

Debottlenecking Existing Offshore Production Facilities To Safely Extend Their Operation Capacities

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Summary

For most early oil-production facilities, higher oil production is expected in the initial days of field life because of the high reservoir pressure. In this specific case study, oil production has reached the floating production, storage, and offloading (FPSO) unit topside-design limit of 100,000 BOPD (662 m³/h) with only five wells connected. Most of these wells were choked to between 30 and 50% to limit the oil production within the installed nameplate capacity of the FPSO unit. Therefore, to take advantage of the low gas/oil ratio (GOR) and the high dry-oil production during early field life, the client has requested this study be undertaken to identify the maximum oil production without compromising process safety and while avoiding major modifications.

The main bottlenecks to increasing crude-oil production to greater than the design capacity are identified as

- Increased temperature resulting in nonstabilized crude: On the basis of the oil composition, stabilization of the crude before leaving the topside is an issue. With an increase in production and the limitation on utility systems, there is not enough heat available to flash off light components and/or cool down the crude before it enters the cargo tank to prevent any further flashing. These flashed-off gases will be vented through the cargo-vent system, along with displaced gases (which are significantly higher than flash gases). The cargo-vent system is designed for vent load during cargo-terminal loading, and, therefore, crude-flash rates will not be anywhere near the design capacity of the vent system.
- Limitations of utility systems: As expected, all utility systems are designed for 100,000-BLPD (662-m³/h) liquid production, and any increase in production will impact the utility balance. To accommodate for additional flow, either modifications are required to the existing system (e.g., additional duty, exchanger modifications) or optimization of the existing system is required.
- Separator capacity: As long as water cut is negligible and a small quantity of water can be allowed to settle in the cargo tank, current crude-oil production has the potential to be increased with the separator operating as a two-phase separator.

A potential crude-oil-production rate of 120,000 BOPD (795 m³/h) can be achieved immediately, without any modifications, if slightly-higher-temperature crude oil is allowed, along with some additional hydrocarbon venting of the gases, from the cargo vent to the atmosphere. Changes in process parameters (pressure and temperature) will allow optimization of the utilities and will allow the crude-storage specifications to be met without any physical modifications.

The actual field trial did match with the study results, and production was increased to greater than the design capacity without many process-stability issues. Therefore, this type of study provides a quick but thorough method of investigating the way forward to improve production without compromising safety limits, allowing the operator to take full advantage of favorable reservoir performance to optimize field economics.

Introduction

During early field life, higher oil production is possible because of higher reservoir pressure and lower GOR. In most offshore facilities, the gas-handling system, compressor, and dehydration system are the governing scenarios or capacity bottlenecks. However, if the GOR is low during early field life, additional oil can be produced during that period without exceeding the gas capacity if the liquid-handling process and associated utility system can handle additional oil.

In this specific case study, oil production has reached the FPSO-unit topside-design limit of 100,000 BOPD (662 m³/h) with only five wells connected. Most of these wells have been choked to between 30 and 50% to maintain oil production within the installed capacity of the FPSO unit. To take advantage of the low GOR and high dry-oil production during early field life, the client has requested to undertake this study.

The objectives of this debottlenecking study are to address the following requirements without the need for hardware modifications (or, if required, with minimal hardware modifications) and to determine operational methods to accommodate the high production rate of early field life (and, hence, to generate the commensurate revenues).

1. Determine the maximum achievable production capacity (not to exceed 120%) without compromising process safety/integrity and while minimizing environmental impact.
2. Identify limiting/bottleneck factors and equipment theoretical limits.
3. Formulate suggestions to overcome these limits and to determine new limits.

Study Assumptions

The following assumptions were determined for the study, to simplify the results and to achieve better clarity:

1. The composition of the production fluid does not vary considerably, and the carbon dioxide (CO₂) or hydrogen sulfide (H₂S) levels are within the existing design limit.
2. Water production is negligible (less than 5%).
3. The maximum gas production is within the nameplate capacity of the existing facility (i.e., the GOR is below the design value); therefore, the study is to be focused on debottlenecking of the oil-processing side only.
4. There is no change in the crude-oil export specifications [i.e., the Reids vapor pressure=10 psia (0.69 bara) at 37.8°C and the crude-oil export temperature ranges from 35 to 55°C].

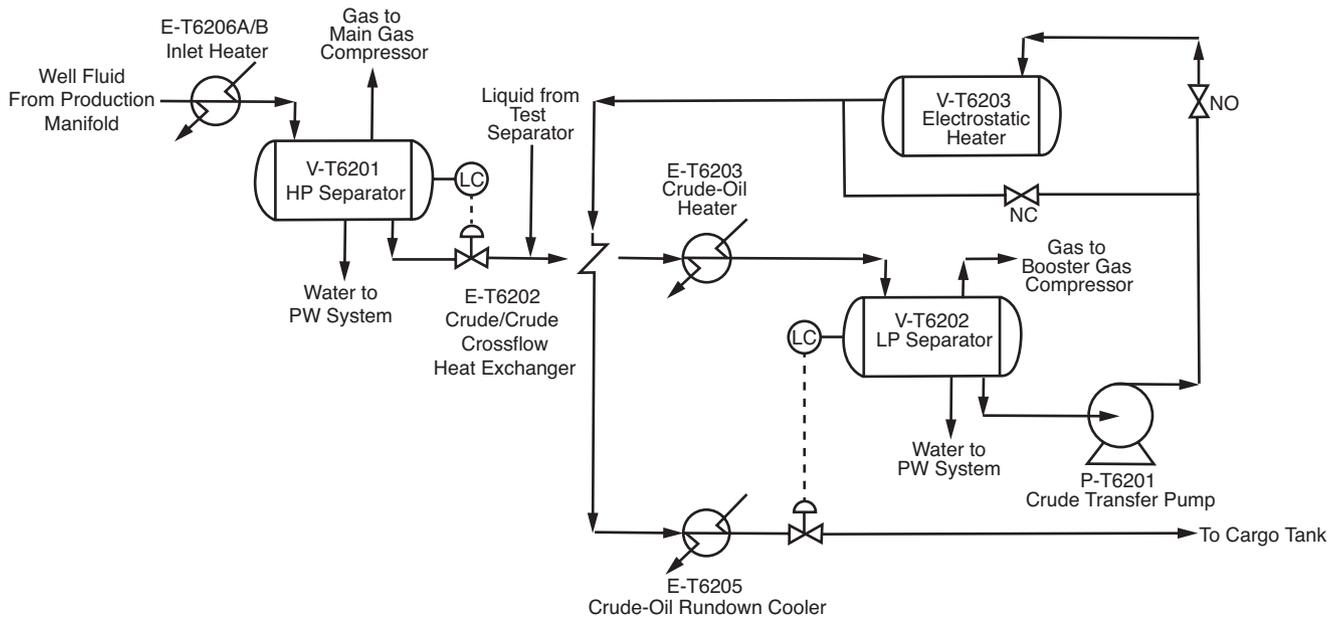


Fig. 1—Simplified oil-processing flow chart for liquid processing. (LC=level controller, NC=normally closed, and NO=normally open.)

5. There is negligible (or manageable) pressure-drop increase in the exchanger and in the piping with an increase in flow.
6. All utilities, but primarily the heating medium (HM) and the cooling medium (CM), are operating per their design specification/design duty, and no spare capacity is available in those systems.
7. The environmental impact associated with additional venting/flaring is excluded from this study. This was assessed by the client.
8. This paper deals with process bottlenecks. When implementing debottlenecking processes, it is important to assess the integrity envelope of the facilities that are affected;

however, the integrity-envelope aspect is not covered by this paper.

Simulation

The topside process simulations were performed with the Aspen HYSYS® (2011) simulation package by use of the Peng-Robinson equation of state. As per the current production profile, five wells were connected to the production manifold from three different fields. The current flow rates from these wells were provided by the production support team and were equal to a total rate of 100,000 BOPD (662 m³/h). In agreement with the production-support team and for simulation purposes, it is assumed that the

TABLE 1—DESIGN PARAMETERS OF EXISTING OIL-PROCESSING HEAT EXCHANGERS

Equipment Tag Number	Title	Design Duty (kW)	HM/CM	HM/CM		Process Side		Observation
				Inlet Flow of HM/CM (kg/h)	Inlet/Outlet Temperature of HM/CM (°C)	Inlet Flow of Process Fluid (kg/h)	Inlet/Outlet Temperature of Process Fluid (°C)	
E-T6206 A/B (2×50%)	Crude-oil inlet heater—S&T*	21 274 (per unit)	Inhibited fresh water (HM)	325 552	160/105	368 246	42/95.1	10% surface margin included in design duty
E-T6202 (1×100%)	Crude/crude crossflow heat exchanger—P&F**	5355	Process Fluid	590 800	115.1/100.5	587 100	95.3/110.1	Extension capacity available to add 52 plates (8.5% additional duty)
E-T6203 (1×100%)	Crude-oil heater—P&F**	4324	Inhibited fresh water (low-temperature HM)	182 400	140/120	551 600	106.2/116.9	18% surface margin included in design duty Extension capacity available to add 45 plates (50% duty)
E-T6205 (1×100%)	Crude-oil rundown cooler—P&F**	15 885	Inhibited fresh water (CM)	500 454	33/60.33	600 523	101/55	10% surface margin included in design duty Extension capacity available to add 68 plates (20% duty)

*S&T = shell and tube.
**P&F = plate and frame.

increase in flow from these wells will be proportional (to current flow rates).

Existing System

The main factors that govern the oil-processing capacity of the FPSO unit are the separators, the heat exchangers, the crude-oil pump, the level-control valves, the oil-metering system, and the associated pressure safety valves (PSVs). **Fig. 1** provides a simplified flow chart of these systems.

The high-pressure (HP) separator and low-pressure (LP) separator are designed to operate as three-phase separators. The HP separator is originally sized to handle rates of 110,000 BLPD (729 m³/h) and 110 MMscf/D of gas. The LP separator was sized for rates of 100,000 BLPD and 12 MMscf/D of flashed gas. The specifications used to size the separators were 0.1-gal/(MMscf/D) liquid in gas and 1,000 ppm oil in water. The water-in-oil specification is 10 and 5% for HP and LP separators, respectively. The separators were sized such that a minimum of 5 minutes of residence time is available between the normal interface level (NIL) and the normal liquid level (NLL) and a minimum of 30 seconds of residence time is available between the NLL and the high liquid level (HLL).

The existing system also includes four exchangers on the liquid side—the inlet heater, the crude/crude crossflow heat exchanger, the crude-oil heater, and the crude-oil rundown cooler. The design parameters of the existing exchangers are listed in **Table 1**.

The inlet heater is an S&T-type and is sized on the basis of the maximum water-cut case. The expected flow pattern with high water cut will be oil dispersed in the water phase. In that scenario, because of high water cut, the expected emulsion viscosity is very high, as is the specific heat of the crude mixture [3.65 kJ/(kg·°C)]. The crude/crude crossflow heat exchanger is a P&F-type heat exchanger in which the stream leaving the HP separator receives the heat from the stream leaving the electrostatic treater, which is operated at 115°C.

The crude-oil heater is also a P&F-type heat exchanger, and its main purpose is to adjust the temperature of the stream leaving the crude-oil exchanger (E-T6202) to 116°C to flash the gases. This will stabilize the crude oil traveling to the cargo tank (by means of the rundown cooler). All the P&F-type exchangers have the capability to add an additional number of plates, which should help to increase the exchanger duty.

The crude-oil transfer pump is provided to develop the head required to overcome the system pressure drop between the LP separator and the crude cargo (piping, exchanger, electrostatic treater). The available net positive suction head (NPSH) can sometimes pose an issue because the liquid leaving the LP separator is at its vapor pressure. An increase in flow will result in additional pressure drop (between the LP-separator liquid nozzle and the pump), which may affect the NPSH available to the crude-oil transfer pump. However, in this particular case, enough of a margin is open between the NPSH available and the NPSH required. Hence, this is not considered a bottleneck for further study. However, in similar studies on other facilities, this was noted as an issue that could be resolved with various options, including a change in impellers.

The design valve-sizing coefficient of the HP and LP separators level-control valve is 1,422 and 3,000, respectively, with an equal-percentage flow characteristic. The current valve opening for these control valves is 50 and 40% for HP and LP separators, respectively; therefore, a great deal of margin is available on these valves. Because these valves can handle much higher flow, they are not considered a bottleneck for further increase in flow. However, for some other project studies, it was noticed that these level-control valves can be bottlenecked, and, hence, care should be taken when performing similar studies.

For crude-oil-flow measurement, at any given time, two out of three turbine-type flow elements are available. Each flow element is designed for a maximum flow rate of 113,056 BLPD (749

m³/h), with a design temperature of 100°C; therefore, the maximum crude that can be measured at any time is 226,102 BLPD (2×749=1498 m³/h). With these results, the crude-oil meter is not considered a bottleneck for this study.

Once the crude enters into the cargo tank, it will displace the blanketing gas, which will be vented by means of the cargo vents. The existing vent system is designed for vent load during cargo-terminal loading (i.e., 6875 m³/h for automatic venting and 4000 m³/h for manual venting).

Processing Limits of the Separators

On this particular FPSO unit, the HP and LP separators are designed for three-phase separation—oil, water, and gas. The key parameter for the separation is oil and water residence time provided within the vessel. According to American Petroleum Institute (API) Specification (SPEC) 12J (2008), for crude >35°API, the minimum required oil residence time (considering separation length within the separator) for a three-phase separator is 3 minutes, and for crude <35°API, the required oil residence time is 5 minutes. Also, the time between normal operating level and alarm level and between alarm level and trip level shall not be less than 30 seconds or 150 mm—whichever is greater—for both high and low ranges. Please note that the 3 to 5 minutes of residence time is a good estimate when no data are available for the emulsion tendencies of the oil. If oil tends to make a tight emulsion, longer residence times (and/or chemical addition) may be required to achieve the desired separation.

The design basis for the existing-separator crude is 18 to 31°API, meaning 5 minutes of residence time for three-phase separation. However, according to recent field data, the crude density is less than 850 kg/m³ (i.e., more than 35°API). Therefore, a separation time of 3 minutes is sufficient for HP and LP separators as long as crude gravity is more than 35°API. In later field life, with change in the crude density and increase in the water cut, the original 5 minutes of residence time needs to be reinstated for better separations.

Tables 2 and 3 show the residence times available for the HP and LP separators, respectively, over the range of oil-processing capacities (on the basis of the final dead crude oil). This additional crude processing is performed without exceeding the design capacity for gas processing downstream (i.e., taking advantage of lower GOR). Therefore, considering the overall HP- and LP-separator system, an additional 20% [up to 120,000 BOPD (795 m³/h)] of oil can be processed without any modifications to the HP and LP separators. On the basis of the overall integrity of the field (mainly from the flow-assurance point of view), the client requested to limit the increase in production to 20%; hence, further increase in the production was not studied.

Additional Higher Liquid Processing. During the early phase of field production, water cut is negligible and the actual requirement is for two-phase (oil/gas) separation. However, a small quantity of water can settle at the bottom of the cargo tank during the storage period. If these separators are used for two-phase (oil/gas) separation only, then the separation-time requirements will be reduced dramatically (to 2 minutes or less), meaning these separators can process a greater amount of liquid without any internal modifications. Once water cut increases, the original design duties can be resumed. However, two-phase separation was not taken into consideration for this study because the targeted 20% increase in oil processing was achieved by operating separators as three-phase separators to reduce the risk of overlooking the water breakthrough between well-testing periods.

Test Separator. As the liquid from the test separator is routed to the LP separator, the test separator can be used to process additional liquid. On the basis of the separation-time requirements mentioned in the preceding subsection, the maximum amounts of liquid (oil/water) and gas that can be handled by the test separator are

TABLE 2—HP-SEPARATOR LIQUID-HANDLING CAPACITY

*Parameter	Dead-Crude Flow Rate (100,000 BOPD)	Dead-Crude Flow Rate (105,000 BOPD)	Dead-Crude Flow Rate (110,000 BOPD)	Dead-Crude Flow Rate (115,000 BOPD)	Dead-Crude Flow Rate (120,000 BOPD)	Required Per Design Standard	
Separation time between NIL and NLL (seconds)	300	285	275	265	250	180	
Separation time between NLL and HLL (seconds)	40	40	35	35	30	30	
Oil-outlet velocity (m/s)	1.19	1.33	1.32	1.37	1.43	< 2	
Gas-outlet velocity (m/s)						16.1**	—
Momentum criterion (ρV_s^2) for gas-outlet nozzle (Pa)						1751	≤ 3,750

*These separation times are based on the separation length (the length between the weir and the inlet baffle). If taking into consideration the tan-to-tan length of the separator, the holdup time will be higher.
 **Gas production is assumed to be at maximum per the design case and is managed by well management.

TABLE 3—LP-SEPARATOR LIQUID-HANDLING CAPACITY

Parameter	Dead-Crude Flow Rate (100,000 BOPD)	Dead-Crude Flow Rate (105,000 BOPD)	Dead-Crude Flow Rate (110,000 BOPD)	Dead-Crude Flow Rate (115,000 BOPD)	Dead-Crude Flow Rate (120,000 BOPD)	Required Per Design Standard
Separation time between NIL and NLL (seconds)	300	290	275	260	240	180
Separation time between NLL and HLL (seconds)	50	40	40	40	40	30
Oil-outlet velocity (m/s)	1.1	1.2	1.33	1.39	1.45	< 2
Gas-outlet velocity (m/s)	29.5*	30.9*	5.84	6.11	6.37	—
Momentum criterion (ρV_s^2) for gas-outlet nozzle (Pa)	2844*		98**	107**	116 [§]	≤ 3,750

*Based on highest gas-flow rate of 12 MMscf/D.
 **Based on simulated flash gas-flow rate.

41,000 BLPD and 54 MMscf/D, respectively. However, the maximum amount of gas that can be separated in the test separator is 50 MMscf/D because of the erosion velocity and the nozzle size of the gas outlet.

In line with standard industry guidelines, the ρV_s^2 for the gas nozzle should be less than 3750 Pa. Any further increase in gas-flow rate will increase the momentum applied to the nozzle and can affect the performance of the demister. Therefore, it is recommended to operate the test separator at less than 41,000 BLPD and 50-MMscf/D gas-flow rate, with the total gas on the platform limited to the emergency gas-handling capacity of flare, which is 121 MMscf/D (by means of well management).

Electrostatic Treater. The electrostatic treater is designed for a flux ratio of 115 B/D/ft². The current performance guarantee for the electrostatic treater is provided for 100,000 BOPD at a minimum water cut of 3% to achieve 0.5% of basic sediments and water (BS&W) and the salt specification of 500 mg of salt/L for offloaded crude oil.

For the increased-flow case, per vendor feedback, the salt specification can be met provided that the washwater-flow rate is increased to remove the additional salt. This additional water is available from an existing source. However, per vendor feedback and owing to the flux-ratio limitation (on the basis of transformer sizing), if the flow is increased by 10%, the BS&W will increase by 0.1% and if the flow is increased by 20%, the expected BS&W will be 0.7%, which, compared with the required 0.5%, would be an approximately 0.2% increase in BS&W. The additional settling time is available in the hull tanks during storage, which will facilitate the gravitational separation of the water from the oil. Consid-

ering the relatively high API gravity and zero emulsion, it is fully expected that the retention time in the cargo tanks will be sufficient to achieve BS&W reduction.

On the basis of this information, it was agreed that an electrostatic treater should not be considered as a bottleneck for this FPSO unit (one of the advantages of FPSO). Otherwise, capital modifications would be required to meet BS&W requirements. Additionally, during early field life, if water cut is less than 5%, the electrostatic treater can be bypassed because cargo can provide residence time for oil/water separation.

Processing Limits of the Liquid-Side Heat Exchanger

The following subsections comprise our observations of the performance of the heat exchangers made during the trial to achieve the targeted production of 120,000 BOPD. **Table 4** provides the revised oil-processing limits of the heat exchangers on the basis of these observations. All calculated values are based on the theoretical composition of crude. Any changes in flow and to the crude characteristics will result in changes to the heat-transfer characteristics. Similarly, any changes in the temperature profile will result in a change to the driving force for the heat exchange. Therefore, the duty for the heat exchanger will not remain exactly the same. The effects of this are relatively small and, for the purpose of this study, will be further neglected.

Inlet Heater (E-T6206). The original design case for this S&T-type heat exchanger is the maximum water case (75% water and 25% oil). In that scenario, because of the high water cut, the expected emulsion viscosity is very high, as is the specific heat of the crude mixture [3.65 kJ/(kg·°C)]. Considering this high spe-

TABLE 4—REVISED DUTIES OF EXISTING OIL-PROCESSING HEAT EXCHANGERS

Equipment Tag Number	Title	Design Duty (kW)	HM/CM	HM/CM		Process Side		Observation
				Inlet Flow of HM/CM (kg/h)	Inlet/Outlet Temperature of HM/CM (°C)	Inlet Flow of Process Fluid (kg/h)	Inlet/Outlet Temperature of Process Fluid (°C)	
E-T6206 A/B (2×50%)	Crude-oil inlet heater—S&T	21 274 (per unit)	Inhibited fresh water (HM)	325 552	160/105	428 629	42/95	No issues.
E-T6202 (1×100%)	Crude/crude crossflow heat exchanger—P&F	5355	Process fluid	590 800	115.1/100.5	725 511	95/108	This is a process/process fluid heat exchanger followed by other exchangers. Duties can be compensated by those exchangers. No issues.
E-T6203 (1×100%)	Crude-oil heater—P&F	4324	Inhibited fresh water (low-temperature HM)	182 400	140/120	725 511	106.2/116.4	
E-T6205 (1×100%)	Crude-oil rundown cooler—P&F	15 885	Inhibited fresh water (CM)	500 454	33/60.33	724 660	103.1/65	Without altering the utility balance, the outlet temperature of crude will be 65°C, which is 10°C higher than the required temperature of 55°C.

cific-heat value [3.65 kJ/(kg·°C)] for the oil/water emulsion, the achieved temperature for increased process flow rate is 90°C, as compared with the required temperature of 95.1°C.

However, according to the current production profile, water production is almost negligible (i.e., dry-oil production). Therefore, the current oil viscosity and the specific heat [2.1 kJ/(kg·°C) per simulation] are expected to be much less than in the governing scenario. With current oil characteristics, the required HM-flow rate will be much less compared with the design flow rate; hence, the temperature of 95°C for the stream leaving the heat exchanger is achievable. Therefore, this exchanger is not a bottleneck.

Crude/Crude Crossflow Heat Exchanger (E-T6202). In this P&F-type heat exchanger, one process stream exchanges heat with another process stream. Therefore, a change in one stream will result in a change to the other process stream. Consistent with the original design duty of 5355 kW, the expected temperature for the stream leaving the exchanger is 110.1°C for cold stream and 100.5°C for hot stream. For a revised increased flow rate of 120,000 BOPD (795 m³/h), the calculated values are 108 and 103.1°C for the cold and hot streams, respectively. This means that the stream receiving heat is approximately 2°C colder than expected. This drop in temperature can be compensated in the heat exchanger (E-T6203), which is in series with this exchanger, if duty is available.

On the other side of the exchanger, the stream that is cooling down will have a temperature rise of 2.6°C (103.1°C compared with the original 100.5°C), which will affect the duty of the crude-oil rundown cooler (E-T6205).

Crude-Oil Heater (E-T6203). The existing design duty of 4324 kW is suitable to provide the increased flow rate, mainly because of the lower specific heat of the dry crude. The design basis for this heater was high water cut, meaning higher specific heat. Therefore, no issues or concerns were identified for this exchanger.

Crude-Oil Rundown Cooler (E-T6205). With a design cooling duty of 15.89 MW, the temperature of the process stream leaving this exchanger is calculated as 65°C (assuming there is no change

on the utility side). This crude-oil rundown cooler is identified as one of the bottlenecks to enhancing oil production.

To check the real limit of this cooler, other simulation cases with reduced flow were observed. For a flow rate of 105,000 BOPD, the temperature of the process stream leaving the cooler was 57°C; for a flow rate of 100,000 BOPD, the temperature was 54.5°C; and for a flow rate of 102,000 BOPD, the temperature was 55°C. This indicates that the current capacity of this cooler is a bottleneck to increasing production to greater than 102,000 BOPD. Note that during the trial run with production of 105,000 BOPD, the reported temperature of the crude-oil rundown cooler was 57°C, which matches the theoretically calculated temperature of 57°C.

Debottlenecking for Higher Crude-Oil Production

From the preceding subsections, it is clear that to increase the production to greater than the design limit, the real bottleneck is the crude-oil temperature leaving the topside. But is it really a bottleneck? Can a higher temperature be allowed for crude traveling from the topside to the cargo storage tank? If not, how can the bottleneck be corrected? There are three options for debottlenecking in this scenario:

- Allow a higher temperature for crude leaving the topside.
- Allow operational changes only (without additional utility demand).
- Allow equipment modifications.

Allow a Higher Temperature for Crude Leaving the Topside. According to the project specification, “CONTRACTOR shall limit exported oil temperature through export hoses, from a minimum of 35°C to a maximum of 55°C, to comply with shuttle tankers requirements” and “Export oil Reid vapour Pressure of 10 psia @ 37.8°C” (Gaidhani and Hollaar 2013).

During the design phase, calculations were performed to determine the offloading temperature of the crude from the crude cargo tank. These calculations were performed on the assumption that the produced crude oil would enter into the crude storage tank at 55°C. Because the crude cargo tank is in contact with the ocean, it is expected that the crude entering the tank will be cooled as a result

TABLE 5—VARIOUS OPERATING SETS TO OPTIMIZE THE OPERATIONS

	HP-Separator Operating Conditions	LP-Separator Operating Conditions	Crude-Oil Temperature Leaving Topside (°C)	Crude Oil Produced (BOPD)	Flash Gases in Cargo (m ³ /h)	Comments
Design Case	1000 kPa at 95°C	350 kPa at 115°C	55	100,000	0	Base-design case; however, with current actual-crude composition from various fields, some flashing is expected with current operating parameters (< 40 m ³ /h).
Option 1	1000 kPa at 95°C	350 kPa at 115°C	62.5	115,000	197	All existing process parameters other than crude-oil temperature will be retained. In case of high water cut, no change will be required. Heavier-crude separation can be handled.
Option 2	1000 kPa at 95°C	350 kPa at 115°C	65	120,000	222	Considering current heating-system duty availability, light crude and higher crude up to 120,000 BOPD can be produced without exceeding the heating-duty requirement. However, crude exit temperatures will be higher and additional cooling-system modifications will be required to reduce the temperatures.
Option 3	1000 kPa at 80°C	350 kPa at 100°C	55	120,000	405	Change in HP- and LP-separator operating temperatures will reduce the heating- and cooling-duty requirements. Meets crude storage-temperature requirement. Once the desired heavy-crude/water separation is reached, temperature decrease might be needed to reinstate to the original values.
Option 4	1000 kPa at 80°C	300 kPa at 100°C	55	120,000	210	All advantages of Option 3, plus reduces cargo-vent gases. Saved utilities can be used for additional liquid processing, if separator capacity allows.
Option 5	1000 kPa at 80°C	300 kPa at 100°C	55	Approximately 130,000	> 221	Good only until negligible water cut (two-phase separation). Once heavier crude produced and oil/water separation is required, production cut will be needed to allow separation-time requirements of the separators.

of the ship-bottomplate cooling from seawater and the ship-deck cooling from the ambient air.

It was concluded on the basis of these cooling calculations that the offloading temperature would be approximately 35°C, with a maximum of 39°C at the highest temperature condition (during peak summer). In this case, the produced-oil temperature leaving the topside was calculated at 65°C for a higher oil-production rate of 120,000 BOPD, with the maximum temperature of the offloading crude from the cargo tank being approximately 49°C (39°C+change of 10°C–heat losses), which is still below the maximum allowable export temperature per the specifications.

The cargo tank (bottom only) is coated with Interbond 808, a high-performance epoxy coating that is suitable for crude-oil application up to 70°C. The zinc anodes in these tanks are suitable for high temperatures; therefore, the relatively high-temperature crude will not be an issue for these tanks because the crude will soon be cooled down either by mixing with the cold fluid already in the tank or by heat transfer with the ocean.

If the aforementioned temperature of the crude traveling to storage and the additional gas venting are acceptable, then the crude-oil production can potentially be increased to 120,000 BOPD without any additional requirements or modifications, with the assumption that the gas produced is still within the design capacity of the FPSO unit. In this scenario, the real bottleneck will be the processing capacities of the separators. However, the real concern resulting from the higher crude temperature is the amount of vapor generated in the cargo tank, which will be greater than the normally expected vent load described in the next subsection.

Allow Operational Changes Only (Without Additional Utility Demand). High temperature in the separators is required for good oil/water separation and to stabilize the crude by vaporizing the light ends. With negligible water cut, the required temperature for

the HP separator, the LP separator, and the electrostatic treater can be reduced from the current operating temperatures of 95°C (for the HP separator) and 115°C (for the LP separator and electrostatic treater), but this is possible only if slight flashing of crude is allowed to the cargo tank. This will reduce the requirements of the additional heating and cooling duties.

Instead of increasing the temperature, the operating pressure of the LP separator can be reduced slightly to allow for additional flashing of the light ends of the crude, which will help to stabilize the crude and reduce or eliminate the crude flashing in cargo tanks. The booster gas compressor is sized with a design margin of 30%; hence, there is enough room available in the booster gas compressor to accommodate the additional flash gases.

As mentioned previously, by retaining the current HP and LP separator temperatures and pressures, the temperature of the crude reaching the cargo tank will be 65°C. The specification of 55°C can be achieved with a reduction in the process temperature. This will increase the flashed gases generated in the cargo tanks, but will reduce the heating- and cooling-duty requirements, which are currently operating at their limit.

On the basis of the Aspen HYSYS (2011) simulation results, various sets of operating scenarios were developed. Assuming that three-phase separation is required, the first three operating scenarios are recommended for increased production; however, selection of the operating set depends on the operation requirements, the acceptance of high-temperature crude in the cargo tank, the amount of venting allowed, and the oil/water-separation requirements. Table 5 provides an indication of the operating parameters to be used to find the optimum conditions in the field.

Option 1 allows for a production rate of 115,000 BOPD and does not require any modifications to water cut or heavy-crude processing. Option 2 also allows for processing of the heavier crude, but requires an increase in the water cut and creates a need for ad-

ditional heating duties. The drawback to both of these options is higher crude temperature, which can be resolved by making a modification to the cooling system.

When taking the current crude profile into consideration, Options 3 and 4 are the better scenarios because they allow for separation of water in the separators. Option 5 is the best scenario for two-phase separation in HP and LP separators because it allows for much higher crude handling and takes into account dry oil and the knowledge that there is no real requirement for the removal of small quantities of water (if any exists) from the liquid. Once the water cut increases, the FPSO unit can be operated according to Option 1 or 2. However, physical modification to the utility system, heat exchangers, and separators can allow for further increase in crude production, which is suitable for higher water cut or heavy-crude processing.

It should not be forgotten that the real concern stemming from these scenarios is unstable crude entering into the cargo tank, which will then be flashed in the cargo tank. Because the cargo tanks are blanketed, the produced crude oil will displace the gases, which will also be vented, and those displaced gases will have a much higher rate (>700 m³/h) than the flash gases. In spite of this, the amount of vapor generated by flashing and displacement will be much lower than the vent-system design capacity, and this should not be a concern.

Note that the concern about flashing crude is based on the original crude composition, which is heavy crude. The fact is that the current produced crude is light crude, and, as such, we expect fewer issues in meeting the vapor-pressure specifications and expect a lower quantity of vent gases in the cargo tank.

Equipment Modifications. As noted in the preceding, some of the exchangers installed on the FPSO unit can be modified to enhance the heat-transfer capability to overcome the limitation of high-temperature crude and to stabilize the crude completely (i.e., zero flashing in the cargo tank). However, this option will require the process to be shut down, which will result in production and revenue losses. Therefore, the viability of this option depends on local environmental regulations and the duration of the higher production.

- Crude/crude crossflow heat exchanger (E-T6202): This P&F-type heat exchanger has an extension capacity of 52 additional plates, providing an increase of 8% to the heating duty.
- Crude-oil heater (E-T6203): This P&F-type heat exchanger has an extension capacity of 45 additional plates, providing an increase of 50% to the heating duty.
- Crude-oil rundown cooler (E-T6205): This P&F-type cooler has an extension capacity of 68 additional plates, providing an additional 20% to the cooling duty.

Any changes in the duties of the exchanger either will result in a change in the return temperature of the HM/CM or will require an additional amount of medium fluid. Both of these options will result in a change in the utility balance. As stated previously, the utility system was assumed to be running at design capacity. The recent operating data show that most of the utility-control valves are fully open, although water cut is zero. Therefore, the utility balance is the bottleneck for further increase in production. To match the increase in production, a more-detailed review of the utility system is required to optimize the parameters, such as increase in export-gas temperature, which will allow for additional CM availability for the liquid system.

Concerns and Limitations

Destabilized Crude. The stabilization of crude oil (i.e., making sure the true vapor pressure is less than atmospheric) is the main issue in increasing the production of crude oil to greater than the nameplate capacity. With increased throughput and limitation on the utility system, additional flashing may be expected in the cargo

tank. This flashed-off gas will be vented through the cargo-vent system along with the displaced gas. This vent system is designed for vent load during cargo-terminal loading; therefore, the flash rates will not be anywhere near the design capacity of the vent system (6875 m³/h for automatic venting and 4000 m³/h for manual venting). The cargo-vent system is a critical safety system because it serves as the ultimate protection against cargo overpressurization, which is the reason the design pressure of the cargo-storage tanks is only 120 kPa absolute. The actual amount of flashing is very sensitive to the actual composition and conditions. However, it is important to maintain awareness of this phenomenon because it may not become obvious from any of the instruments installed on the FPSO unit.

Crude Characteristics. Currently, the crude oil is light (>35°API). However, once the water cut increases or a well with heavy crude becomes tied in, the separation of oil and water will require longer residence time in the separators. This will reduce the processing capacity of the separators, which may result in a return to the design capacity of 100,000 BOPD in those separators. Further evaluation of these separators is required in consultation with the separator internal vendors. There have been various new technologies developed in the last 5 years that might allow producers to retrofit new internals into the separators to increase the oil/water/gas separation capacity. As suggested in the preceding sections, if water cut is low and some water tolerance is accepted in the cargo tank, the separators can be used for two-phase (oil/gas) separation only. This will reduce the residence-time requirements, and a much greater amount of liquid can be processed in the same vessels.

In general, it is very hard to theoretically predict separation efficiency. It is therefore recommended to monitor the separation efficiency in the field and adjust the production rate as needed to accomplish acceptable water/oil separation.

Water Cut. Currently, water cut is almost negligible, which leaves the specific heat of the crude at approximately 2 to 2.2 kJ/(kg·°C) and the viscosity dominated by the crude oil. Once the water cut reaches greater than 10%, the crude-oil/water specific heat will increase along with the emulsion viscosity, which will be much higher than the crude-only scenario. Both these factors will alter the heat-transfer duty of most exchangers.

Corrosion Inhibitor. A corrosion inhibitor is added to the flowlines/risers because the material used in the pipes leading to the HP separator is carbon steel, which is likely to corrode in the presence of water and CO₂. Corrosion inhibitor is not effective when the velocity of the fluid in the pipe increases to greater than 15 m/s. Considering the current low/negligible water cut, this is not an immediate concern, but as the water cut increases, it will be necessary to make sure that two-phase velocity in the riser and flowlines is limited to 15 m/s (and is below the erosion limit of the material).

Process Safety

Separator PSV. The governing case for the HP-separator PSV is the “closed gas outlet.” For the LP separator and the electrostatic treater, the governing case is the “external fire.” Currently, gas production is much lower than the design capacity because of a lower-than-expected GOR. Therefore, at the start of this study, it was assumed that the gas capacity was unchanged and that operations would maintain the gas flow from wells within the normal gas-processing capacity of the separators for higher liquid flow. Therefore, an increase in the oil flow would not alter the governing scenario for the HP-separator PSV, and, hence, the PSV size.

Similarly, the inventories of the LP separator and the electrostatic treater are the same, with the only change being in volumetric flow rate. Therefore, PSV size in relation to the fire case is still unchanged.

Flare Drums. The flare drums will remove the liquid droplets from the flare gas to prevent liquid carry-over to the flare stack. The sizing of the flare drums is performed according to the method described in API STD 521 (2014), where the horizontal part of the flare drum should be designed for the removal of liquid droplets with a diameter of 300 to 600 μm . Both flare drums are sized for the removal of liquid droplets with a maximum diameter of 500 μm .

HP and LP flare-drum design-sizing criteria are stated as “to handle 5 minutes of the full facility production of oil plus the emergency gas relief rate” (API STD 521 2014). The new scenario, with a crude-production rate of up to 120,000 BOPD, plus the emergency gas-relief rate, is checked for the 5-minute criterion and is found to be within the design limit of the installed flare drum. Similar results are true for the LP flare drum. These flare drums are sized on the basis of the emergency gas load of 121,000 MMscf/D. Therefore, at any given time, the total gas on the platform shall be less than the emergency gas-handling capacity of flare, which is 121 MMscf/D.

Cargo Vent. As mentioned previously, because the crude may not be fully stabilized before entering the cargo tank, slightly increased venting of the cargo gas may be required under loading conditions, which is already required for the displaced-gas to maintain the pressure in the cargo tanks. The cargo-vent system is designed for vent load during cargo-terminal loading; therefore, the flash-gas and displaced-gas rates will be well below the design capacity of the vent system (i.e., 6875 m³/h for automatic venting and 4000 m³/h for manual venting). Hence, this is not a safety concern, but may need environmental attention for additional venting permit.

Conclusion

On the basis of this relatively simple and straightforward desk study, it can be proved that oil production can be increased safely through performance of due intelligence and assessment and without exceeding the gas-processing capacity or requiring significant modification to the facility. The low GOR can be taken advantage of during early field life to produce more oil and to obtain higher revenue.

Despite best efforts to develop an accurate basis of design, the actual production conditions, such as composition, GOR, arrival temperature, and pressure, can still be quite different in most cases. This desk-study provides an initial screening of the production increase that is expected to be achieved in a safe manner with the use of the various options outlined in this paper. In this case, the maximum increase under consideration was 20%.

As discussed in the preceding sections, increasing the oil-production rate by 20% may have consequences on the performance of the facility, such as additional flashing of the crude in cargo tanks, which will result in a slight increase in hydrocarbon venting to the atmosphere. Given the deviation of the actual production conditions compared with the basis of design, changes in performance are expected and may be acceptable as long as they do not compromise the process safety and overall product-export specifications.

Most of the equipment is designed within a margin and with contingencies in place for various phases of the project. This, in combination with the difference from the design basis, makes it very hard to provide exact field-performance prediction from behind the desk, especially when predictions are based on the theoretical simulation model. The exact limits of the facility can be verified by actual field trials only. However, this desk-study provides a very important guidance for the operations that need to be observed and the consequences that need to be monitored. The

study also provides important assurance that all safety systems can deal with the increased oil production, and it provides the limits to be observed during field test, such as maintaining the overall gas capacity to below the emergency flaring limit.

Additionally, this study provides assurance to the field team by defining the constraints to be observed during field testing. It is recommended to increase crude-oil production gradually, by only a few thousand B/D, by allowing dry and low-GOR wells to be opened first. This will allow the utility system to adjust to the increase in demand, and will help to identify the exact requirements for utility-system and equipment modifications. Any unforeseen bottleneck can be evaluated to run the process plant for steady operation while increasing crude production to greater than the name-plate capacity.

The actual plant trial was carried out in the field with an initial production rampup of 115,000 BLPD (762 m³/h) that gradually increased over the next 10 days. All process parameters were stable during that period. However, as expected, the HM/CM system reached a point at which further increase resulted in fluctuations in the process stability. Therefore, no further increase was carried out over 115,000 BLPD (762 m³/h). As forecasted previously, the frequency of the manual cargo vent increased during this period because of flashing of the crude in the cargo tanks.

In summary, this type of study provides a quick but thorough method to investigate the way forward to improve production without compromising safety limits, allowing the operator to take full advantage of favorable reservoir performance to optimize the capital value of the asset.

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