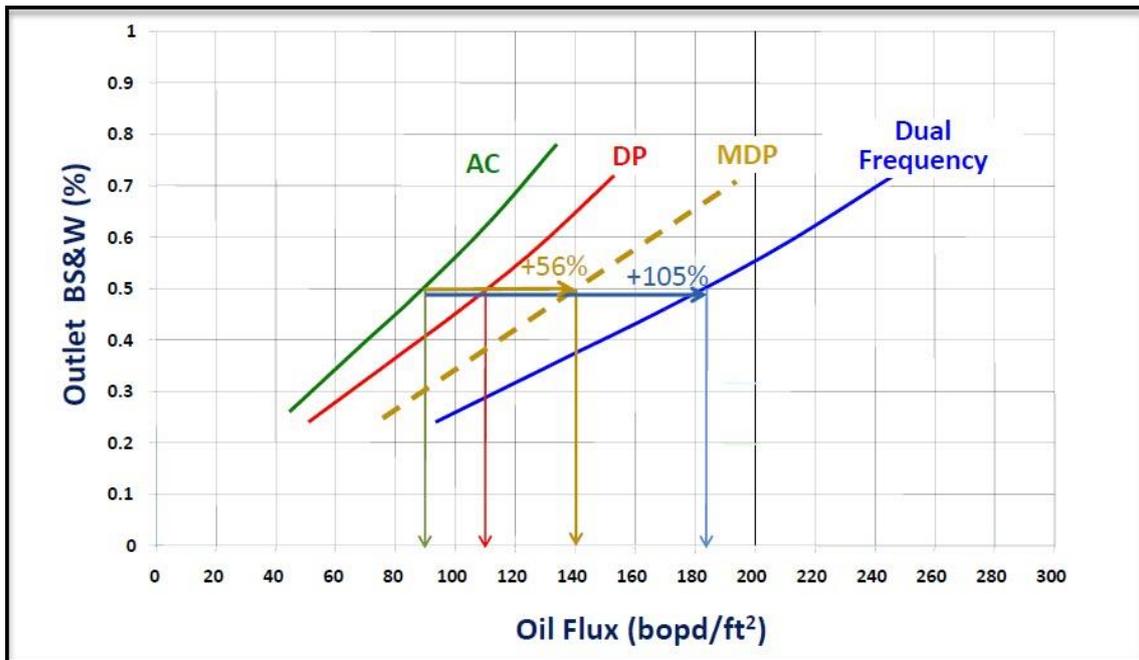


Figure C3: Higher flux from technology upgrade, Improved Dehydration of Heavy Oil and DilBit Using Modulated Electrostatic Treaters, Erik Sellman, Dan Senft, Terry Murtagh, Cameron Process Systems, World Heavy Oil Congress, 2016, [141].

The sizes required for each technology, given a constant diameter 14' ID for dehydrating 100.000 BOPD of API 20.6° crude oil from 15% to 0.5% BS&W are demonstrated in the following figure.

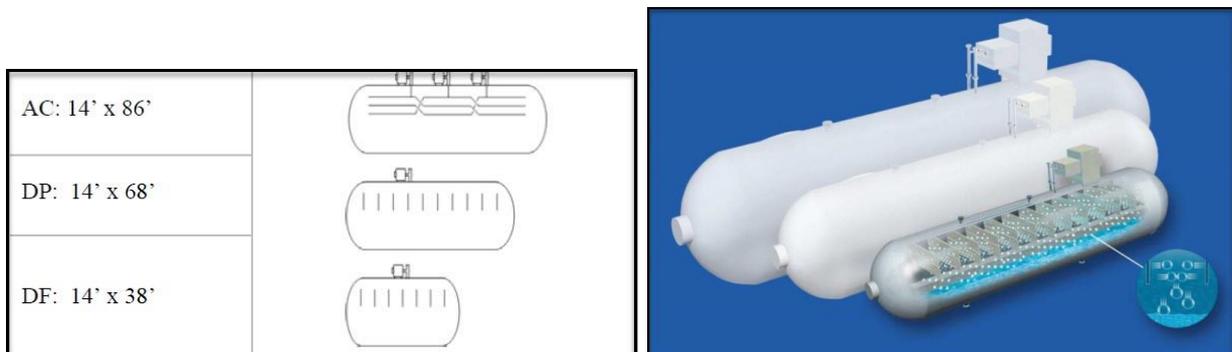


Figure C4: Reduction in the size of the treater using different electrostatic technologies, Compact and lightweight equipment for floating production systems, Erik Sellman, Pavan Mandewalkar, Gary W. Sams, Cameron Process and Compression Systems, OTC-24727-MS, 2014, [104].

Based on the results, the Dual Frequency technology would require a 66% shorter treater while maintaining the same BS&W in the crude oil.

3. Savings on crude oil heating with Dual Frequency technology

This case study involves electrostatic treaters in a heavy crude oil field in Colombia and the fluid flow loop testing which was conducted to determine the optimum technology among AC, Dual Polarity AC/DC and Dual Frequency AC/DC technologies. Also, the potential of reducing the operating temperature 20°F was examined. The dimensions of the treaters were of 3.6 m x 18 m and the treated fluid a blend of 18° API with 30% BS&W which had to reach the specification of 0.5% BS&W. The flow rate of 60.000 BOPD and the results for each technology are presented in the following table.

Technology	Viscosity cP	Oper. Temp (°F)	Delta-T (°F)	Savings (M BTU/hr)
AC	5.6	200	-	-
Dual Polar	6.9	186	14	12.4
Dual Freq	8.9	170	30	26.5



Table C4: Operating temperatures for three different technologies, Improved Dehydration of Colombian Heavy Crude Oil Using Dual Frequency Treaters, Erik Sellman, Juan Alvarez, Juan Carlos Guillen Cameron Process and Compression, ACIPET 2013, [139].

The heat saving with the Dual Frequency technology was 26.5 M BTU/hr corresponding to annual savings of 1.500.000 USD [141].

4. North Sea Heavy Oil Offshore dehydration

A development in the North Sea needed compact treater for dehydration of 84.500 BOPD of a 16.9° API crude. The treatment temperature was 105°C and Dual Frequency technology was selected as it has the ability to be more compact for the given flow rate compared to AC and Dual Polarity technologies. In addition, Dual Frequency technology can handle the normal dehydration case from 10% BS&W to 0.5% BS&W in a 2 x 50% split flow configuration in two parallel treaters and also the case of full flow in a single treater with 1% BS&W target. Consequently the installation of two treaters in parallel with Dual Frequency technology were 4 x 10m in size. For the same operating conditions AC technology treaters would have been 4 x 19.7m in size and thus 97% longer and heavier [104].

5. 12.5° API Heavy Crude Oil dehydration in Colombia

An onshore facility in Colombia treats a heavy crude oil of 12.5° API without the use of diluent. The BS&W specification is reached despite the apparent difficulties in processing. That is because Colombian crude oil is characterized by low formation water salinity which reduces the density differential between crude oil and produced water, the tendency to create tight w/o emulsions and the high water cuts downstream the FWKO that overload the electrostatic treaters. The separation facility includes FWKO, wet/dry crude interchanger (device used to transfer heat between a solid object and a fluid), crude oil heater and several Dual Frequency electrostatic treaters in parallel configuration. (Figure C5)

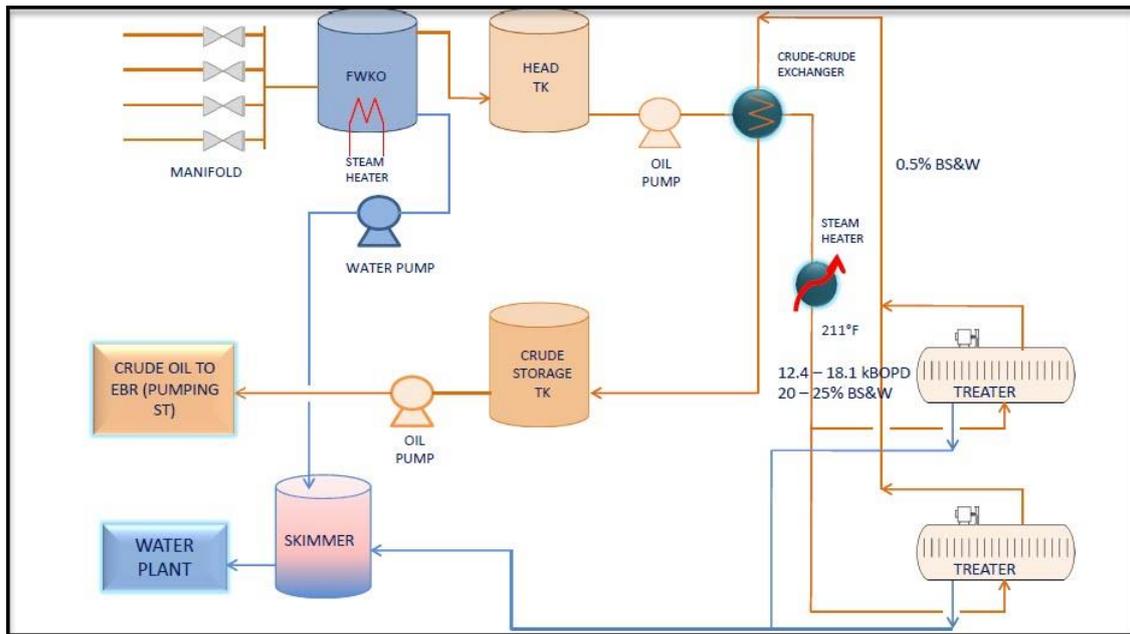


Figure C5: Onshore processing facility for dehydration of 12.5° API heavy crude oil without diluent addition in Colombia, 2-Prong Approach for Enhanced Dehydration of Heavy Crude Oil and DilBit, Erik Sellman, SPE, Terry Murtagh, PE, Pavan Mandewalkar, SPE, Cameron Process Systems, World Heavy Oil Congress, Canada, 2015, [265].

The process temperature is around 100°C (211°F) at the final stage of dehydration. The inlet stream contains 20% BS&W and the target of 0.5%BS&W is achieved.

6. El Furrial oil field Eastern Venezuela, 25° API

EL Furrial oil field has been discovered in 1985 and has been one of the most important discoveries in Venezuela’s oil history. The field produces light to medium crude oil but its production is considered very challenging due to the intensity of asphaltene precipitation problems.

Initially the production facility included five AC/DC treaters in parallel service which were gradually upgraded. The first upgrade concerned the inlet spreader design and the electrode material from steel to composite. The second upgrade was the addition of Dual Frequency power units which, in the existing configuration almost doubled the flow rate through the treaters. Eventually five treaters operated in parallel and treated a 25° API crude with 11% BS&W which reached the specification of 1% BS&W [265] (Table C5). The upgrade from Dual Polarity with HiFlo spreader and composite electrodes to Dual Frequency technology provided an additional 25.000 BOPD capacity increase to 85.000 BOPD per treater [103].

Technology	OT °F	Spreader	Electrodes	BOPD / Treater
DP	140	Open bottom	Steel	43,000
DP	140	HiFlo	Composite	60,000
DF	140	HiFlo	Composite	85,000



Table C5: Higher capacity using improved internals and Dual Frequency technology, 2-Prong Approach for Enhanced Dehydration of Heavy Crude Oil and DilBit, Erik Sellman, SPE, Terry Murtagh, PE, Pavan Mandewalkar, SPE, Cameron Process Systems, World Heavy Oil Congress, Canada, 2015, [265].

7. 11.7° API Extra Heavy Crude Oil dehydration

An onshore Extra Heavy Oil production facility in Colombia treats an 11.7° API extra Heavy crude with the addition of diluent, heating and Dual Frequency technology. The demanding crude rendered the choice of the appropriate diluents important and also the setting of the optimal parameters for the Dual Frequency controller. The diluted oil has an API gravity of 16.4° and the treating temperature is 162°F. The inlet fluid contains 20 – 25% BS&W and the flow rate for each treater is 18.400 BOPD. The 16.4° API diluted crude reaches the specification of 0.5% BS&W after dehydration [265] (Figure C6).

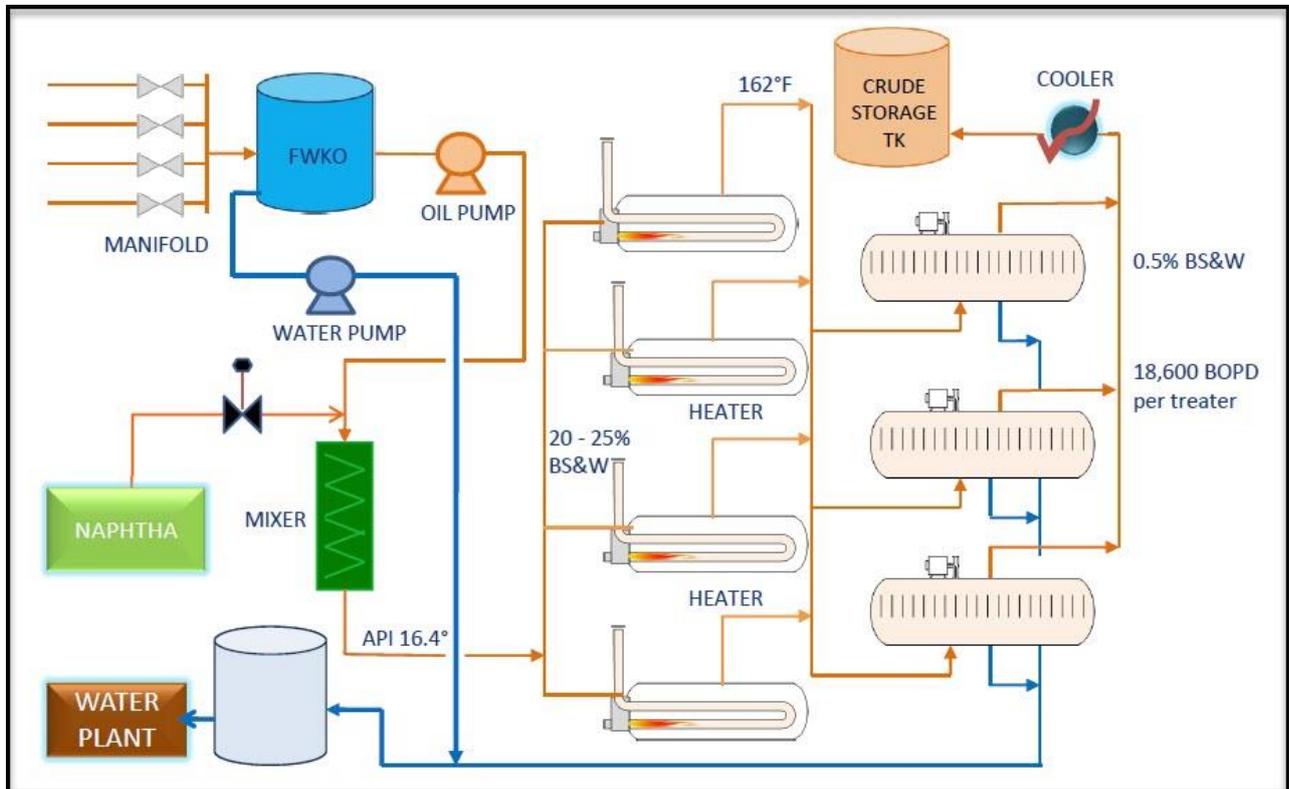


Figure C6: Onshore processing facility for dehydration of 11.7° API extra heavy crude oil with diluent addition in Colombia Improved Dehydration of Heavy Oil and DilBit Using Modulated Electrostatic Treater, Erik Sellman, Dan Senft, Terry Murtagh, Cameron Process Systems, World Heavy Oil Congress, 2016, [141].

8. Dual Frequency AC/DC SAGD treaters in Alberta, Canada

Bitumen is a black highly viscous hydrocarbon mixture that often contains high amount of solids and has a very high specific gravity which is typically heavier than water. The density and the viscosity of bitumen is reduced with dilution with a suitable diluents solvent. If the solvent is a lighter crude oil than the blend is called DilBit and if the bitumen is diluted with synthetic crude oil, then the blend is called SynBit. The properties of DilBit and SynBit are very similar to the ones of heavy oil with 14 – 16 API and they usually have high solid content. There is no formation water associated with DilBit and SynBit besides the steam from SAGD process so the water salinity is very low [102].

In this case study, treaters are operating on SAGD in Alberta and treat a blend with almost 16° API DilBit. The emulsion stability is very high and so are the flow rates, thus the treaters are large vessels with 14' ID and S/S length of 110'. The treatment temperature is 131°C. The high concentration of solids and fines in the feed as well as the very low water salinity make the dehydration of the DilBit emulsion very difficult. The high viscosity of the emulsion renders the water removal in the upstream FWKOs limited so the

treaters have to operate with a feed of high water cut, which often reaches 18 – 20% v/v. A schematic illustration of the process facility is given in the figure below.

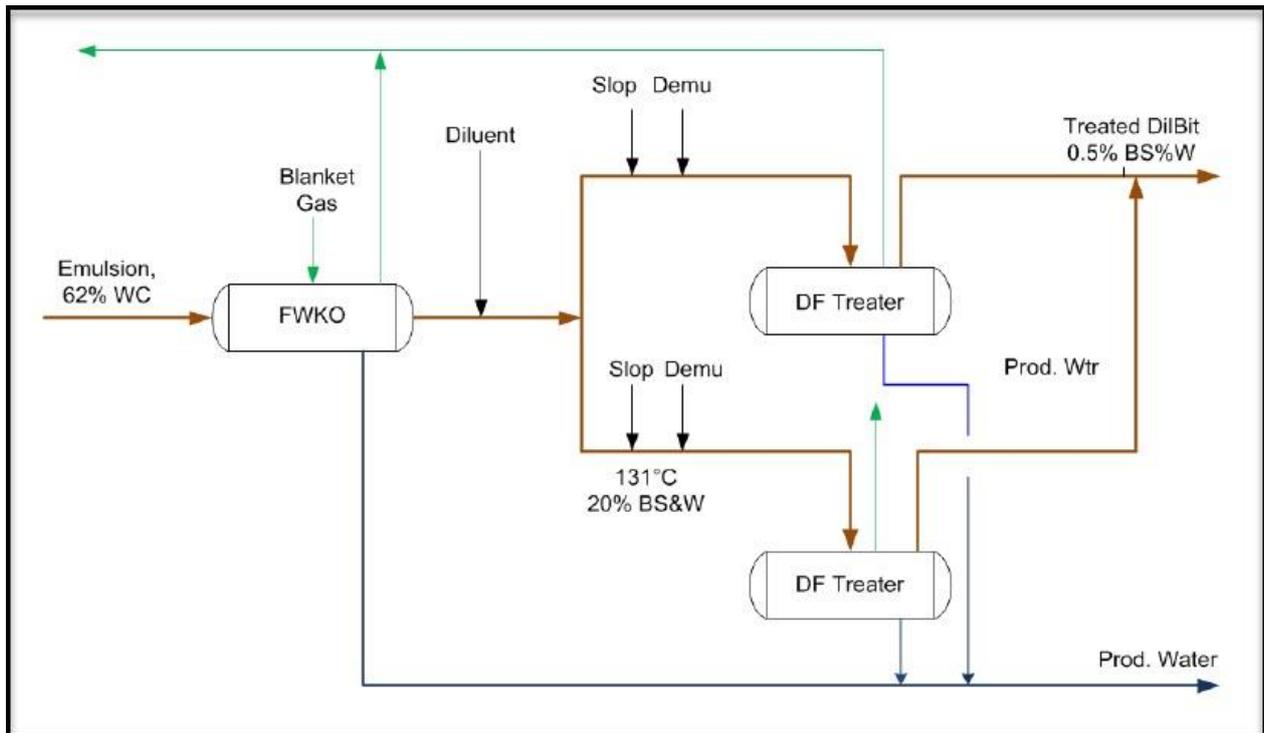


Figure C7: SAGD DilBit production facility in Alberta, Effective Dehydration of Canadian Heavy Crude Oil and Dilbit, Erik Sellman, SPE, Terry Murtagh and Benjamin Manuel, Cameron Process and Compression Systems, SPE 165464, 2013, [102]

The Dual Frequency treaters provide a DilBit with a BS&W content at or below 0.5% v/v. The size of the treaters requires double Dual Frequency power units which operate in separate sections on the electrostatic grids.

9. Sizing comparison of treaters dehydrating API 18 and API 30 crude oils with different technologies

The sizing comparison is based on 30° API crude oil with 12% formation water salinity and 18° API crude oil with 4% formation water salinity using conventional AC technology, Dual Polarity technology (DP) and Dual Frequency (DF) at 50,000 BOPD flow rate and BS&W content of 0.2%. The size of the vessels is particularly important on FPSOs and offshore platforms where there are space restrictions.

Technology	Crude API°	Brine Salinity %	Inlet BS&W, %	Temperature °C	Vessel Diameter ft	Length ft	Length Savings %
AC	18	4	15	100	14	94	-
DP	18	4	15	100	14	78	17
DF	18	4	15	100	14	58	38
AC	30	12	15	40	12	56	-
DP	30	12	15	40	12	46	18
DF	30	12	15	40	12	36	36

Table C6: Treater sizes required to treat 50,000 BPPD of wet crude to 0.2% BS&W, Benefits of Using Advanced Electrostatic Fields in Crude Oil Dehydrators and Desalters, Erik Sellman, SPE, Gary W. Sams and S. Pavan Kumar B. Mandewalkar, SPE, Cameron Process Systems, SPE 156299, 2012, [106]

10. Upgrade from Dual Polarity to Dual Frequency for 5 different crudes

In the following table the outlet of BS&W (%) for the two different technologies, DP and DF are represented for crudes from 14 - 25° API.

Crude ID	API°	Temp (°F)	Feed BS&W (%)	Flux (BOPD/sq ft)	Dual Polarity BS&W (%)	Dual Frequency BS&W (%)
Venezuela	25	150	11	200	1.8	0.6
Venezuela	17	280	28	94	0.3	0.1
Wyoming	24	130	11	100	1.45	0.85
Oklahoma	23	135	11	75	1.35	0.5
Brazil	14	240	5	80	0.5	0.27

Table C7: Upgrade of existing crude oil treaters from Dial Polarity to Dual Frequency API 14 – 25, Improved Dehydration and Desalting of Mature Oil Fields, Erik Sellman, SPE, Gary Sams and S. Pavan Kumar Mandewalkar, SPE, Cameron Process Systems, SPE 164289, 2013, [103].

To conclude over the efficiency of Dual Frequency AC/DC electrostatic treatment technology, it has been demonstrated that conventional AC treaters are less efficient in treating heavy crude oil (and DilBit), mainly because of lower voltage gradient and higher sensitivity to lower water cuts.

The Dual Polarity and Dual Frequency AC-DC treaters provide important benefits for heavy oil dehydration that concern production, Operating Expenditures and Capital expenditures.

The production benefits include higher flow rates in existing treaters, lower BS&W in the treated crude oil, tolerance to high water cuts in the feed, improved effectiveness in treating conductive crudes and resolving rag layers.

The OpEx benefits include lower operating temperatures, reduced demulsifier dosage and reduced electric consumption.

The CapEx benefits are associated with the smaller treater sizes, the lower installation cost and the transportation savings due to the lower BS&W.

- **Vessel Internal Electrostatic Coalescer (VIEC™)**

1. Troll C platform installation (Start-up)

The first installation of VIEC from Norsk Hydro was at Troll C platform in July 2003. Before the actual installation the electrocoalescer device was still in prototype stage. Before VIEC was certified for installation, an extensive qualification review program was performed along with a multiphase rig test using real crude oils ranging from 21 to 32 API. One of these oils was from the Troll field at 29°API.

The extensive qualification review included checking over the durability of epoxy modules in elevated temperatures and their resistance in moving debris inside the separator, the penetrator design safety (durability in 400barg), the zone requirements and the motion robustness of the steel structure. All the aspects that were checked showed excellent results [145].

The multiphase rig test was performed at temperatures of 50-80°C and pressures of 40-70 barg with synthetic produced water and no additional chemicals. The separator was 24" diameter and 3 meters long. VIEC showed an excellent ability in reducing the emulsion layer and the produced water quality was significantly better.

The installation of VIEC in the first stage separator of Troll C platform took almost 6 days. The Troll C first stage separator is a split flow symmetrical separator and the installation of the two VIEC electrocoalescers was done simultaneously in both sides of the separator. One end of the symmetrical split flow is shown in the following figure.

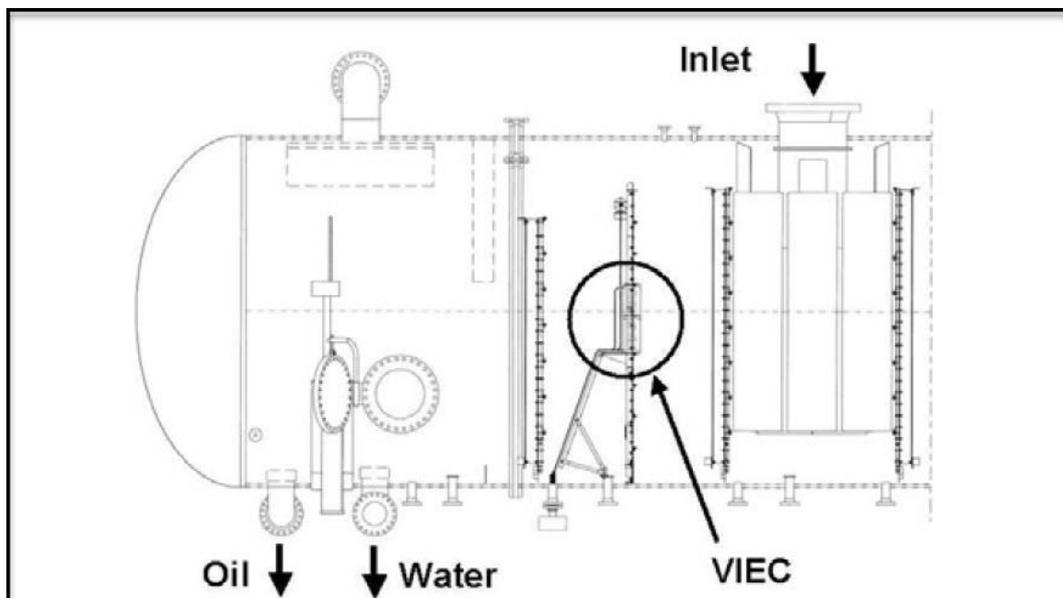


Figure C8: Troll C separator side view, flow towards left, Advanced Electrostatic Internals in the 1st Stage Separator Enhance Oil/Water Separation and Reduce Chemical Consumption on the Troll C Platform, Erik A. Wolff/ABB Offshore Systems, Trond L. Knutsen/Norsk Hydro, Wojciech Piasecki/ABB Corporate Research, Peder Hansson/ABB Offshore Systems, Pål J Nilsen/ABB Offshore Systems, OTC 16321, 2004, [145].

VIEC was activated in July 2003 and the improvements in operating conditions were noticed almost immediately: There was a significant reduction in the thickness of the emulsion layer by 2/3 and the interfaces of oil, emulsion and water layers were clearly defined. The demulsifier dosage before the VIEC operation was 10 ppmv on fluid basis whereas after VIEC it was reduced gradually to 2ppmv. The possibility of removing chemical addition completely was also tested but the results were not positive. After the first months of VIEC operation, in January 2004, the demulsifier dosage was further decreased to 1.5ppmv. In terms of cost savings, the decrease in demulsifier dosage saved the operator 600.000\$/year.

During the initial period of VIEC operation, the antifoam injection in the first stage separator appeared to have a negative effect in the separation performance. The injection of the antifoam was stopped in November 2003 and no problems were observed in the separation performance. Lastly, the average oil quality from the first stage separator was significantly improved by the end of 2003. Before VIEC, the water in the oil outlet was more than 10% whereas by the end of 2003 it was only 4.8%. This first application of VIEC technology provided also positive indications for possible use in subsea separation projects since it could eliminate hydrate problems in subsea pipelines [145].

2. Dukhan field, Qatar, VIEC and VIEC LW installation 40° API (Start-up)

The first well in Dukhan oil field in Qatar was drilled in 1938 and since then the field has been developed in several stages. Production challenges appeared in a specific well in the field (no further information is available) involving operating pressure decline from 400psig to 170psig causing changes in viscosity and droplet size due to shearing and increased GOR. Until February 2012 when Hamworthy, Wärtasilä Oil &

Gas Systems AS commissioned and installed VIEC system (VIEC and VIEC LW) the exact operating conditions during the winter period were: operating temperature down to 4°C, operating pressure 170psig, flow rate 64000BPD, water content in oil outlet more than 5% and average demulsifier consumption during winter 30ppmv. These conditions deviated from the targeted specifications which were less than 0.2% BS&W and salt content in export crude less than 50PTB. [240].

Qatar Petroleum carried out 2-week-long performance tests to verify and confirm performance improvement due to VIEC system. The VIEC LW system was piloted at the 13-V-4001 degassing vessel outlet in Fahahil South degassing station in Dukhan field. The design capacity of 13-V-4001 vessel was 100000BPD and its dimensions 4m ID and 21m T/T. The two VIEC LW walls were installed in the vessel as it can be seen in the figure.



Figure C9: VIEC LW electric sheets installed in 13-V-4001 vessel, Vessel Internal Electrostatic Coalescer Technology (VIEC), SPE 156087, Ali AlQahtani, Qatar Petroleum, 2012, [240].

VIEC LW system was installed downstream the VIEC at an elevation that covered the liquid layer. The whole VIEC installation is shown in the following figure.



Figure C10: VIEC technology installation in Qatar, Electrostatic Destabilization of water-in-oil Emulsions: Use of laboratory scale experiments to debottleneck and optimize separation processes, Erik Bjørklund, Toulouse Workshop, 2012, [153].

A third party employed by Qatar Petroleum collected the samples every 6 hours and tested them in Qatar Petroleum laboratories. Test samples were collected before and after VIEC system installation to enable comparison. The results were the following:

- Water in oil outlet: Before modification the demulsifier injection was at 30ppmv
The water in oil outlet varied from 0.2 to 1.2% BS&W

After modification the demulsifier injection was at 30ppmv

The range was 0.05 to 0.2% BS&W

After optimization with demulsifier injection 12ppmv (60% reduction)

The range was 0.1 – 0.3 BS&W
- Salt content in oil outlet: Before modification the demulsifier injection was at 30ppmv
The salt content was varying over a large range from 40 to 200 PTB

After modification the demulsifier injection was at 30 ppmv

The fluctuations were minimized and the range was 20 – 46 PTB

After optimization with demulsifier injection 12 ppmv (60% reduction)

The range was 20 – 64 PTB
- Oil in water outlet: Before modification the demulsifier injection was at 30 ppmv
The oil in water outlet was varying quite a lot reaching the value of 900 ppmv

After modification the demulsifier injection was at 30 ppmv

The fluctuations were minimized and the range was 180 – 357 ppmv

After optimization with demulsifier injection 12 ppmv (60% reduction)

The range was 400 – 600 ppmv

Consequently, without any change in the demulsifier's dosage, the water in oil outlet was 0.05 – 0.2 BS&W, the salt content was between 20 and 46 PTB and the oil in the water outlet was reduced by almost 50%. These results demonstrate that further reduction in demulsifier dosage is feasible since the target for BS&W was less than 2% [153].

3. Energy savings and size reductions in a 20°API crude oil (Batch test/Flow-loop test)

This case involves the development of a heavy oil offshore field in Canada which would be processed in a stand-alone platform. The operator evaluated the VIEC technology and the purpose was to reduce the process temperature at the intermediate stage of the separation train and also reduce the size of the separator. The difficulty of the task was due to space limitations on the platform and the lack of gas in the reservoir [151].

The benefits of VIEC were evaluated by two tests in collaboration with the chemical vendor close to the field and a laboratory in Norway. Both tests utilized oil from a well drilled previously on the field.

The results showed that the target of 10% BS&W at the outlet of the medium pressure separator (intermediate stage) could be achieved with VIEC even if the operation temperature was reduced to 60°C. The operation temperatures that were initially considered were 80 - 90°C. Additionally, the size reduction would be from 29 to 21.5 m, saving a lot of space in the platform. One important result was that with VIEC the separation efficiency was much less dependent on the type of demulsifier.

The following chart in figure C11 displays the batch test results concerning separation efficiency with and without VIEC in 8 and 15 seconds at 60°C.

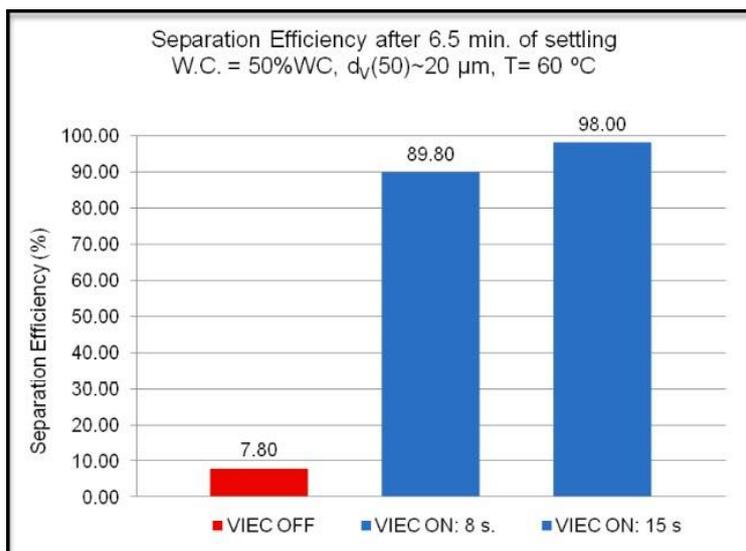


Figure C11: Chart displaying separation efficiency with and without VIEC in 8 and 15 seconds at 60°C, Efficient and Flexible Crude Oil processing technology for Oil Separation facilities, MORAD AMARZGUIOUI & PER CHRISTIAN JACOBSEN, Wärtsilä Oil & Gas Systems AS, Norwegian oil and gas partners, 2014, [143].

These results are indicative since the batch tests cannot be directly extrapolated to a full-size separator. The reason is that these tests are static and they lack the dynamic effects that are met during production such as turbulence and shear [143].

4. Energy savings in a 14°API crude (Batch test/Flow-loop test)

This study involves the development of a heavy oil field in a remote location in the Middle East. The crude derives from several different wells and the API range is from 11 to 18. The wells are produced from two different separation trains. The separation equipment was subjected to size limitations due to the remote location and the available energy supply for heating was also limited by the quantity of the produced gas.

The design liquid capacity was 180 MBLD with feed BS&W ranging from 33 to 83% and GOR from 23 to 64 scf/stb (limited amount). The export specifications were BS&W from 0.1 to 0.5%, crude salt content 7 PTB (pounds per thousand barrels) and oil content in water 10ppm [153].

The first design that was proposed included two conventional separation trains that would operate in a three-stage separation scheme. It included a single very large first stage FWKO operating at 80°C, then heating of the outlet stream up to 125-145°C in order to split it to two parallel second stage gravity separators before it was recombined, mixed with fresh water for desalting purposes and treated in a conventional electrostatic coalescer.

In order to mitigate the high operating costs that resulted from the conventional design the oil company investigated the possibility of using VIEC. For this reason, 100lt of crude oil from one of the wells, along with produced water, was delivered to Wärtsilä Oil & Gas Systems laboratory where a feasibility study was performed.

The API gravity of the sample was 14 and based on the results of the study the application of VIEC solution led to significant reductions both in heating and water consumption. The scheme included a first stage FWKO operating with no additional heating, then the stream would be heated to 70°C and routed to the second stage gravity separator equipped with VIEC system. The stream from the second stage separator, which contained water less than 15% of the total volume, was heated to 95°C and routed to the next gravity separator which was also equipped with VIEC coalescer. The outlet stream from the last gravity separator had 2% water content. The last stage included mixing with fresh water and treatment in a conventional electrostatic treater operating at 120°C.

For the calculation of the operating costs for both solutions computer simulations were performed. As it is shown in the following table, with VIEC technology there is approximately a 65% reduction in heating energy consumption and 41% reduction in wash water consumption [151].

	Conventional	Wärtsilä
Total Steam Energy Consumed [kW]	14351-48622	10378-17247
Total Heat Loss to Air [kW]	6965-29278	5166
Wash water consumption, m ³ /h (BWD)	21 (3168)	12 (1810)
Process train weight	472 Ton	314 Ton
Number of oil treaters required	TWO	ONE

Table C8: Comparison of heating consumption, heat loss, wash water consumption, process train weight and number of treaters required for conventional solution and VIEC solution, *Overcoming Separation Challenges by Use of Electrocoalescence Technology*, Morad Anarzuoui and Per Christian Jacobsen, Wärtsilä Oil & Gas Systems AS, SPE-171483-MS, 2014, [151], *Electrostatic Destabilization of water-in-oil Emulsions: Use of laboratory scale experiments to debottleneck and optimize separation processes*, Erik Bjørklund, Toulouse Workshop, 2012, [153].

Furthermore there is a significant reduction of 58 Tons in the weight of the process equipment and the required treater is one instead of two. The following tables for CapEx and OpEx were calculated based on steam price 20USD/Ton, demulsifier price 0.4USD/ppm/bld/year and electric power 0.12 USD/kWh.

CAPEX		Savings		OPEX		Savings	
Steel cost		3.16 MUSD		Steam cost		21.5 MUSD/year	
Heat exchangers		1.5 MUSD		Demulsifier		3.0 MUSD/year	

Table C9: CAPEX and OPEX savings with the application of VIEC, *Electrostatic Destabilization of water-in-oil Emulsions: Use of laboratory scale experiments to debottleneck and optimize separation processes*, Erik Bjørklund, Toulouse Workshop, 2012, [153].

5. Temperature and size reduction in 20°API crude oil (Batch test/Flow-loop test)

A heavy oil onshore field in Africa is located in a preserved wild-life area and the challenge was the process facilities to be as compact as possible in order not to disturb the surroundings. The oil API gravity of the field is approximately 20. An additional restriction was that the operating temperature could not be above 90°C, due to scale precipitation risk from the fact that the produced water was from carbonates.

Emulsion stability problems were also possible because of the polymer EOR that was implemented to produce from the field.

The feasibility study evaluated VIEC's performance both with polymerized and non-polymerized produced water. The results showed no significant difference between the two cases. More profound impact was observed on the operating temperature. With VIEC installed in the low pressure stage the water in oil specification could be achieved with 75°C. In addition, the size of the separator with VIEC would be 3.2m ID and 14.3m T/T (seam-to-seam length is also referred as Tangent-to-Tangent or tan-tan or T/T), while the conventional solution, without VIEC, required a very large vertical settling tank at the last stage of the process train. The VIEC solution, on the other hand, required a moderately sized conventional electrostatic treater. Thus, a significant reduction in the footprint of the process facility could be achieved by implementing VIEC technology as well as significant reduction in the required process temperature [151].

6. One stage separation with 17 API crude (Flow-loop test)

This case study involves a project that aimed to define the operational parameters for a separation process with heavy oil using one-stage separation. The equipment used was the Hamworthy Mobile Test Separator that has been installed in Norsk Hydro's multiphase test rig in Porsgrunn, Norway. The test rig tries to simulate real offshore conditions and fluids. The rig includes a reservoir, feed pumps, a compressor, a multiphase pump, instrumentation and a test separator (here the Hamworthy test separator) that is supplied with the electrostatic internals of VIEC and VIEC LW (Figure C12).

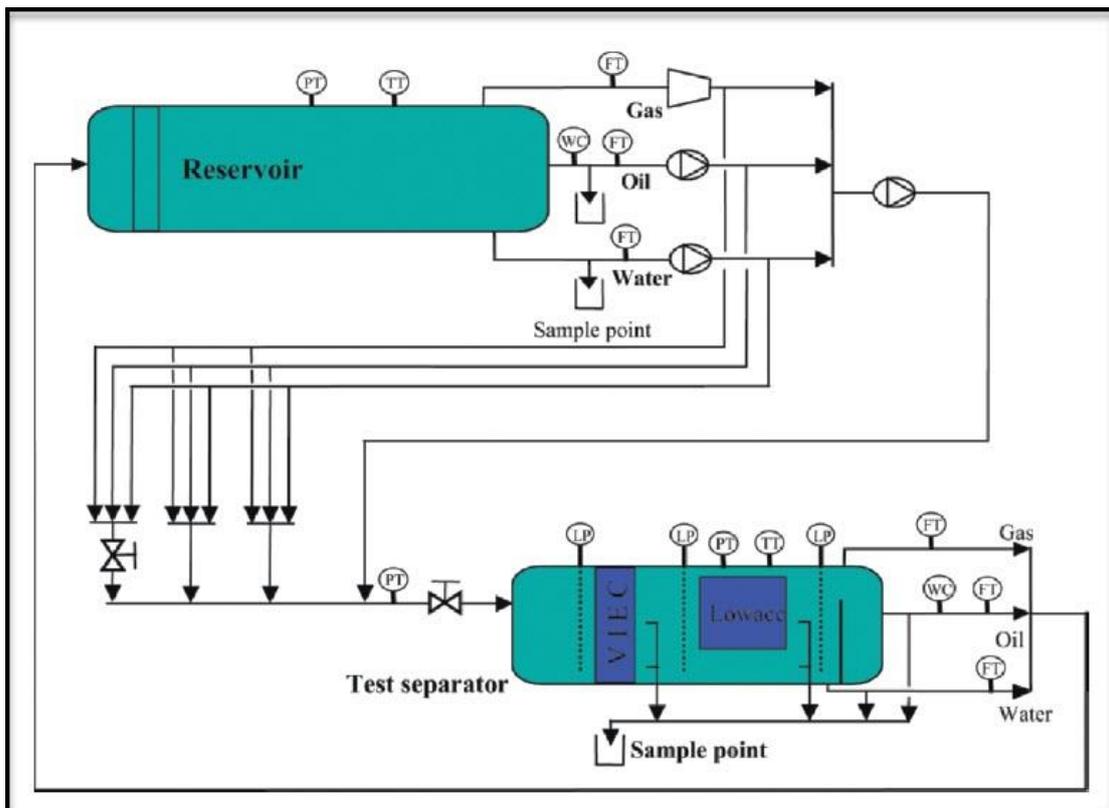


Figure C12: Simplified schematic representation of the Norsk Hydro Multiphase test loop, *A Holistic Solution for Compact Oil Treatment Separation Systems*, M. Chiesa, W. Hamid, Hamworthy Oil & Gas Systems AS, SPE 136836, 2010, [266].

The fluid mixture is separated in the reservoir and water, oil and gas are routed through separate pipes to pumps and compressors. From there, under preset target values for GOR, water cut and flow rates, the separate phases are mixed in mixing systems that simulate the production conditions of one, two or three wells. The water/oil/gas mixture passes through a valve that simulates the choke and is subjected to high shear forces which create a very stable emulsion that enters in the Hamworthy test separator. The two inlet coalescers of the test separator break down the emulsion layer and separate the phases once again. The phases from the three different outlets of the test separator are once again mixed and sent back to the reservoir where they are separated in order to repeat the procedure. This whole circle is continuous in order for the stream to enter the test separator in the set flow rate and achieve as close as possible the field conditions. The result of the multiphase test loop can be conservative compared to the real operation conditions due to the fact that the very complex emulsions that are created may be more stable than the ones from the wells.

The tests covered various operational conditions taking into account a wide range of parameters. The demulsifier concentration was 0.2 or 80ppm, shear forces ranged from 0 to 10 bar in pressure drop, retention time was between 1.5 to 10minutes, GLR varied from 0 to 40 in standard conditions, operating temperature from 65 to 110°C and operational pressure from 10 to 50 bar. In the case of heavy oils the heating during separation is a common practice. With electrocoalescers operating in the three-phase separator, water can be separated from heavy oil at much lower temperatures. Typically at 60°C the viscosity is above 100mPa s and the oil must be heated up to 100 - 120°C to achieve viscosity reduction down to 10 – 18mPa s. Flow-loop tests for heavy oils are difficult to simulate the real conditions and they are even more challenging if the test conditions include high flow-rates and high shear and viscosity.

The results showed that no separation was observed in the API 17 oil without voltage application. On the other hand, when voltage was applied, the wide range of viscosities influenced only partially the separation performance. Additionally, a fully open choke valve gave the best separation performance, with less than 5% water in oil in just 4 minutes retention time and oil viscosity reaching 80mPa s. It should be noted that in order to achieve the desired pressure loss in the loop a multiphase pump was necessary. However the high shear in the pump created very small droplets that gradually shifted the droplet size distribution in the loop towards smaller diameters. Thus, the emulsions created in the flow loop test are possibly more stable than the ones created during actual production [266].

7. Reduced cost for heating and CO₂ footprint in Hebron oil field, 20.5°API

Hebron is a heavy oil field located offshore Newfoundland and Labrador, Canada. The challenge concerning the production of the 20.5°API crude involved the optimization of the processing equipment with respect to separator size and heat consumption. The limited amount of gas available for heat production led to heating fueled by diesel shipped to the platform and consequently increased cost. Additionally, the very strict weight and size requirements due to limited available space demanded for compact separator sizes.

The production was 270 MBPD, the water cut 52% the arrival temperature of the fluids 55°C and the arrival viscosity 50.7 cP. The VIEC implementation in the process design resulted in a size reduction for the separator vessel from 5m x 29m to 5.4m x 21.5m and an additional decrease in process temperature from 90°C to 60°C. The produced oil quality from the MP separator where VIEC was installed improved from 15% to 10% BS&W. In the following table the conventional design and the VIEC are compared with respect to the required viscosity in coalescer and separator, the watercut in the feed of the coalescer, the heat duty and the CO₂ footprint.

Effect of VIEC®	Conventional	With VIEC®
Required Visc in Separator	15cP	30cP
Required Visc in Coalescer	15cP	15cP
Water Cut to Coalescer	15%	10%
Required heat duty	87.8 MW	29.0 MW
Cost of Heating	\$11.99m/year	\$3.93m/year
CO ₂ footprint	55 M tonn/year	18.1 M tonn/year

Table C10: Hebron design optimization with VIEC implementation in contrast to a conventional design, Robust coalescence technology: Increasing production for existing deepwater facilities, Vessel Internal Electrostatic Coalescer VIEC®, Erik Bjørklund, INTSOK Navigating Deepwater Conference, 2014, [267]

The VIEC installations (Available data up to 2010) besides Troll C field Start-up in July 2003 (Statoil) related to heavy oil treatment are:

- Grane field (Statoil). Start-up in June 2005, API 19. VIEC installation improved separation and increased production.
- FPSO Jubarte P-34 (Petrobras), Start-up in 2006, API 17. VIEC installation improved separation of heavy oil.
- FPSO P-35, Marlim Field (Petrobras), Start-up in 2006, API 20. VIEC installation resulted in cleaner produced water from heavy oil and reduced heating
- FPSO P-37, Marlim Field (Petrobras), Start-up in 2006, API 22. VIEC installation improved separation
- Vincent FPSO (Maersk Contractors/Woodside), Start-up in 2009, API 19 – 21. VIEC installation improved separation and reduced residence time.
- Siri, HT-VIEC®, (TeekayPetrojarl/Petrobras), Start up in 2010, API 12.

[147].

Besides the improved separation efficiency the implementation of VIEC, HT VIEC and VIEC LW in heavy oil production has shown very positive results in terms of cleaner produced water and reduction in demulsifier requirements [144] [268] [269].

- **Compact Electrostatic Coalescer CEC™**

1. Reduction in size and weight of processing equipment for 19°API crude

In order to reduce the size and weight of processing equipment in the case of 19°API crude and flow rate 150000BPD a client has chosen the CEC implementation in conjunction with an upflow type electrostatic coalescer. The results that are in the following table were the smaller equipment size, smaller footprint and weight saving that led to a reduction in the CapEx.

CAPEX	CEC™ System	Upflow type electrostatic coalescer
Separator vessel sizes	Ø 4.2 m x 16 m T-T	Ø 5.9 m x 29m
Footprint	22 m x 6 m	35 m x 6 m
Weight all equipment	BENCHMARK	+ 450t
Total weight savings	BENCHMARK	+1400t

Table C11: CAPEX reduction in terms of size, footprint and equipment for 19°API crude, Offshore processing challenges with heavy oils and use of robust high performance technologies, 2nd Tekna Conference on Heavy Oil Technology for Offshore Applications, Mika Tienhaara, Aker Process Systems Norway, 2011, [154].

2. CEC upstream horizontal flow dehydrator at ExxonMobile, Hebron oil field, 18 - 25°API

Another case study that was offered by Fjords Processing [98] involved the implementation of CEC in Hebron oil field for the ExxonMobile. The produced oil was of 18°API and CEC was installed upstream a horizontal flow dehydrator. The CEC helped the facility to meet export BS&W specifications requiring less space and offered significant weight savings compared to the conventional design. The conventional electrostatic coalescer would have been 30 – 40% longer and around 110 metric tonnes heavier. Additionally CEC can tolerate of up to 40% water cut and still achieve export BS&W specifications.

3. CEC skid with degasser in OSX-1 oil field, 18°API crude

OSX-1 oil field produces heavy oil of 18°API. In this case the vertical degasser vessel is used in order to remove the gas that is generated in the choke and heater. Both the degasser and the CEC are positioned downstream the first stage separator in order to reduce the gas loading and increase the size of water droplets in the water-in-oil emulsion. The fluid downstream the CEC enters into a horizontal flow dehydrator for further treatment. The degasser vessel and the Compact Electrostatic Coalescer help the horizontal flow dehydrator to deliver crude oil within the required BS&W specifications [98].

In the following table selected references of CEC implementation in oil dehydration processes are demonstrated and almost half of them concern heavy oil production.

Client	Field	API
Statoil/Petrojarl	Glitne	32
Chevron/KBR	Captain	19
Mærsk O&G Qatar	Al Shaheen	22-24
Det Norske	Alvheim	46
Petrobras	Jubarte	14/19
OSX	OSX1	18
BP	Quad204 x 2	-
Statoil	Kårstø-Gudrun	32
ExxonMobil	Prolab test unit	20
Toyo	Jacos x 2	n.a
ExxonMobil	Hebron	18-25

Table C12: CEC implementation in oil fields. The red squares denote the fields that produce heavy crude oil. (published with permission of Fjords Processing)

- **3D Horizontal Flow Electrostatic Coalescer**

1. 19°API dehydration in Offshore platform with 3D Horizontal Electrostatic treaters in split flow configuration (2 x 50%)

A split flow configuration of 2 x 50% 3D electrostatic dehydrators was chosen. This configuration includes one central emulsion inlet and two end outlets in each side, for oil and water, as it is shown in Figure C13. Its nominal capacity was 200.000BOPD (1300m³/hr) and its dimensions were 4.6m ID and 22.0m T/T length (Tangent to Tangent length).

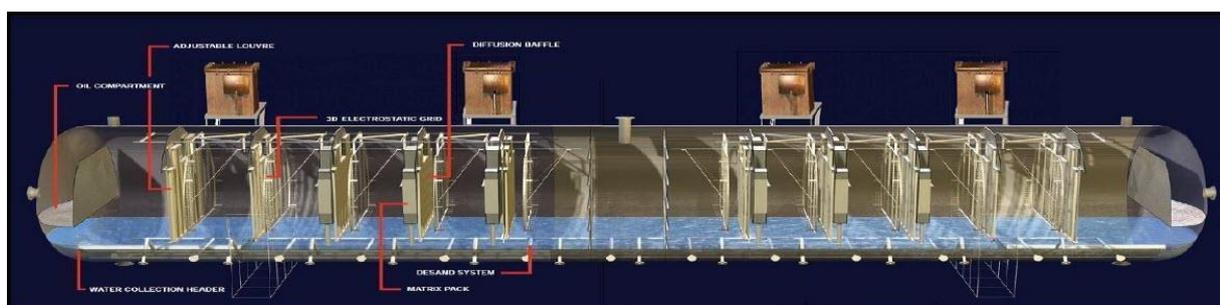


Figure C13: 2 x 50% split flow configuration for the 3D Horizontal Flow Coalescer, (published with permission of Fjords Processing)

The performance of the split flow 3D electrocoalescer was optimized with the use of 5 adjustable electrostatic grid assemblies in each direction, adjustable louvered baffles and matrix packs. The voltage control for the electrostatic grid was adjusted to the water cut in a low/high range between 12 and 25kV and the field strength was controlled by grid space adjustments ranging from 3' to 12'.

The split flow 3D Horizontal Flow treater could operate with flow rates of 1300m³/hr at 92°C and produce less than 0.5% BS&W. At 1750 m³/hr and 100°C it could still meet the required BS&W specifications during six years of continuous production. After the loss of two transformers the 3D electrocoalescer was able to deliver the same BS&W specification with 1750 m³/hr at 100°C by operating at increased voltage.

2. 16°API dehydration in Onshore facility with 3D Horizontal Flow Electrostatic treater

In this case a 3D Horizontal Flow Electrostatic treater with nominal capacity of 19.300BOPD in 120°C was chosen to dehydrate a 16°API heavy crude oil. The client had initially chosen a conventional upflow electrocoalescer of 14' x 110' s/s but later turned to the horizontal flow electrocoalscer due to its ability to handle free and entrained gas and because it could meet BS&W specifications at maximum capacity even with transformer or grid failures.

The feed had 10 – 20% BS&W and a horizontal flow electrocoalescer of 12' x 70' s/s (more compact than the alternative solution) managed to produce at less than 0.5% BS&W. The configuration included 3 adjustable electrostatic grid assemblies, each with own 37.5KVA transformer (Figure C14).

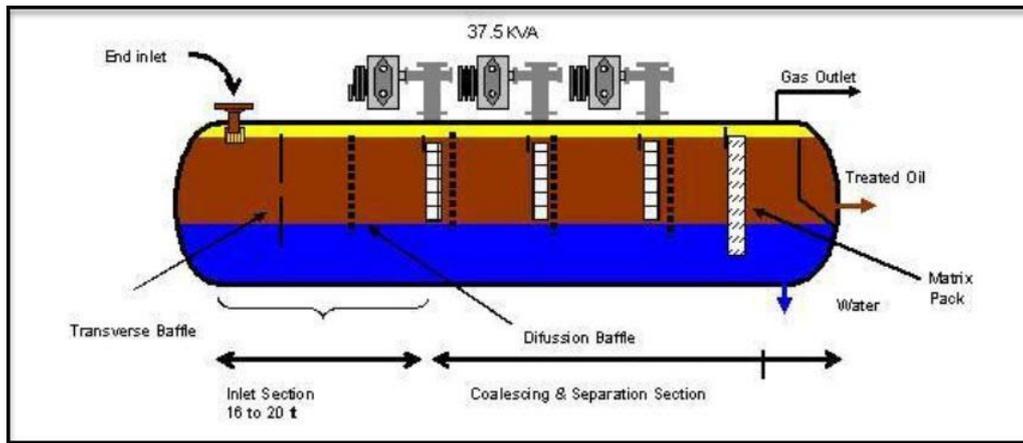


Figure C14: 3D Horizontal Flow Electrostatic treater for 16°API dehydration in Onshore facility, (published with permission of Fjords Processing)

3. Conversion of FPSO Separation system to Heavy Oil

In a Brazilian field an operating FPSO had to be converted from processing light crude of 32°API to heavy crude oil of 17°API. The objective was to reuse as much existing equipment as possible except from the third stage electrostatic treater that was going to be replaced by a 3D Horizontal Flow Electrocoalescer (Desalter). The initial process equipment consisted of a first stage three-phase separator, a second stage two-phase separator (Degasser) and a third stage conventional upflow electrostatic unit (Dehydrator). The difficulties were the limited reservoir data, the uncertain design basis and the fact that desalting was required.

The process solution included a combination of CEC with degasser upstream the second stage separator and a 3D Horizontal Flow Electrocoalescer (Desalter) at the third stage of separation. Additionally the second stage separator was retrofitted with internals adequate for heavy oil dehydration. The schematic dehydration process is in the figure below.

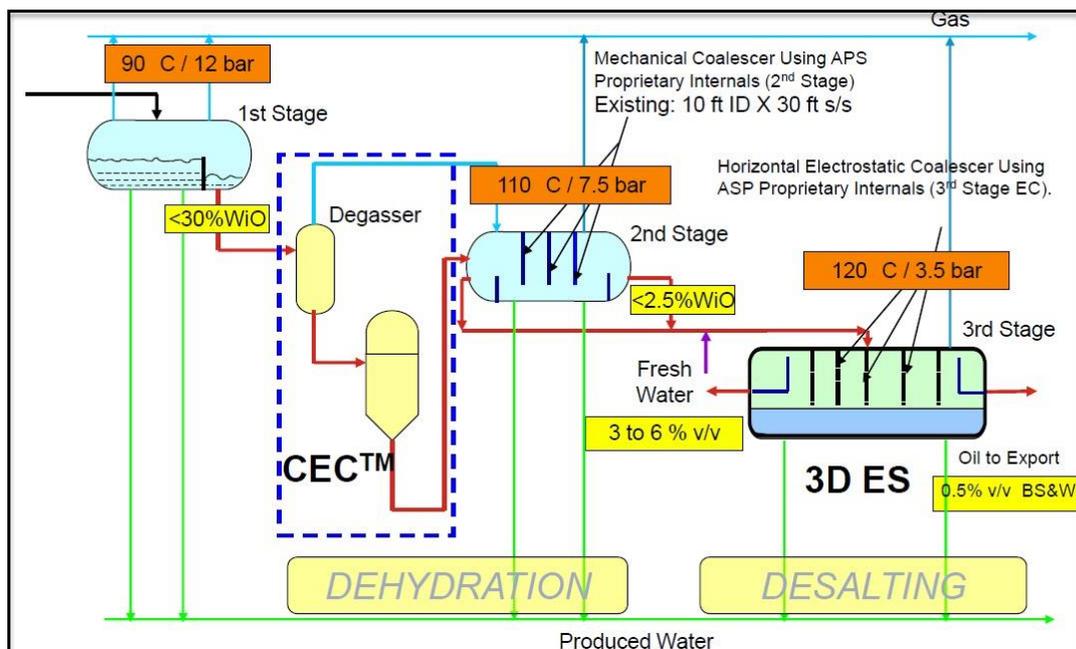


Figure C15: CEC, Degasser, second stage retrofit and 3D Horizontal Flow Electrostatic Treater in FPSO, (published with permission of Fjords Processing)

The production data were 40.000BOPD flow rate, water cut up to 30%, GOR 21 scf/sbbl and 17°API crude oil density. The compact degasser, CEC and skid for the revamping project are presented in the following table.

	SKID	CEC™	Compact Degasser
Length, width	5x5 m	3.5m T/ T	3m T/ P
Diameter Or height	H: 11m	Ø 30 in	Ø 30 in

Table C13: Dimensions of skid, CEC and compact degasser for the FPSO modification project, (published with permission of Fjords Processing)

The project had to be flexible due to the limited reservoir data and in order to manage potential risks. Flexibility was introduced by the following factors:

- The CEC is fairly independent from the operating temperature and it is proven to break stable emulsions that conventional electrostatics cannot handle
- CEC and 3D horizontal flow electrostatic coalescer can handle high water cuts.
- Separation rating has been based on a relatively low operating temperature of 120°C, although the equipment was designed to operate at 140°C or even higher temperatures in case of heavier inlet.
- The 3D horizontal Flow electrostatic coalescer is designed for three-phase operation and therefore it can tolerate gas.

APPENDIX D

PARTIAL UPGRADING TECHNOLOGIES

- **Aquaconversion**

Up to the year 2001 all the heavy oil production in eastern Venezuela was diluted before transferred and traded. For this reason PDVSA-Intevep has developed the Aquaconversion process since 1991. In brief, Aquaconversion is a catalytic steam conversion process which can upgrade an 8°API extra heavy crude oil into a 16°API synthetic crude by transferring hydrogen atoms from water molecules. The resulting syncrude has relatively low viscosity which is accomplished with a rather low investment cost and without the need for a diluent.

In 2001, eastern Venezuela and Orinoco Belt were the main areas of heavy and extra heavy oil production. At that time the Morichal area produced about 40000bbl/day of heavy and extra heavy crude from which the 32000barrels/day were Pilon crude with API gravity 13 and the 8000 barrels/day were Morichal and Jobo extra heavy crudes of 7.5 and 10°API respectively. These two crudes were diluted to 13°API with Leona 21°API gravity crude. The dilution was done in the well to facilitate the lifting of the crude to the surface where it went through a process including degasification, dehydration and desalting. After processing the diluted 13.5 - 16°API crude was commercialized.

The main restriction of the dilution procedure was the availability of the light to medium crudes used as diluents and this led to the development of the Aquaconversion technology. The main target of PDVSA was to increase Venezuela's heavy and extra heavy oil production significantly and to overcome the constraints imposed by the potential shortcoming of light crudes that were required as diluent.

A schematic of the upgrade facilities, integrated with the surface production facilities, is shown in the following figure.

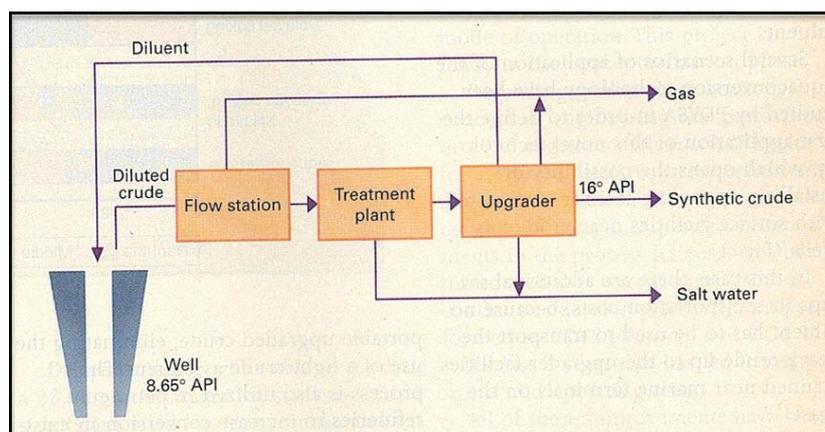


Figure D1: Heavy crude production scheme incorporating on-site upgrading, Aquaconversion technology offers added value to E. Venezuela synthetic crude oil production, Pedro Pereira, Cauri Flores, Hugo Zbinden, Jose Guitian, Rodolfo B. Solari, Howard Feintuch, Dan Gillis, *Oil & Gas Journal*, May 14, 2001, p.79-85, [270].

The integration of upgrading and process facilities offers additional savings in transportation costs since there is no use of diluent in order to transport the heavy crude to the main upgrading facilities located near marine terminals on the coast.

Aquaconversion can be also used in petroleum refineries to increase conversion in the existing visbreaking units. As it was mentioned before, this technology is a catalytic steam conversion process that transfers hydrogen atoms from water vapor into the unconverted residues of the crude and it increases the stability of the process by avoiding the formation of polycondensates, which act as coke precursors. The reactions involved are a thermal cracking mechanism to produce free radicals of hydrocarbons, a catalytic formation of free radicals of hydrogen (nickel-based catalyst) and hydrogen addition to the hydrocarbons' free radicals to avoid polymerization reactions and formation of polycondensate compounds [271] [225]. The process scheme is similar to a conventional visbreaker with a coil furnace and a soaker reactor plus product fractionation. The difference is that Aquaconversion needs larger volumes of water steam and requires the addition of catalytic components and a catalyst recovery section which is widely used in refineries [272]. The process chemistry and flowchart as well as some applications that were foreseen at the time are presented in the following figure.

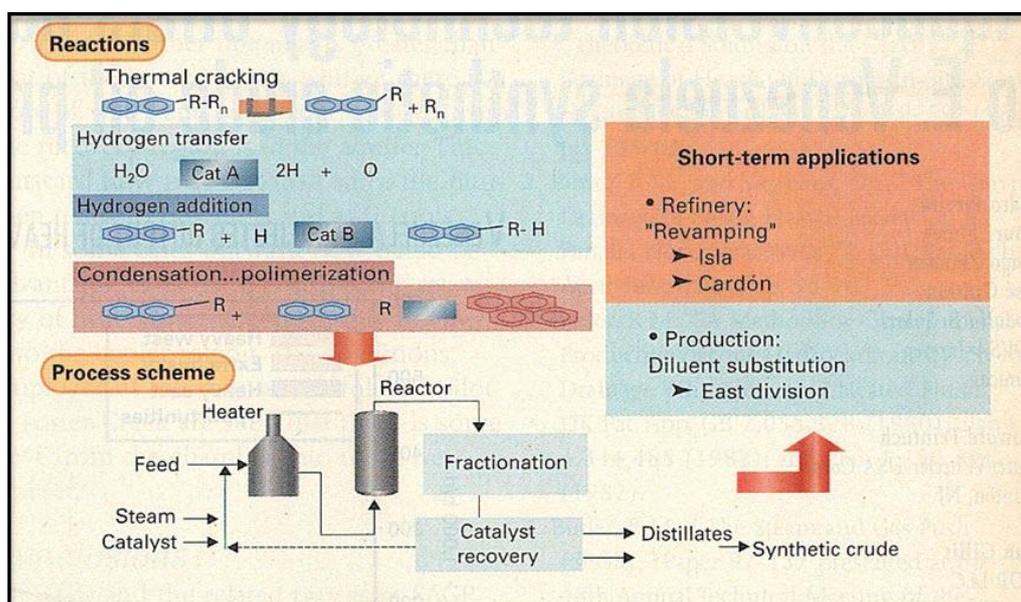


Figure D2: Chemical reactions, process scheme and scheduled applications of Aquaconversion technology, Aquaconversion technology offers added value to E. Venezuela synthetic crude oil production, Pedro Pereira, Cauri Flores, Hugo Zbinden, Jose Guitian, Rodolfo B. Solari, Howard Feintuch, Dan Gillis, *Oil & Gas Journal*, May 14, 2001, p.79-85, [270].

The advantages of Aquaconversion are significant compared to the thermal conversion processes of visbreaking and delayed coking. Visbreaking can produce a 11.9°API from a 8.4°API crude with lower yield compared with Aquaconversion that produces a 16°API crude with a much better product distribution. Additionally delayed coking produces a 28.4°API crude but with a much lower yield due to coke and gas formation. Even if the delayed coking product is partially mixed with the feed 8.4°API crude to produce 16°API upgraded crude (same API as the Aquaconversion product) the yield will still be significantly less.

After pilot plant development and two successful commercial short-term tests from 1991 to 1997 [273] the revamping of Isla refinery was performed in 1998 with UOP, LLC and Foster Wheeler colicensors of the technology. Other refinery applications as well as wellhead applications were designed by PDVSA in the period between 1998 and 2000 taking into account some important improvements in the process scheme aiming to reduce capital and operating costs.

PDVSA's business planning was to reach 5.2 million bbls/day of total production capacity up to 2009 with Aquaconversion technology and without relying exclusively on light crude oil availability for commercializing heavy crudes. However, no further information was found concerning the results.

- **Heavy-to-Light (HTL) upgrading process**

This technology was originally developed in 1998 as the Rapid Thermal Process (RTP) by Ensyn Group of Canada, which was acquired in 2005 by Ivanhoe Energy of the United States [225]. The HTL is a small-scale field-based heavy oil upgrading process which converts heavy viscous oil into a lighter and more valuable product that meets pipeline specifications, without any requirement for diluent.

The HTL plant can be located at or near the wellhead and it is integrated with the surface processing facilities. It can upgrade heavy oil from onshore or offshore locations and provide an upgraded product that has limited coking tendency [219].

The HTL process is based on fluid coking but it is essentially a visbreaking technology. It utilizes a circulating transport bed of hot sand to heat the heavy feedstock and convert it to lighter products. The feedstock is sprayed on the circulating hot sand at very high temperatures and for a very short residence time. Under these circumstances the heavy molecules crack into lighter molecules and produce a liquid of lower viscosity along with some coke. All the coke that is formed on the sand surface is burned in a reheater to heat the sand which is then recycled into the reactor. The lower viscosity upgraded product is quenched at the exit of the reactor cyclone and is led to the atmospheric distillation unit. The distillate and lighter materials are then sent to the product tank for blending with Gas oils. The residue of the atmospheric distillation tower is routed to the front end of a vacuum distillation tower to separate Vacuum gas oil and lighter materials. The Vacuum tower residue can be used for the energy demands of the plant. The flow process scheme is shown in the following figure.

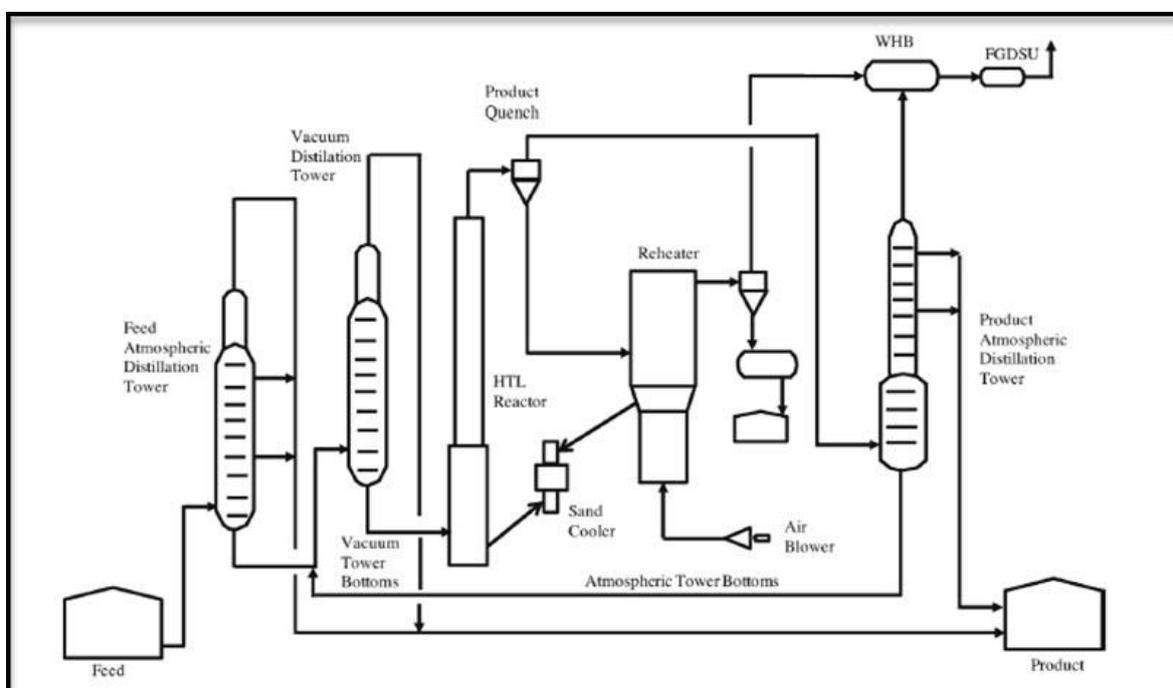


Figure D3: HTL flow process, E. Veith, *Releasing the value of heavy oil and bitumen: HTL upgrading of heavy to light oil*, World heavy oil conference, paper 727, Beijing, China, 2006, [274], E. Koshka, J. Kuhach, E. Veith, *World Heavy Oil Congress*, vol. 329. Edmonton, Alberta, 2008, [275].

Ivanhoe energy claimed that an integration of HTL technology into the topsides of a conventional FPSO was feasible for upgrading heavy crude oils in offshore locations. This would result in an offshore Floating Production, Upgrading, Storage, and Offloading facility (FPUSO). The FPUSO could be energy self-sufficient because the by-products from upgrading can provide energy that is utilized onboard the ship and at the well sites both for control and for powering Electrical Submerged Pumps (ESPs). The schematic illustration of a FPUSO is shown in the following figure.

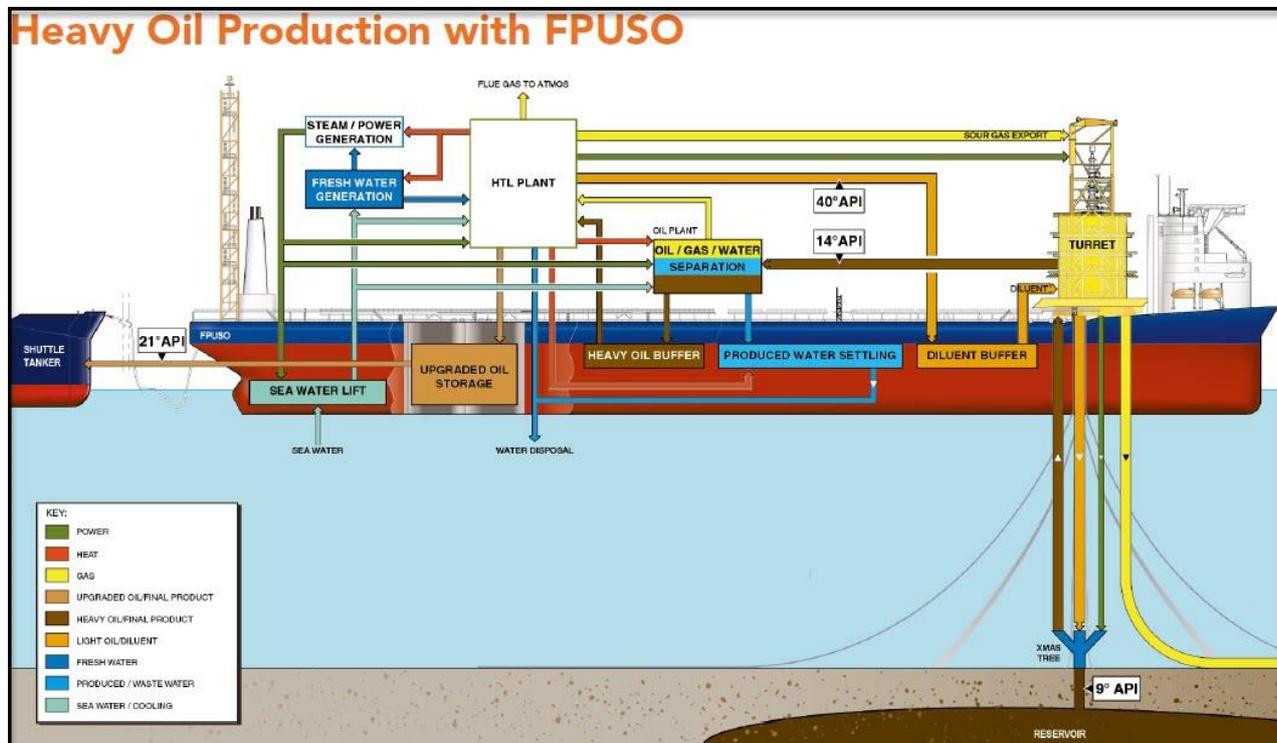


Figure D4: Schematic illustration of a FPUSO, Ivanhoe Energy, FPUSO PRODUCES & UPGRADES HEAVY OIL OFFSHORE, product flyer, 2014, [276].

The main disadvantages of HTL technology are the large size of the equipment, the relatively low volumetric yield of the upgraded oil, the high formation of coke, the low reduction of sulfur content, the high concentration of olefins and diolefins in the upgraded product that are susceptible to polymerization and the production of sand waste which is rich in heavy metals, fine particles and ash.

The HTL process is economic at feed capacities of 10 to 30 Mbbls/day and the properties of the HTL product compared with the heavy oil feedstock are shown in the following table.

Property	Athabasca bitumen	Upgraded crude oil
API gravity	8.5	18.8
Kinematic viscosity		
@40 °C, cSt	>23,000	23
@100 °C, cSt	161	11
Sulfur, wt%	5.02	2.91
Nitrogen, wt%	0.66	0.40
Nickel, ppm	79	15
Vanadium, ppm	209	27
Residue, 1000 °F+, wt%	52	6
Composition, volume%		
Naphtha, IBP–375 °F	0	2.5
Kerosene, 375–455 °F	1	4.5
Distillate, 455–650 °F	12	24
VGO, 650–1000 °F	35	63
Residue, 1000 °F+	52	6

Table D1: Properties of the feedstock and the product of the HTL process, Current situation of emerging technologies for upgrading of heavy oils, Luis C. Castaneda, José A.D. Munoz, Jorge Ancheyta, *Catalysis Today* 220–222 (2014) 248–273, [219].

- **Genoil process**

This is a catalytic hydroconversion technology developed in May 1998 by Genoil [277]. The purpose was to upgrade heavy crude oil on-site to meet the pipeline specifications even for small-scaled field units of 10000 bbls/day capacity.

Genoil hydroconversion technology upgrades heavy crude oil, bitumen and refinery residues. The conversion capacity of the GHU is estimated around 70 – 90% based on various types of feedstocks. The integration of a distillation unit after the GHU and the routing of the residue to a synthesis gas unit achieves an API gravity increase from 24 to 34°. The processing scheme consists of a fixed-bed reactor system with a series of reactors; HDM guard reactors are followed by a reactor using HDS or a combination of HDS and HDN beds. A schematic illustration of GHU is given in the figure below.

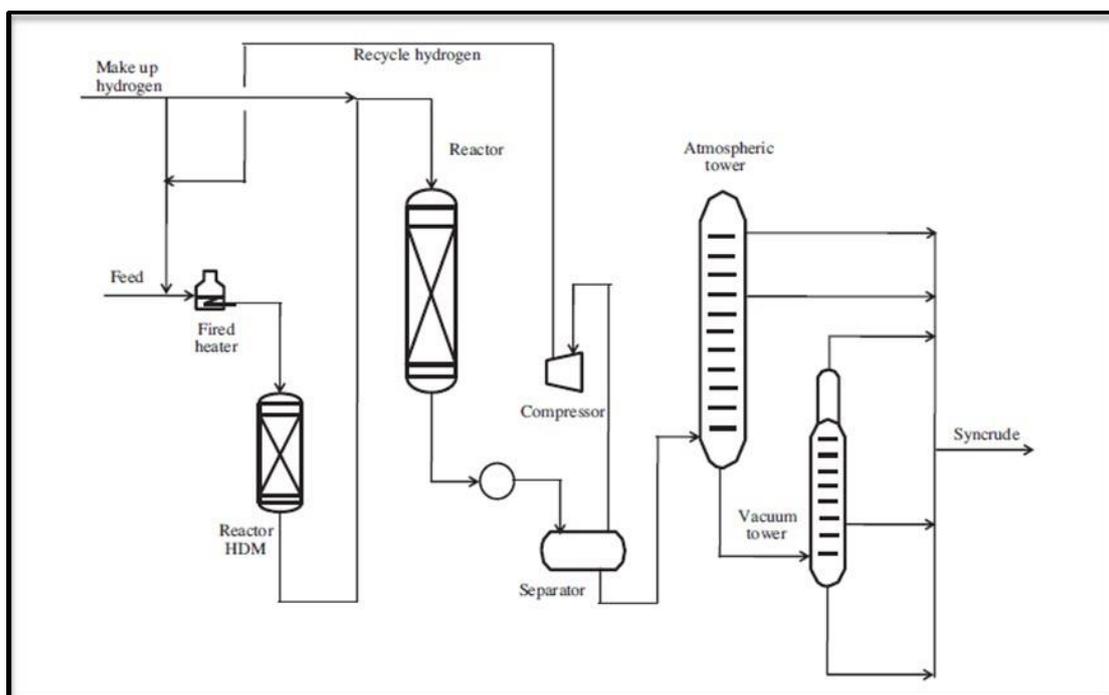


Figure D5: Scheme of Genoil heavy oil upgrader, Current situation of emerging technologies for upgrading of heavy oils, Luis C. Castaneda, José A.D. Munoz, Jorge Ancheyta, Catalysis Today 220– 222 (2014) 248– 273, [219].

The Genoil hydroconversion technology uses extra hydrogen in the feedstock. Genoil reported that the hydrogen was produced from water by electrolysis. In a possible commercial scale plant Genoil will introduce a combination of fixed-bed and moving-bed reactors [225].

The properties of a bitumen feedstock from Western Canada and the crude produced by GHU process are shown in the following table.

	Bitumen feed	Syncrude
API gravity	8.5	24.8
Sulfur, wt%	5.14	0.24
Nitrogen, wt	0.27	0.14
C ₅ Asphaltenes, wt%	17.3	1.6
C ₇ Asphaltenes, wt%	12.6	1.2
CCR, wt%	12.8	2.6
Composition		
IBP– 340 °F	0.0	8.7
340– 450 °F	2.0	11.5
450– 649 °F	12.4	33.0
649– 975 °F	32.3	36.7
975 °F	53.3	10.1
Process results		
-API increase	16.3	
%HDS	95	
%HDN	48	
CCR conversion, %	80	
nC ₇ asphaltenes conversion, %	90	
975 °F+ conversion, %	81	

Table D2: Properties of the feedstock and product of the Genoil process, P. Chung, T. Bugg, Genoil Hydroconversion Upgrader (GHU®)-High Sulphur Heavy Crude and VTB/ATB Residue Upgrading-Engineering Technology for the Future, 2008, [278].

- **Viscositor process**

The Viscositor technology was initially developed by the University of Science and Technology in Norway and the University of Oslo. The owners of this technology are Ellycrack A/S (Norway) and Westcorp Energy Inc. (Canada) and responsible for the continuous research and technological development of Viscositor is SINTEF Energy Research, Trondheim, Norway [279].

Viscositor technology is a process that upgrades heavy oil at the field and does not require any catalyst. Hot fine grained sand is heated in a reactor by coke combustion and is transferred into a “collision” riser by the combustion gases. Pre-heated steam atomized heavy oil (steam under pressure that breaks the oil into droplets in order to expose as much oil surface as possible to the heat) is injected into the riser and collide with the sand particles causing instant evaporation and cracking. Afterwards the generated coke and the solids are separated by a cyclone from the stream. The coke is led to a generator to fuel the process and the solids used as heat carriers are routed again into the reactor and riser in a loop. The oil gas and non-condensable gases are routed to a dual condensation system (Figure D6).

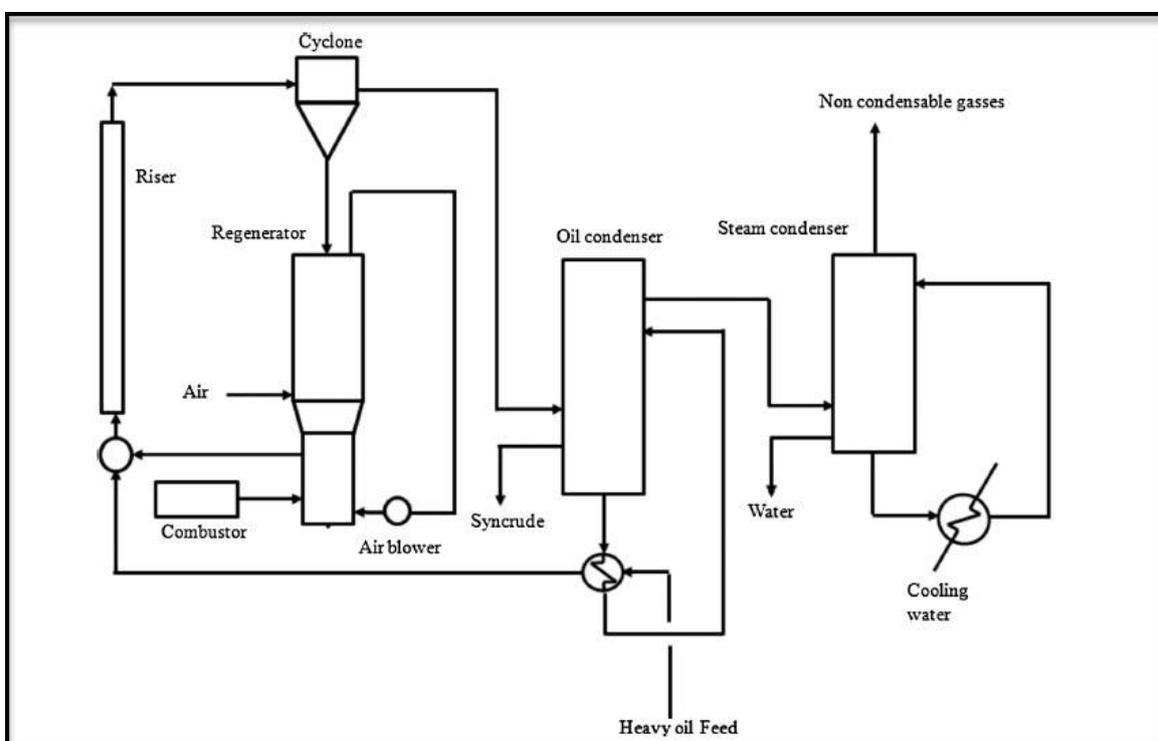


Figure D6: Scheme of Viscositor process, Ellycrack-on-site heavy oil upgrading technology-Technology preview, Oslo, October 2011, [280].

This procedure is self sustained with the energy provided by the combustion of coke and it does not require an advanced catalyst. The metal reduction is close to 90%, sulfur reduction between 50 – 60% and also 5% hydrogenation is achieved. The 2007 testing program yielded the results shown in the following table.

Feed	Initial API	Upgraded API
Pit oil from Venezuela, PDVSA	6.2	22.5
Crude from Venezuela, PDVSA	9.3	20
Blended oil from Venezuela, PDVSA	14.9	21.5
Blended oil diluted from Venezuela, PDVSA	14.9	29.3
Crude from NORSK HYDRO	18	29.5
Canadian crude SUNCOR	13.18	25.2

Table D3: Tests performed with different crude oils using the Viscositor process, Current situation of emerging technologies for upgrading of heavy oils, Luis C. Castaneda, José A.D. Munoz, Jorge Ancheyta, *Catalysis Today* 220–222 (2014) 248– 273, [219].

Besides the aforementioned partial upgrading processes there are other promising technologies such as:

Roseflow: This is a solvent based upgrading technology that utilizes a first separator in which the feed and solvent are introduced and the asphaltenes precipitate as solids and a second separator in which the deasphalted oil and solvent are routed and separated. The recovered solvent is recycled. The first separator works in a subcritical condition whereas the second separator works in supercritical condition. The separation is affected by the temperature, pressure and the type of solvent. The process is licenced by KBR [225].

IYQ upgrading: This is a technology developed by ETX, Calgary, it is based in fluidized-bed coking and utilizes a horizontal reactor. The bed is fluidized perpendicularly but it moves horizontally entering from the one end of the reactor and exiting from the other. The company claims that this reactor design provides 9% by volume higher yield than a delayed coker with less coke formation and lower gas yields.

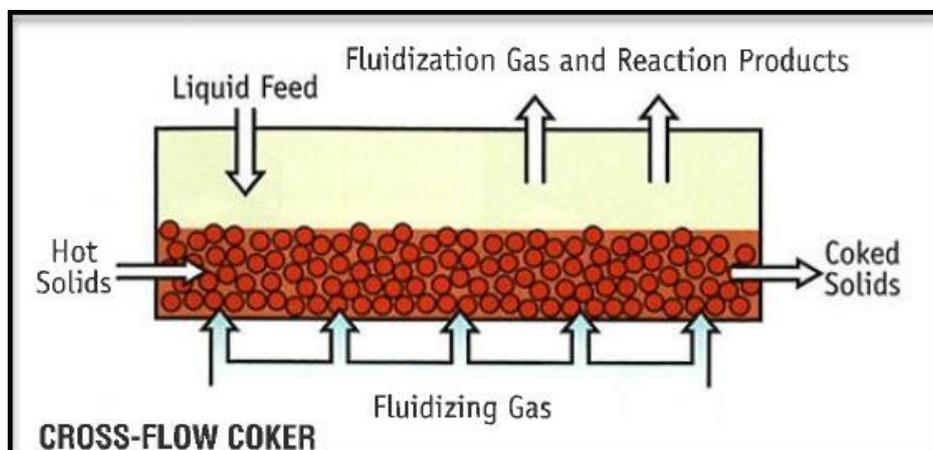


Figure D7: Cross flow coker in IYQ upgrading process,source: www.ntm.nickles.com, *New Technology Magazine*, 2008, pp. 25 – 28, [281].

The following table lists the field upgrading technologies that were analyzed by Kline & Company, Inc. in 2014.

Process route	Bench scale	Pilot	Demo/commercial
Carbon rejection			
Thermal	HOUP write	CPJ EADIEMAC CCU	Aquaconversion HTL Vicositor
Solvent			Roseflow
Thermal and solvent	TRU	HI-Q Value creation	
Extraction	Selex-ASP		
Hydrogen addition	Wildcatter Chattanooga	Genoil Ceramatec	IMP-HRH ENI, H-Oil
Ultrasound	Pristec	Petrosonic Sonocracking	
Other	Petrobeam	I ² Q	Jetshear

Table D4: Emerging technologies for partial upgrading on site, Partial Upgrading of Heavy Oil: Evaluation of Commercial and Emerging Technologies, Report #YIVA01.60, 2014 Kline & Company, Inc, [239].

Despite the efforts through private communication with Kline & Company, Inc it was impossible to get access to the specific report. However, further information on several partial upgrading technologies with potential on-site application have been found in a technical paper published in 2014 [219].

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