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## Thesis Title

# Risk Assessment of the Gas-Condensate Separator D-401 in the Associated Gas Treatment Unit at the TFT Oil Field


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## Dedication

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## List of Abbreviations, Symbols, and Acronyms

| | |
|---|---|
| **AGTU** | Associated Gas Treatment Unit |
| **ANN** | Artificial Neural Network |
| **ANFIS** | Adaptive Neuro-Fuzzy Inference System |
| **API** | American Petroleum Institute |
| **BLEVE** | Boiling Liquid Expanding Vapor Explosion |
| **BP** | Low Pressure |
| **C1, C2, C3, ... C8** | Hydrocarbon compounds (Methane, Ethane, Propane, etc.) |
| **COMAH** | Control of Major Accident Hazards |
| **$CO_2$** | Carbon Dioxide |
| **CPC** | Central Processing Center |
| **DNV** | Det Norske Veritas (now DNV, a global risk management company) |
| **EIGA** | European Industrial Gases Association |
| **ETA** | Event Tree Analysis |
| **FTA** | Fault Tree Analysis |
| **HAZOP** | Hazard and Operability Study |
| **HP** | High Pressure |
| **IChemE** | Institution of Chemical Engineers |
| **IDLH** | Immediately Dangerous to Life or Health |
| **IEC** | International Electrotechnical Commission |
| **KT** | Turbine Unit Designation |
| **LEL** | Lower Explosive Limit |
| **LOPA** | Layer of Protection Analysis |
| **LPC** | Loss Prevention Council |

| | |
|---|---|
| **$N_2$** | Nitrogen |
| **NOAA** | National Oceanic and Atmospheric Administration |
| **OREDA** | Offshore Reliability Data |
| **OSHA** | Occupational Safety and Health Administration |
| **PCV** | Pressure Control Valve |
| **P&ID** | Piping and Instrumentation Diagram |
| **PFD** | Process Flow Diagram |
| **PG class** | Pasquill-Gifford Atmospheric Stability Class |
| **PHAST** | Process Hazard Analysis Software Tool |
| **PoF** | Probability of Failure |
| **PRV** | Pressure Relief Valve |
| **QRA** | Quantitative Risk Assessment |
| **RBI** | Risk-Based Inspection |
| **SIS** | Safety Instrumented System |
| **TFT** | Tin Fouyé Tabenkort |
| **TCV** | Temperature Control Valve |
| **UDM** | Unified Dispersion Model |
| **UVCE** | Unconfined Vapor Cloud Explosion |
| **VCE** | Vapor Cloud Explosion |
| **WT** | Water Tank |

# GENERAL INTRODUCTION

**Introduction:**

The oil and gas industry is inherently exposed to significant risks due to the handling of flammable, toxic, and high-pressure substances. Ensuring the safety of personnel, the environment, and industrial assets is therefore a critical priority across all stages of design, operation, and maintenance. Major accident hazards such as fires, explosions, and toxic releases remain central concerns, particularly in upstream facilities where complex fluid mixtures are processed under extreme conditions. Among the various safety engineering tools available, risk assessment techniques such as HAZOP (Hazard and Operability Study) and consequence modeling using advanced software like PHAST (Process Hazard Analysis Software Tool) developed by DNV have become indispensable. These methods allow engineers to identify potential process hazards, assess their likelihood and consequences, and define preventive and protective strategies to mitigate their impact.

This final project focuses on a high-pressure separator vessel located within the Associated Gas Treatment Unit (AGTU) at the Tin Fouyé Tabenkort (TFT) site, a major oil field in southern Algeria operated by Sonatrach and its partners. This vessel is designed to separate condensate from associated gas—natural gas that is produced alongside crude oil—and operates under demanding thermodynamic conditions, with a pressure of 80 bar and a temperature of 20 °C. The presence of pressurized flammable liquids and vapors introduces considerable risk. Given the stored energy and hydrocarbon inventory, any mechanical failure—such as a rupture, flange leak, or instrument malfunction—can escalate into serious safety and environmental consequences, including jet fires, pool fires, or vapor cloud explosions, especially in the presence of ignition sources.

The objective of this project is twofold. First, risk identification and analysis is performed using the HAZOP methodology to systematically examine potential deviations from normal operation. This involves a detailed evaluation of each process line and equipment node, considering deviations such as "More pressure", "No flow", or "High level", and assessing their possible causes (e.g., valve failure, instrument drift, human error) and consequences. The study also examines the existing safeguards and proposes recommendations to reinforce safety barriers and minimize residual risk.

Second, consequence modeling is conducted using PHAST DNV to simulate several accidental scenarios, including full vessel rupture and leaks of varying sizes (6 mm, 25 mm, and 100 mm). These scenarios are assessed under a range of realistic atmospheric conditions representing summer and winter, day and night. Meteorological parameters such as wind speed, temperature, atmospheric stability class, humidity, and solar radiation are varied to understand their influence on the behavior of hazardous releases. The simulations predict the thermal radiation intensity from jet and pool fires, the dispersion distance of flammable or toxic clouds, and the duration of hazardous exposure zones. The aim is to quantify the severity of each event and determine safe separation distances, vulnerable zones, and the need for additional risk-reduction measures.

By combining qualitative hazard identification (HAZOP) with quantitative consequence analysis (PHAST), this study provides a comprehensive and integrated risk assessment of the separator vessel.

# Process Hazard Analysis and Risk Evaluation Using HAZOP and CTA

**Introduction:**

According to ISO 31000:2018, risk assessment is a systematic, iterative and collaborative process of identifying all potential hazards and deviations—from loss of containment and overpressure to temperature excursions—that could compromise the safety or reliability of a gas–liquid condensate separator in an associated gas treatment unit; quantifying for each scenario the probability (P) and severity (S) of its consequences—whether equipment damage, environmental release or operational downtime—and calculating a risk index ($R = P \times S$); and then prioritizing these risks against predefined tolerance criteria to select the most effective and cost-optimal mitigation measures, such as engineered barriers, procedural controls or instrument redundancy, thereby ensuring transparency and traceability of every decision in line with organizational and regulatory performance thresholds.

# 1 Hazop hazard and operability and consequence tree analysis

## 1.1 Purpose and Objectives of HAZOP

A Hazard and Operability Study (HAZOP) is a structured, systematic examination of a process or operation to identify potential hazards to personnel, equipment, or the environment, and to uncover operability problems that could impair efficient operation[1]. HAZOP assumes that risk events arise from deviations from design or operating intent. By applying a set of guide words to process parameters (flow, temperature, pressure, level, composition, etc.), the HAZOP team systematically reveals how the process might deviate from its intended function[1]. The main objectives are to:

- **Identify deviations** from design intent, and determine all associated hazards and operability issues[2].
- **Determine controls or actions** needed to manage identified hazards or operability problems, and specify how to implement solutions[3].
- **Flag issues requiring further study**, where immediate conclusions cannot be reached, and decide what additional information or actions are needed[4].
- **Ensure follow-up on agreed actions** so that recommended safeguards are implemented and tracked until closure[2].

- **Increase operator and stakeholder awareness** of process hazards and operability concerns, improving the overall safety culture.

HAZOP is primarily a qualitative, inductive, bottom-up risk assessment tool – often described as a brainstorming technique – relying on subject-matter experts to imagine deviations and their consequences[2]. It is widely accepted in process industries (chemical, petrochemical, oil & gas, pharmaceuticals, etc.) and is a key component of Process Safety Management.

Standards and guidelines (e.g. IEC 61882) emphasize that HAZOP should be conducted by a multidisciplinary team in a critical-thinking atmosphere[1].

The ultimate aim is to enhance design safety by revealing and mitigating hazards early, rather than as an afterthought.

## 1.2 HAZOP Methodology

### 1.2.1 Study Nodes and Design Intent

The first step in a HAZOP is to define the system boundaries and nodes. A node is a discrete section of the process (often a segment of pipe, vessel, or unit operation) for which a clear design intent can be stated[1].Nodes are typically identified on process flow diagrams (PFDs) or piping & instrumentation diagrams (P&IDs); each node's extent is chosen so that its *normal operation* is well-defined (for example, "flow from Pump-101 into Column-201 at X rate and temperature")[1]. The HAZOP team records the design intent or function of each node before analysis. The goal is to ensure nodes are neither too large (so that issues get overlooked) nor too small (creating redundant work)[1].

### 1.2.2 Guide Words and Process Parameters

For each node, the team systematically applies guide words to the relevant process parameters to identify deviations from the design intent[1]. Typical process parameters include flow (or flowrate), pressure, temperature, level, and composition[1]. The basic HAZOP guide words (with their typical interpretations) are shown below[1]:

- **No (None):** Complete negation of the design intent (e.g. "no flow")[1].
- **More:** Quantitative increase (e.g. higher flow or pressure)[1].
- **Less:** Quantitative decrease (e.g. lower flow or level)[1].

- **As well as:** An additional activity (qualitative increase, e.g. an extra reactant)[1].
- **Part of:** Only part of the design intent (qualitative decrease, e.g. "part of the flow")[1].
- **Reverse:** Logical opposite of the intent (e.g. reverse flow)[1].
- **Other than:** Complete substitution or "other" (e.g. the wrong fluid in the stream)[1].

The exact list of guide words may be tailored to the process. IEC 61882 notes that guide words should be chosen to be neither too specific (which might stifle ideas) nor too general (which may lose focus)[1]. In practice, each selected guide word is paired with each relevant parameter at a node (for example, *"No Flow"*, *"More Temperature"*, etc.), and the combination is reviewed for credibility as a deviation[1]. If meaningful, it is recorded as a potential deviation from intent (e.g. "no flow through reactor feed line").

### 1.3 Application Workflow with Examples

In practice, HAZOP is applied at the detailed-design stage or during major modifications, when P&IDs and process documentation are available. The workflow typically starts with a kick-off meeting to define scope (process boundaries, nodes to be studied) and then proceeds with a series of workshop sessions, each lasting a few hours. Successful HAZOPs require good preparation: finalizing the design ("freezing" drawings), distributing documents to team members beforehand, and allocating enough time and resources. For large facilities, this can mean many sessions: for example, a mid-size chemical plant with ~1200 equipment items might require on the order of 40 separate node-review meetings[1].

When applied to oil, gas or petrochemical processes, HAZOP finds a wealth of potential hazards and operability issues. Consider a three-phase gas–oil–water separator in a gas processing unit. In such a vessel, parameters include inlet flow rate, level controls, pressure setpoints, etc. A HAZOP team would examine deviations like "No flow" at the inlet (perhaps caused by a closed feed valve or pump failure[5]) or "More inlet flow" (e.g. pump over-speed).

They would consider what happens if liquid level gets too high ("High level" deviation – possibly caused by level controller failure[5]). Consequences of a high-level deviation might include carry-over of liquids into the gas outlet, overpressure due to suppressed volumes, or even spillage from the weir overflow[5]. The team would note existing safeguards (level alarms, dump valves) and likely recommend additional measures (e.g. high-level shutdown, emergency flaring arrangement) if needed. Each scenario is logged with its causes and effects, so that the separator's design can be made safer and more robust.

More generally, HAZOP is used on reactors, distillation columns, heat exchangers, compressors, pipelines, and even procedural operations in refineries and chemical plants. By methodically forcing "what-if" deviations at each node, HAZOP often uncovers non-obvious failure modes. For example, a HAZOP on a reactor feed might note "No flow – valve closed" leading to reactor run dry, or "Low pressure – pump trip" causing underfeeding and runaway chemistry. The result is a list of design improvements or operating procedures that significantly reduce risk. (As a historical example, one study noted that a "High level" deviation in a storage tank would lead to overfill and a flammable release[5], which in practice triggered installation of a fail-safe high-level trip.)

## 1.4 Documentation Outputs and Action Tracking

The primary output of a HAZOP study is the HAZOP worksheet (or series of worksheets). This is a structured table (paper or electronic) where each row corresponds to one deviation scenario. Key columns typically include: the *Node* name, the *Guide Word/Parameter* (e.g. "No/Flow"), the *Deviation* description, *Possible Causes*, *Consequences*, *Existing Safeguards*, and *Recommended Actions.*

During the workshop, the recorder fills in each entry. According to IEC 61882, the HAZOP "produces minutes or software to record the deviations, their causes, consequences and recommended actions together with marked up drawings"[2]Thus, the worksheets (often backed by markup on the P&ID) become an auditable record of the study. After the sessions, these worksheets are consolidated into a final report which lists all identified hazards and follow-up items.

Crucially, recommended actions must be tracked to completion. Typically, each action is assigned to an individual (often called the "action") with a deadline. Modern HAZOP practices use action-tracking systems or software databases to log and monitor these items.

As noted in literature, specialized tools can assist in managing the HAZOP output and ensuring that "the meeting minutes and tracking [of] recommended actions" are maintained[1]. In large projects, one often maintains an action log or safety issues register arising from the HAZOP, which is reviewed in subsequent project meetings until all safety improvements are implemented.

## 1.5 Strengths and Limitations

### 1.5.1 Strengths

HAZOP’s systematic, team-based approach makes it extremely thorough at identifying process hazards. Its guideword methodology prompts examination of virtually all plausible deviations, yielding a comprehensive hazard list[6]. The use of a multidisciplinary team ensures diverse perspectives – for example, operators may spot issues designers miss – leading to a more balanced analysis[6]. Because HAZOP is qualitative, it is effective even for hazards that are difficult to quantify or predict (such as human errors or complex chemical behavior)[7]. It also does not require extensive prior data; experts use experience and process knowledge to brainstorm outcomes. In essence, HAZOP is a powerful creative *brainstorming* tool (supported by structure) that often catches “hidden” or non-intuitive failure modes. Its widespread use means that its outcomes are well-respected by engineers and regulators, providing documented due diligence on process safety.

### 1.5.2 Limitations

The HAZOP method is resource-intensive. It can require many person-hours of preparation and meeting time, especially on large plants, and thus can be costly and time-consuming[1]. It also relies heavily on the experience and attentiveness of the team; if key expertise is missing or if participants are inattentive, important deviations may be overlooked. The outcome is inherently qualitative and somewhat subjective – HAZOP does not, by itself, quantify likelihood or severity. It focuses on *single* deviations from intent and does not systematically consider multiple simultaneous failures or interactions between nodes[7]. As noted by commentators, HAZOP alone does not assess the effectiveness of existing controls; additional analyses (e.g. LOPA or reliability studies) may be needed for that. Finally, the voluminous documentation (long tables of deviations and technical jargon) can be difficult for non-specialists to digest, and ensuring timely closure of action items requires diligent project management.

Despite these limitations, HAZOP’s benefits in enhancing safety have been judged to outweigh its costs[6]. In many complex industries, HAZOP is still the preferred tool for detailed design-phase hazard analysis precisely because of its thoroughness and adaptability.

## 1.6 Why HAZOP Suits Complex Continuous Processes

HAZOP was originally developed for fluid-processing industries and remains especially well-suited to complex continuous processes[1]. Continuous units like reactors, separators, columns and heat exchangers involve many interdependent parameters and control loops. HAZOP's node-and-parameter approach naturally maps onto such systems: each vessel or pipeline segment can be a node, and its flow, pressure, temperature, level, etc., are parameters. By exhaustively combining guide words with these parameters, HAZOP explores an extensive set of process states.

For example, in a gas–oil–water separator, the HAZOP would examine deviations such as *"No flow at inlet"* (perhaps a blocked line), *"High pressure in vessel"* (valve or control failure), *"Low level (no liquid)"* or *"High level"* (level transmitter or control malfunction). Each deviation is analyzed: a *"High level"* might be caused by a failed level-control valve[5], and its consequence could be liquid carryover or overflow releasing flammable mixture[5]. These insights directly drive design choices (level alarms, relief valves, backup controls) and procedures. In this way, HAZOP turns the complexity of continuous processes into a structured set of "what-if" scenarios.

## 1.7 Consequence Tree Analysis in Industrial Risk Assessment

Consequence Tree Analysis (CTA) is a forward-looking risk modeling technique that maps out the possible sequences of events ("chains of consequences") following an initiating incident. In a consequence tree, a primary event (e.g. a leak, rupture or equipment failure) is placed at the root, and branches represent subsequent events or outcomes. Each node may split via logical OR (alternative paths) or AND (joint prerequisites) relationships. In the words of Schäbe (1994), *"a consequence tree is a set of events logically combined by OR and AND connections that occur in sequence, some being prerequisites for others"*[8]. In practice, one identifies the initiating loss-of-containment (LOC) event and then systematically lists follow-on events (e.g. dispersion, ignition, fire, explosion, toxic release) until all final damage states (e.g. injury, fatality, environmental harm, asset damage) are reached[8].

Constructing a consequence tree typically involves these steps: first define the initiating event (e.g. a pipe rupture, tank overpressure, or valve leak). Next, enumerate all significant intervening events or barriers (e.g. ignition source present/absent, relief valve opening, emergency response) that affect outcomes. Use OR gates where either of multiple events can lead to the same consequence, and AND gates where multiple conditions must coincide. Continue branching until reaching terminal outcomes of interest (such as a fire, explosion, toxic cloud, release cessation, etc.)[8]. For example, EIGA guidelines instruct: *"for each type of LOC … construct the sequence of events that can result from it, stopping with all end events that result in serious harm to people, the environment or property"*[9]. In this way the consequence tree systematically captures all plausible accident sequences (including escalation paths and mitigated outcomes) from a given initiating event.

## 1.8 Identifying and Classifying Consequence Chains

Once constructed, a consequence tree is used to *identify and classify* the chains of consequences from an initiating failure. Analysts examine each branch from the root to a leaf, interpreting it as a distinct scenario. For example, one path might represent "leak → vapor cloud formation → delayed ignition → vapor cloud explosion (VCE) → equipment damage"; another might be "leak → gas dispersion → no ignition → flammable cloud dispersal (no fire)". Each branch is labeled with its outcome (e.g. "BLEVE", "pool fire", "toxic inhalation", "no effect"). This systematic listing ensures no major outcome is overlooked. In quantitative risk assessments (QRA) for hydrocarbon systems, engineers often subdivide the plant into sections and, *"for each section, potential failure scenarios are mapped out… Subsequently, a series of consequence chains is developed for each failure scenario, illustrating the possible hazard outcomes."*[10].

In other words, CTA serves much like an event-tree diagram: it explicitly enumerates all sequences of events after the initiating incident. The analyst "moves forward from the critical event" to list hazards and damages [10]. In complex systems, many branches may arise (tens or even hundreds in large analyses), each representing a unique pathway to harm. For instance, a pipeline rupture could lead to projectile hazards, toxic gas clouds, jet fires, pool fires or VCEs depending on conditions[11]. A good consequence tree will capture all such sequences, often prompting the analyst to distinguish cases (e.g. immediate ignition vs. delayed ignition) and to include mitigating factors (e.g. pressure relief, flame arresters, inerting).

## 1.9 Probability Calculation and Propagation

Consequence trees can be made quantitative by attaching probabilities or frequencies to branches. Typically, one assigns probabilities to the intermediate events (e.g. "ignition occurs"), then calculates the likelihood of each end outcome by combining these branch probabilities. If two events on a branch are independent prerequisites, their probabilities are multiplied (AND logic); if two alternative branches lead to the same outcome, their probabilities are added (OR logic, usually under the small-probability approximation)[12]. In practice, one often uses historical failure rates or barrier reliability estimates to assign branch probabilities. For example, one QRA procedure estimates leak frequencies from industry databases (UK HSE, TNO Purple Book, OREDA, etc.), *"then constructs event trees… using probabilities for each event branch to determine the frequency of the final hazard outcomes."*[10]. In this way, each branch's probability is the product of the probabilities of events along that path, and the total probability for a given final consequence is the sum of probabilities of all branches leading to it (assuming distinct scenarios).

These probability propagation rules mirror those of event-tree analysis and fault-tree analysis. For instance, as one guidance note puts it: *"Add the probabilities which sit below an OR gate… Multiply the probabilities which sit below an AND gate."*[12]. Thus, consequence trees apply the same mathematical logic as more familiar risk models. In software or calculations, each branch's likelihood is computed and mapped to its outcome, producing a distribution of consequences. In the offshore example above, after assigning leak frequencies and barrier failure probabilities, the method *"merges the consequences and likelihoods of all incident outcomes… to produce an overall measure of risk"*[10].

### 1.10 Strengths, Use Cases, and Limitations

Consequence Tree Analysis has several strengths. It provides a comprehensive, visual mapping of complex accident chains, helping engineers ensure that all plausible outcomes are considered[10]. It is especially valuable in process safety and major hazards industries (chemical plants, oil & gas, pipelines, etc.) where a single release can follow many paths. By explicitly listing outcomes, CTA supports scenario identification, consequence modelling (e.g. fire and explosion sizing), and risk quantification. It also links naturally with fault trees and Bow-Ties, giving a holistic cause–consequence [13]. For instance, regulators and industry guides (e.g. EIGA, COMAH/LPC) recommend consequence sequences be developed for each loss-of-containment scenario[10]. In practice, QRA software like SAFETI-OFFSHORE or PHAST often embeds CTA-style logic to compute population and individual risk based on leakage outcomes.

CTA is best applied in complex systems where multiple outcomes must be differentiated. It is commonly used in hydrocarbon and chemical plant risk assessments, offshore safety cases, pipeline QRA and domino-effect analysis. For example, an offshore oil platform QRA might develop consequence chains for well blowouts or pipeline ruptures, considering fire, explosion, toxic hazard and environmental releases[9]. Similarly, safety engineers use CTA in early design reviews to verify that protective features and emergency plans cover all identified hazard.

However, consequence trees also have limitations. Because they can grow very large (dozens of branches) CTA can become unwieldy for complex facilities. Building and updating a detailed consequence tree requires significant effort and expert judgement on scenario plausibility. Probabilities of some branches (e.g. rare ignition cases, hidden assumptions) may be highly uncertain. Moreover, CTA usually assumes branch events are independent, which may not hold if common-mode failures exist. Finally, consequence trees by themselves do not illustrate causes – they must be paired with fault-tree or bow-tie analysis if root causes need quantification. Despite these caveats, CTA remains a powerful tool when used in conjunction with other methods, offering clear documentation of possible accident sequences.

## 1.11 Industrial Examples

### 1.11.1 Hydrocarbon pipeline rupture

Consider a natural gas pipeline rupture. A consequence tree would branch on whether the gas cloud finds an ignition source or not. Possible outcomes include: immediate ignition at the break (jet fire), delayed ignition after dispersion (vapour cloud fire or explosion), or no ignition (gas disperses safely). Further branches capture effects like flame impingement, overpressure and thermal radiation zones. In a QRA, each branch's probability is found by multiplying the leak frequency by ignition probabilities (often drawn from industry data)[10]. Common outcomes in pipeline leak analyses include jet fires, pool fires (if liquid), vapor cloud explosions, projectiles from ejected pipe fragments, toxic clouds or environmental contamination[9]. For each failure mode (e.g. small leak vs. large rupture) the analyst *"develop[s] a series of consequence chains… illustrating the possible hazard outcomes."*[10]. The probabilities of each final consequence (fatality, explosion, etc.) are then estimated by following each branch and combining event probabilities (e.g. probability of ignition AND probability of critical cloud concentration leads to explosion).

### 1.11.2 Chemical storage or pressure vessel scenarios.

In a tank or vessel rupture, a CTA might capture sequences like: catastrophic rupture → immediate fuel pool + boiloff → ignition → pool fire; or rupture → loss of coolant → BLEVE (boiling liquid expanding vapor explosion); or containment breach → vapor cloud formation → delayed ignition → VCE. Each branch would reflect different mitigation (e.g. pressure relief working or not) and final outcomes. Similar CTA logic is used in quantitative risk assessment of major accident hazards (e.g. offshore flare release, LPG terminal accidents) to ensure all explosion and fire scenarios are considered.

### 1.11.3 Offshore/QRA practice.

As one risk management report notes for offshore oil and gas: first potential leak scenarios are identified in each plant section, then *"a series of consequence chains is developed for each failure scenario, illustrating the possible hazard outcomes."*[10]. The report goes on to explain that leak frequencies are taken from databases (e.g. OREDA, UK HSE) and *"event trees are constructed… using probabilities for each event branch to determine the frequency of the final hazard outcomes."*[10]. Thus, in real QRA work CTA (often implemented via event-tree algorithms) yields quantitative likelihoods of, say, pool fire causing human harm or a vapor cloud explosion. These result values are then used in risk metrics (individual and societal risk contours).

In summary, consequence trees are a key component of modern risk assessment for industrial hazards. They systematically capture what happens *after* an initiating incident, complementing fault trees (why it happened) and supporting comprehensive safety analysis. By explicitly laying out sequences of fires, explosions, toxic releases and other damages, CTA helps engineers classify outcomes, estimate their probabilities and make informed risk-reduction decisions[8].

# Theoretical Basis for Risk Simulation Using PHAST

**Introduction**

PHAST (Process Hazard Analysis Software Tool) by DNV is a leading consequence-analysis package for hydrocarbon releases, fires and explosions. It models the physics of each scenario from release through dispersion and ignition to thermal and blast effects. Key phenomena include jet fires, pool fires (early and late), flash fires, fireballs, unconfined vapor cloud explosions (UVCE), BLEVEs, and boilovers. Each is governed by fluid and combustion physics (mass flow, combustion energy, flame radiation, gas dynamics), which PHAST approximates with empirical and semi-empirical models. This chapter reviews the underlying theory and equations that PHAST uses for each scenario, as well as its treatment of dispersion under varying meteorological conditions and its overall modeling workflow.

## 2 Fire and Explosion Phenomena Considered in Risk Assessment

### 2.1 Jet Fires

A jet fire occurs when pressurized flammable gas or vapor escapes (typically from ≥2 bar piping or vessels) and ignites immediately, producing a high-momentum turbulent flame plume. The flow is often choked, leading to near-sonic exit velocity. PHAST models jet flames using correlations by Johnson et al. (1994) (cited in the PHAST manual) for flame length $L$ and emissive power. For example, the PHAST manual notes that the Johnson correlation predicts horizontal jet flame lengths within 10% of large-scale data. In practice, PHAST computes the jet's heat release rate is the radiative fraction (empirical, ~0.2–0.3 for hydrocarbons). The flame is approximated as a slender cylindrical or point-radiating source. The peak flame temperature (≈2000–2200 K) and emissivity (∼0.1–0.4) yield an emissive power, and radiative flux at distance $r$ is computed via view factors (often decaying roughly as $\frac{1}{r^2}$ or by a specific pool-radiation model). Wind can tilt and stretch the flame; higher wind speeds shorten and deflect the flame plume, reducing downstream thermal radiation. Jet fires have minimal growth time and rapidly reach full intensity. Typically, no significant overpressure is produced by an open-air jet fire (it is a diffusion flame), so PHAST only computes thermal radiation and flame impingement effects.

### 2.2 Pool fire

A pool fire forms when a liquid hydrocarbon collects on the ground or in a containment area and burns. PHAST distinguishes early (growing) and late (steady-state) pool fires. In an early pool fire, the liquid release rate into the pool exceeds the burning rate, so the pool grows until fuel accumulation equals evaporation. In a late pool fire, the pool has reached its maximum size (all released liquid is pooled) and burns at the maximum burning rate. PHAST computes the burning rate per unit area from empirical correlations (e.g. Bosch and Weterings, 2005) that depend on fuel type and environmental heat feedback. The flame emissive power (radiant intensity per unit area) also follows experimental correlations and typically ranges 100–400 kW/m² for large hydrocarbon pools. PHAST then treats the pool fire as a uniformly radiating disk at flame temperature; the radiative flux to receptors is computed via flame-wall view factors or point-source approximations. If wind speed is significant, PHAST reduces the flame height and emissivity (due to plume tilt) and may apply a view-factor model accounting for elevation.

## 2.3 Late Pool Fires

A late pool fire in PHAST is simply the steady-state burning of the pooled liquid after an initial release phase. For example, in a flammable liquid spill, an initial vapor cloud may burn off (flash fire), after which the remaining liquid burns as a pool fire. PHAST models this by assuming the pool has reached a maximum diameter (equal to the spill footprint) and that it burns at the empirically defined maximum burn rate. The flame length and heat flux are then calculated as for any steady pool fire of that size. In practice, PHAST reports both early and late pool fire results: the early (growth-limited) fire with smaller heat release, and the late (size-limited) fire which typically has higher heat flux at close range because of the larger pool.

## 2.4 Flash Fires and Fireballs

A flash fire refers to the diffusive combustion of a flammable vapor cloud without appreciable overpressure. It occurs when a dispersed gas/vapor cloud ignites and burns as a flame front travel through the cloud. In PHAST, a flash fire is treated as an instantaneous ignition of a uniformly mixed cloud: the fuel mass $m$ in the cloud is assumed to burn, yielding a heat release over the flame duration. The radiative output is computed similarly to a large flame: PHAST often assumes an “equivalent” pool/jet flame or fireball of that energy. Because there is no confinement, the dominant effect is thermal radiation and convective heat flux (the latter typically of secondary importance). Overpressure is negligible. Fireball scenarios in PHAST are similar to flash fires but assume a roughly spherical flame (as from a large instantaneous release). For a fireball, empirical scaling is used: for example, the fireball radius R may be estimated by empirical scaling laws (with $k$~0.2–0.3 for hydrocarbon fires) and the flame duration on the order of a second. The total energy radiated by the fireball is again times a radiative fraction. PHAST then computes peak radiative flux vs. distance using a point-source or solid-sphere model. (In practice, PHAST′s output for "fireball" is effectively the instantaneous heat pulse and any small blast pressure from the rapid expansion.)

### 2.5 Unconfined Vapor Cloud Explosion (UVCE)

An **unconfined vapor cloud explosion** (UVCE) occurs when a flammable cloud ignites and the combustion is fast enough to drive a pressure wave. PHAST models UVCEs using empirical blast correlations (the Baker–Strehlow–Tang (BST) method or similar) that relate the cloud's energy and reactivity to peak overpressure and impulse at distance. In this approach, only a fraction of the cloud's combustible energy contributes to the blast (since the flame speed is subsonic). PHAST typically estimates a "flame speed" based on fuel type, concentration, and obstruction, then uses BST-based blast curves to compute the side-on overpressure P vs. distance. In essence, PHAST assumes an equivalent spherical explosion with energy and applies standard blast wave scaling (e.g. Hopkinson's cube-root law). The resulting pressure P at distance $x$ often follows empirical fits (e.g. detonation analogies or the DiNenno/Baker curves). The positive-phase duration is also obtained from the model, but PHAST emphasizes peak overpressure and impulse. Unlike pool/jet fires, dispersion modeling of the cloud is key: the cloud geometry and concentration distribution (heavier gases stay near ground) are computed first, then ignition is assumed to consume the flammable portion. A pure flash (without blast) is produced if the cloud is too lean or small; otherwise, an explosion with finite $P$ occurs. PHAST does not model turbulence-driven deflagration details; it relies on validated empirical blast models.

### 2.6 Boiling Liquid Expanding Vapor Explosion (BLEVE)

A **BLEVE** involves sudden failure of a vessel containing a pressurized liquefied hydrocarbon (e.g. LPG tank). The liquid rapidly flashes to vapor and expands. PHAST models a BLEVE as a combination of a vapor explosion and a fireball. First, the vessel's inventory mass is assumed to instantly vaporize adiabatically; the resulting gas, if ignited, is treated like a flash fire or small explosion of energy. In practice, most of the BLEVE hazard is the subsequent *fireball* of burning vapor. PHAST assumes the entire mass forms a fireball with a radiant fraction. The fireball's size and duration are then estimated (similar to a gas fireball as above). A minor blast overpressure may be calculated using the vapor explosion model (as for UVCE) but is usually modest due to venting. The key output is thermal radiation from the fireball. PHAST does not explicitly simulate the vessel's collapse dynamics; it simply treats a BLEVE as an instantaneous inventory release.

## 2.7 Meteorological Effects on Dispersion

Weather conditions critically affect how a released gas cloud disperses before ignition or dilution. In PHAST, the most important meteorological inputs are wind speed and direction, ambient temperature, relative humidity, and solar radiation. Wind speed $u$ drives the plume advection and dilution rate: higher wind yields faster transport but also greater mechanical turbulence, leading to quicker dilution and shorter travel distances. Low wind means slower dispersion and higher concentrations downwind. Ambient temperature (and the resulting lapse rate) influences buoyancy: hot ground (high insolation) produces strong vertical mixing (unstable conditions), while cool or night conditions suppress mixing. Relative humidity affects dense-phase clouds: in cold vapor releases (e.g. LPG, LNG), ambient moisture can condense and release latent heat, warming the cloud and altering its buoyancy and dispersion. PHAST's integral dispersion model even accounts for humidity condensation: including moisture can increase predicted plume rise and reduce ground-level concentrations, as latent heat is released. Solar radiation enters indirectly by setting the stability class (see below) and by heating surfaces, thereby influencing buoyancy and vertical mixing.

## 2.8 Pasquill–Gifford Stability Classes

Atmospheric turbulence is classified by the Pasquill–Gifford (PG) scheme (classes A through F) based on solar insolation and wind speed. Class A is *very unstable* (strong sun, light wind), B/C are unstable, D is neutral (overcast or moderate wind), and E/F are stable (nighttime or high cloud cover with low wind). In unstable classes (A–C), strong thermal turbulence causes rapid vertical and horizontal dispersion – plumes quickly dilute and remain low in concentration. In stable classes (E–F), weak turbulence causes plumes to remain concentrated and travel farther. PHAST uses the chosen PG class to select Gaussian plume dispersion coefficients ($\sigma_y$, $\sigma_z$) for far-field dispersion. For example, in class A, $\sigma_z$ (vertical spread) is large, producing a short, wide cloud, whereas in class F, $\sigma_z$ is small and the cloud remains narrow and dense. The NOAA Ready site summarizes the meteorological criteria (insolation, wind) defining each PG class. In practical modeling, the user inputs either a PG class or parameters (time of day, cloud cover, wind) from which PHAST infers the class.

*Table 2.1Classification of Atmospheric Stability and Associated Dispersion Behaviors*

| Stability Class | Atmospheric Condition | Surface Wind Speed (m/s) | Solar Radiation (Daytime) | Sky Condition (Nighttime) | Turbulence | Dispersion Characteristics |
|---|---|---|---|---|---|---|
| **A** | Very Unstable | < 2 | Strong (sunny, clear) | – | Very High | Very rapid vertical and lateral spread |
| **B** | Unstable | 2–3 | Moderate (partly sunny) | – | High | Rapid dispersion |
| **C** | Slightly Unstable | 3–5 | Weak (cloudy or low sun) | – | Moderate | Moderate dispersion |
| **D** | Neutral | > 5 (day) or overcast (night) | Overcast or anytime with strong wind | Overcast | Low–Moderate | Standard (Gaussian) dispersion |
| **E** | Slightly Stable | < 3 | – | Clear with light wind | Low | Slow dispersion |
| **F** | Stable | < 2 | – | Clear and calm (little/no wind) | Very Low | Very slow; long and narrow plume |

**lass A–C:** Occur during daytime with solar heating → strong vertical mixing.

**Class D:** Neutral conditions, often during overcast skies or with moderate wind.

**Class E–F:** Occur at night, especially under clear skies with little wind → plume remains concentrated and close to ground.

*Table 2.2 Empirical Formulas for Gaussian Plume Dispersion Parameters*

| **Stability Class** | **σᵧ – Rural (m)** | **σ_z – Rural (m)** | **σᵧ – Urban (m)** | **σ_z – Urban (m)** |
|---|---|---|---|---|
| **A** (Very unstable) | 0.22x * (1 +0.0001x) ^−0.5 | 0.20x | 0.32x*(1+0.0004x) ^−0.5 | 0.24x |
| **B** (Unstable) | 0.16x * (1 + 0.0001x) ^−0.5 | 0.12x | 0.22x(1+0.0004x) ^−0.5 | 0.20x |
| **C** (Slightly unstable) | 0.11x * (1 + 0.0001x) ^−0.5 | 0.08x * (1 + 0.0002x) ^−0.5 | 0.16x*(1 + 0.0004x) ^−0.5 | 0.16x * (1 + 0.0002x) ^−0.5 |
| **D** (Neutral) | 0.08x(1 + 0.0001x) ^−0.5 | 0.06x *(1 + 0.0015x) ^−0.5 | 0.11x * (1 + 0.0004x) ^−0.5 | 0.11x (1 + 0.0004x) ^−0.5 |
| **E** (Slightly stable) | 0.06x *(1 + 0.0001x) ^−0.5 | 0.03x * (1 + 0.0003x) ^−1 | 0.08x *(1 + 0.0004x) ^−0.5 | 0.08x * (1 + 0.0003x) ^−1 |
| **F** (Stable) | 0.04x* (1 + 0.0001x) ^−0.5 | 0.016x * (1 + 0.0003x) ^−1 | 0.06x *(1 + 0.0004x) ^−0.5 | 0.06x * (1 + 0.0003x) ^−1 |

## 2.9 Dispersion Models in PHAST

PHAST uses several dispersion approaches depending on the scenario and release properties:

- **Gaussian (Pasquill) plume model:** For neutral or buoyant gas releases in unobstructed terrain, PHAST applies classic Gaussian dispersion. The cloud concentration downwind is given by:

$$C(x,y,z) = \frac{\dot{m}}{2\pi u \sigma_y \sigma_z} \exp\left(-\frac{y^2}{2\sigma_y^2}\right) [\exp\left(-\frac{(z-H)^2}{2\sigma_z^2}\right) + \exp\left(-\frac{(z+H)^2}{2\sigma_z^2}\right)]$$

- where "H" is releasing height. The lateral (σ_y) and vertical (σ_z) spreads are functions of downwind distance and stability class. PHAST's far-field (passive) clouds have been verified to match standard Gaussian formulas. This model is typically used for light gas (e.g. methane) or small releases where buoyancy is negligible or positive.

- **Dense/heavy gas integral models:** For heavy gas releases (LFL vapor denser than air, e.g. LPG, anhydrous ammonia, or cold cryogens), gravity and momentum dominate. PHAST 8.x uses a Unified Dispersion Model (UDM) an integral puff model that transitions through three stages – momentum-dominated jet, gravity-slumping heavy gas cloud, then passive Gaussian diffusion. The UDM solves conservation equations (mass, momentum, energy) via differential equations for the cloud's height, width, mass, and momentum as it travels. For validation, PHAST's heavy gas mode has been compared to established models (e.g. Shell's HEGADAS, HGSYSTEM) and shows good agreement. For example, UDM reproduces HEGADAS results for centerline concentration in a steady gas cloud. Essentially, PHAST's heavy-gas model accounts for slumping, reduced dispersion, and possible droplet formation. For two-phase (liquid) releases, it tracks droplet rainout and re-evaporation – PHAST links a pool evaporation model to the dispersion so that vapor feeding the cloud comes partly from the spilled liquid.
- **Building wake model (not core PHAST):** In congested areas, wake effects can trap or disperse a cloud; PHAST has an optional "building wake" dispersion adjustment. (The standard open-air models assume unobstructed flow.)
- **Instantaneous vs. continuous models:** PHAST handles both "continuous" (long-duration) and instantaneous (explosive) releases. Continuous mode seeks steady-state (time-independent) dispersion, while instantaneous mode tracks a puff evolving in time. PHAST's UDM uses a top-hat to Gaussian similarity profile to smoothly transition from near-field to far-field dispersion.

In summary, Gaussian plume models are applied to far-field, neutrally buoyant clouds (large distance from source), whereas heavy/dense gas models apply when the cloud is cooler/heavier than air and remains near ground. PHAST's unified model blends these seamlessly: the early cloud may behave like a gravity current, but the late far-field behavior converges to a Gaussian plume.

### 2.10 Consequence Thresholds for Thermal Radiation and Overpressure

The following tables summarize widely used hazard thresholds in oil/gas consequence analysis (PHAST, ALOHA), based on API guidance, CCPS, UK HSE, NOAA/ALOHA, etc. For thermal radiation, exposures are given in $kW/m^2$. For overpressure, both psi and bar are shown. (For context, **API 521** design limits continuous exposure to **1.58 kW/m²** (clothed personnel), with **4.73 kW/m²** allowable for short-term emergency operations. Exceedance of these values leads into the injury regimes below.)

### 2.10.1 Thermal Radiation (kW/m²)

*Table 2.3Radiant Heat Flux Thresholds and Associated Human, Material, and Equipment Effects*

| **Effect/Threshold** | **Radiant Heat Flux (kW/m²)** | **Typical Consequences** | **Source** |
|---|---|---|---|
| *Reversible injury (first-degree burn/pain)* | ~1.6–2.0 (bare skin) | Pain/discomfort within minutes; first-degree burns. No lasting injury. | NOAA/ALOHA: pain at 2 kW/m²; Sandia: no harm up to 1.6 kW/m² |
| *Irreversible injury (second-degree burn)* | ~5 kW/m² (20–60 s exposure) | Second-degree burns (blistering); tissue damage requiring medical care. | NOAA/ALOHA: 2nd° burns at 5 kW/m²; Sandia data (9.5 kW/m² → 2nd° in 20s) |
| *Lethality threshold* | ~10–12 kW/m² (60 s) | Lethal burns/thermal shock. ~1% fatal at 12.5 kW/m² in 1 min (ISO); 100% fatal near 25 kW/m² (1 min). | NOAA/ALOHA: "lethal" at 10 kW/m² in 60s; UK HSE: fatalities by 12.5 kW/m² (minutes), immediate death ~37.5 kW/m²; Sandia: 12.5–15 kW/m² gives ~1% lethality in 1 min, 25 kW/m² ~100% lethal in 1 min |
| *Equipment damage / domino* | ~35–37.5 kW/m² (long exposure) | Damage to process equipment/structures, failure of protective cladding, ignition of structural materials. Triggers escalation (domino). | Sandia (HIA Task 19): 35–37.5 kW/m² causes equipment/structural damage (>30 min); industry rule-of-thumb "equipment damage" at ~37.5 kW/m² |
| *Material ignition (for reference)* | 12.5–25 kW/m² | Piloted ignition of wood ~12.5 kW/m²; unpiloted ignition of wood/plastics ~25–32 kW/m². Clothing ignites ≈75 kW/m². | Sandia: 12.5 kW/m² piloted wood ignition, 25–32 kW/m² unpiloted ignition |

### 2.10.2 Overpressure (Blast) Thresholds

*Table 2.4Overpressure Thresholds and Associated Human, Structural, and Equipment Damage Effects*

| Effect/Threshold | Overpressure | Typical Consequences | Source |
|---|---|---|---|
| *Minor injury (glass/debris)* | ~0.02–0.07 bar (0.3–1.0 psi) | Broken windows, flying glass/debris causing cuts. | NOAA: glass failure starts ~0.01–0.05 bar (0.15–0.7 psi); lacerations from debris 0.07–0.55 bar |
| *Eardrum rupture* | ~0.2–0.4 bar (3–6 psi) | Perforation of eardrums and bleeding (risk of permanent hearing loss). | CDC/DoD: 0.35 bar (5 psi) causes ~1% eardrum rupture; NOAA/Lees: 0.17–0.84 bar (2.4–12.2 psi) → 1–90% eardrum rupture |
| *Pulmonary injury (lung damage)* | ~1.0 bar (≈15 psi) | Pulmonary hemorrhage and lung trauma (acute respiratory damage). | CDC/DoD: threshold for lung injury ~1.0 bar (15 psi); VA/DOD news: "50 kPa (7.3 psi) can rupture eardrum and cause lung injury" (hemorrhage by ~14.5 psi) |
| *Serious injury / fatalities* | ~3.0–4.0+ bar (≈45–60 psi) | Severe/blast injuries; chance of death. ~1% fatal at ~2.4 bar (35 psi), ~~100% fatal by ~3.8–4.5 bar (55–65 psi). | Lees (via NOAA): 14.5–29.0 psi (1.0–2.0 bar) → 1–99% fatalities; CDC: 2.4–3.1 bar → ~1% killed; 3.8–4.5 bar → ~99% killed. |
| *Building damage* | ~0.07–0.7 bar (1–10 psi) | Structural damage to buildings: ~0.07 bar (1 psi) can demolish unreinforced houses; ~0.7 bar (10 psi) likely total destruction. | NOAA/Lees: 0.07 bar – house collapse; 0.7 bar (10 psi) – "total building destruction". |
| *Equipment damage / domino* | ~0.07–0.2 bar (1–3 psi) | Overpressure above ~1–3 psi can deform or rupture process equipment (tanks, vessels), triggering escalation to adjacent | Process-safety studies: atmospheric tanks fail at ~0.07 bar (1.02 psi); pressurized equipment at ~0.2 bar (2.90 psi). (Such thresholds are often used |

| | | | |
|---|---|---|---|
| | | units. | as domino/damage criteria.) |

## 2.11 Conditional Probability in Oil & Gas Process Safety

**Conditional probability** is the likelihood of an event occurring given that another related event has occurred. In probability theory, it is formally defined as the probability of **A** occurring *given* that **B** has occurred, denoted P(A|B)[14]. In risk assessment, conditional probability is fundamental: for example, we often compute the probability of a hazardous outcome **given** an initiating failure or scenario. In practical terms, an accident "sequence" probability is calculated as the initiating event probability times the conditional probabilities of subsequent failures. For example, in a simplified event-tree analysis the probability of an accident sequence (AS) with events A (initiator), B (system failure), and C (containment breach) is:

$P(AS) = P(A) \times P(B|A) \times P(C|A \wedge B)$[15].

Here the **P(B|A)** term – the probability of system failure given the initiating event – is provided by a fault-tree or other analysis. Thus conditional probabilities tie together top and intermediate events in risk models[14], [15].

- Example*:* If an equipment leak occurs (event B), the conditional probability of ignition (event C) given that leak – P(C|B) – quantifies the chance of fire or explosion. Such conditional factors (e.g. ignition probability, occupancy) are often used in process safety calculations.

*Table 2.5 Probabilities of Outcome Types Following Different Leak Scenarios and Release Phases Based on IOGP Guidelines*

| Leak Size / Phase | Immediate Ignition | Delayed Ignition | No Ignition | Explosion (VCE) | Flash Fire | Fireball | Pool Fire |
|---|---|---|---|---|---|---|---|
| **Small Gas Leak** | 0.1 – 0.3 | 0.1 – 0.2 | 0.6 – 0.8 | 0.05 – 0.15 | 0.15 – 0.3 | ~0 | Not applicable |
| **Medium Gas Leak** | 0.3 – 0.5 | 0.2 – 0.4 | 0.3 – 0.4 | 0.2 – 0.4 | 0.3 – 0.5 | ~0 | Not applicable |
| **Large Gas** | 0.5 – 0.7 | 0.2 – | 0.1 – | 0.4 – 0.6 | 0.3 – | ~0 | Not |

| | | | | | | | |
|---|---|---|---|---|---|---|---|
| **Leak** | | 0.4 | 0.2 | | 0.5 | | applicable |
| **Liquid Leak** | 0.2 – 0.4 | 0.1 – 0.3 | 0.3 – 0.6 | ~0.05 | ~0 | ~0.2 | 0.6 – 0.8 |
| **Two-Phase Release** | 0.4 – 0.7 | 0.2 – 0.3 | 0.1 – 0.3 | 0.3 – 0.6 | 0.1 – 0.3 | 0.6 – 0.9 | 0.2 – 0.4 |
| **Catastrophic Rupture** | 0.6 – 0.9 | 0.1 – 0.3 | <0.1 | 0.3 – 0.6 | ~0.2 | 0.7 – 0.9 | 0.2 – 0.4 |

## 3 Quantitative Consequence Modeling Studies

Many recent studies have employed tools like PHAST to simulate the consequences of failures involving separators and storage vessels in oil and gas installations. One such case involved a corroded two-phase oil–gas separator on an offshore platform. The simulation results indicated hazardous conditions, including thermal radiation levels of 37.5 kW/m² extending to around 21 m, overpressure of 0.2 bar reaching 30 m, and 0.02 bar extending up to 200 m. Toxic concentrations near 20,000 ppm were reached within a few seconds of the release. Based on these findings, recommendations included rerating or replacing the vessel and relocating critical infrastructure such as access routes and control rooms upwind[16] .

In another study, artificial intelligence models were used to predict the efficiency of a wellhead separator in a gas condensate field, followed by PHAST modeling of a hypothetical explosion. The simulation revealed severe consequences, reinforcing the need for frequent inspection and maintenance. The integration of AI with consequence modeling proved useful for anticipating equipment performance and guiding preventive strategies [17].

Another analysis focused on a high-pressure methane storage tank identified as a worst-case scenario during preliminary hazard analysis. Fault tree analysis linked the explosion risk to mechanical failures, process deviations, and human errors. PHAST simulations predicted a vapor cloud explosion, with probit analysis estimating up to 16 fatalities. To mitigate this risk, recommendations included eliminating mechanical defects, reinforcing structures, and tightening operational procedures[18] .

In a separate case, a risk-based inspection approach based on API 581 was applied to a condensate separator and an associated storage vessel. The calculated probability of failure was in the range of 10^-5 per year. Based on this, inspection intervals were optimized, with the next assessment planned several years ahead. This approach highlighted the value of RBI in identifying low-risk equipment and informing maintenance planning[19].

### 3.1 Dispersion and Fire Modeling (ALOHA/PHAST)

A number of studies have focused on gas dispersion and fire consequences. One study simulated a hydrocarbon venting scenario on an offshore production platform using both ALOHA and PHAST. Various meteorological conditions and vent parameters were considered, but the results consistently showed that the flammable plume rose above personnel areas. In all cases, the vapor cloud did not reach manned modules or escape routes. The conclusion was that no additional risk mitigation was needed beyond standard design, since even the worst-case dispersion posed no hazard to personnel[20] .

In another case, a catastrophic LPG cylinder leak was modeled using ALOHA. The simulation showed very high gas concentrations—around 33,000 ppm (IDLH level)—within 11 m of the release source, and the lower explosive limit (~21,000 ppm) was reached between 12–16 m. The resulting fire/explosion hazard zone extended to approximately 15 m. Based on these results, evacuation distances of about 15 m were recommended to improve emergency preparedness and reduce exposure[21] .

Another study investigated potential fire and explosion consequences from a large LPG storage sphere using PHAST. Scenarios included pool fires and BLEVEs, with estimated thermal and blast impact zones. For example, a large pool fire could affect areas up to 100 m away. The analysis led to recommendations regarding stand-off distances and personnel restrictions around the tank, emphasizing the need for wide exclusion zones around large LPG installations [22].

### 3.2 Qualitative and Risk-Based Analyses

In parallel with modeling, qualitative methods and structured analyses often guide risk assessment. For separators and pressure vessels, HAZOP studies might be used internally by industry, but explicit HAZOP-case publications are scarce. However, Fault Tree Analysis

(FTA) and Event Tree Analysis (ETA) have been employed in published case studies. In one example, FTA was used to identify root causes—mechanical, process, and human—of a hypothetical methane tank failure. This analysis followed an initial Preliminary Hazard Analysis (PHA) to select the worst-case scenario, demonstrating how PHA, FTA, and consequence modeling can be integrated effectively [18].

Similarly, ETA is commonly used in Quantitative Risk Assessment (QRA) to map out potential outcome pathways, although specific ETA applications on condensate separators remain limited in the open literature. Risk-Based Inspection (RBI) provides a semi-quantitative framework that combines the probability of failure (PoF) and the consequence of failure (CoF). One case study applied API 581-based RBI to a condensate separator and a related storage vessel, resulting in prioritized inspection intervals and enhanced understanding of future failure likelihood[19] .

Beyond RBI, other widely adopted methods—such as Layer of Protection Analysis (LOPA) and Safety Instrumented System (SIS) studies—are often referenced in industry practice. However, detailed LOPA or SIS studies specifically focused on condensate separators are rarely published. Overall, these qualitative approaches complement quantitative modeling by helping prioritize safety measures—for example, by recommending corrosion monitoring to address wall thinning[16] or emphasizing procedural documentation as revealed by root cause analysis findings.

In summary, the reviewed studies consistently emphasize that separators and pressure vessels, if left unmonitored, can result in severe fire or explosion consequences. Tools like PHAST and ALOHA quantify hazard zones, enabling emergency planning and design adjustments. Meanwhile, qualitative methods such as RBI and FTA identify failure causes and help define inspection or mitigation priorities. Together, these complementary approaches support a more robust and proactive risk management strategy.

*Table 3.1Notable Risk Assessment Studies Involving Process Equipment Failures and Consequence Modeling*

| Study | Tools/Methods | Key Scenario(s) | Major Outcomes and Recommendations | Year |
|---|---|---|---|---|
| Naemnezhad *et al.* | PHAST v7.x | Corroded oil–gas separator explosion (225 psi) | 37.5 kW/m² radiation to ~21 m; overpressure 0.2 bar to ~30 m, 0.02 bar to ~200 m; ~20,000 ppm toxic at 3 s. Advised rerating or replacing separator and relocating buildings/routes upwind. | 2017 |
| Jahani *et al.* | PHAST v7.x | Gas condensate leakage: jet fire, pool fire, flash fire, vapor cloud explosion from 650 tanks | Vapor cloud explosion and pool fire were worst; hazards extended ~490 m to control room and routes. Recommended moving access/muster points upwind and hardening control center. | 2019 |
| Davarikhah *et al.* | ANN/ANFIS + PHAST v7.2 | Wellhead separator probable explosion | Trained ANN/ANFIS to predict separation efficiency, then PHAST showed severe explosion hazard. Emphasized frequent inspections; ANFIS outperformed ANN. | 2020 |
| Khodadadi Mousiri *et al.* | PHAST 7.2 + RCA (brainstorming) | Actual methane pressure vessel explosion in gas separation unit | At 15–45 m: radiation 12.5–4 kW/m²; at 20 m: overpressure 0.206 bar (0.020 bar at 150 m). Vapor cloud explosion was key. Root-cause analysis identified equipment/procedural failures. Advocated proactive risk management via consequence modeling. | 2023 |
| Suyut *et al.* | ALOHA v3.x & PHAST | Flammable gas venting on offshore platform | Simulated vent dispersion under various weather. Plume reached high altitude without affecting personnel or escape routes. Concluded no extra mitigation needed beyond standard design. | 2024 |
| Shahedi Ali Abadi *et al.* | PHAST + FTA + PHA | 120 bar methane storage tank explosion | PHA identified explosion of methane tank as worst-case; FTA found mechanical, process, human failures. PHAST | 2019 |

| | | | | |
|---|---|---|---|---|
| | | | predicted a vapor cloud explosion; probit analysis estimated up to 16 fatalities. Recommended eliminating mechanical faults, blast-proofing, and stricter procedures to reduce risk. | |
| Siswantoro *et al.* | RBI (API 581) + FTA | Condensate separator (10V2102) and storage vessel (10V2103) failure | Calculated current/future PoF: Separator ~1.4×10^-5 → 4.6×10^-5; Storage ~1.3×10^-5 → 1.6×10^-5. Assigned qualitative CoF levels; prioritized inspection dates (e.g. separator next in 2026). Demonstrated RBI's value in identifying critical equipment. | 2021 |
| Beheshti *et al.* | ALOHA v5.x | Full LPG cylinder rupture (26–107 L) | Catastrophic release model: 33,000 ppm (IDLH) reached ~11 m; LEL (~21,000 ppm) reached ~12–16 m; flame hazard ~15 m. Advised setting evacuation zones ~15 m from cylinder during accidents. | 2018 |
| Malviya & Rushaid | PHAST 6.5 | Rupture/BLEVE of large LPG storage sphere | PHAST modeled thermal/blast effects of pool fire/BLEVE. For example, a pool fire hazard radius ~100 m. Recommended safety stand-off distances (order of 100 m) and blast-resistant design for nearby structures. | 2018 |

# Failure of a Pressurized Vessel

# Case study

## 3.3 SCENARIO No. 1: AGTU – Failure of a Pressurized Vessel

### 3.3.1 process description:

The Associated Gas Treatment Unit (AGTU) at Tin Fouyé Tabankort (TFT) is responsible for processing associated gases through compression, dehydration, and condensate recovery. The recovered condensates are reinjected into crude oil, while a portion of the dry gas is used as lift gas for oil production. The remaining dry gas is exported via the GR1 pipeline to Hassi R'mel for commercialization. Originally, the AGTU was designed to recover gases that were flared at the separation centers CS1 to CS5 and Amassak. The process chain includes four main stages: compression, dehydration, condensate recovery, and recompression. Associated gases are received via two separators: D-201 for low-pressure gas and D-101 for high-pressure gas.

The AGTU is composed of five main sections. Section 100 handles high-pressure gas compression using turbines (KT 101 A/B) and compressors (K101 to K103 A/B), with multiple scrubbers (D101 to D105) and air coolers (E101 to E103) to separate liquids and cool the gas from around 170°C down to 55°C. Section 200 deals with low-pressure compression, utilizing compressors (K201 A/B), air coolers (E201 A/B), and scrubbers (D201 to D203) to raise the pressure from 0.6 to 6 kg/cm². Section 300 is dedicated to gas dehydration and includes adsorption towers (D301 A/B/C), heaters (H301 A/B), and filters (V301 A/B) to remove moisture and solid particles from the gas. Section 400 ensures condensate recovery and stabilization using heat exchangers (E402 and E403 A/B), separators (D401 and D402), a stripper (C401), a heater (H401), and circulation pumps (G401 A/B). Finally, Section 500 contains scrubbers (D501 and D502) for fuel gas purification and a flare drum (D511) for collecting all purged gases and liquids.

The gas coming from the high-pressure (HP) and low-pressure (BP) separation centers is processed as follows:

The BP gas enters the horizontal separator D-201, where liquids are removed and directed to the flare separator D-511 for evacuation to the CS2 oil center. The dry gas is then split into two streams, each passing through vertical scrubbers D-202 A/B to eliminate water content. Afterward, the gas is compressed in compressors K201 A/B from 0.6 kg/cm² to 6 kg/cm². The resulting hot gas, at approximately 150°C, is cooled using air coolers E-201 A/B and then passed through scrubbers D-203 A/B. After this stage, the gas merges with the

HP gas stream.The HP gas enters the horizontal separator D-101 for liquid removal, and the separated liquids are sent to the flare separator D-511, just like the BP liquids. The gas is also split into two streams, entering vertical scrubbers D-102 A/B to remove remaining liquids. It is then compressed in compressors K101 A/B from 6 to 28 kg/cm² in the first compression stage. The gas, now at 150°C, is cooled in air coolers E-101 A/B before continuing to scrubbers D-103 A/B. The second stage of compression occurs in compressors K102 A/B, where pressure is raised from 28 to 80 kg/cm². Again, the gas is cooled using air coolers E-102 A/B and then passed through scrubbers D-104 A/B to eliminate any remaining liquid particles.

The dry, high-pressure gas is then sent to the dehydration section, where it enters molecular sieve towers D-301 A/B/C, operating in three phases: adsorption (production), heating (regeneration), and cooling. During the adsorption phase, water molecules are captured, and the dry gas exits through filters V-301 A/B. The filtered gas is then cooled from 55°C to 20°C using two exchangers: E-402 (gas-gas) and E-403 A/B (gas-liquid).

After cooling, the gas enters separator D-401, where most of the condensate is recovered. This condensate is used as a cooling fluid in the gas-liquid exchanger and later stabilized and recovered through the stripper C-401. After D-401, the gas undergoes a pressure drop from 80 to 20 kg/cm² via PCV-401A/B valves. Remaining liquid is separated in the stripper. The stripper bottom is heated to 150°C using the reboiler H-401 to stabilize the condensate by removing light hydrocarbons. This circulation is maintained using pumps G-401 A/B. Once stabilized and cooled in air cooler E-401, the condensate is sent to the CS2 center.

At the stripper outlet, the gas has a temperature of -3°C and is used for cooling in the gas-gas exchanger. Temperature regulation at the separator D-401 is managed by valve TCV-401, which controls the gas flow used for heat exchange. The gas exiting the stripper and exchanger is then split into two streams, each passing through scrubbers D-105 A/B before being compressed from 20 to 80 kg/cm². After final cooling in air coolers E-103 A/B, the gas reaches 55°C and is ready for dispatch. The pressure at this stage is controlled by valve PCV-110 at 78 kg/cm². Part of the gas is used for gas lift in oil wells, while the remaining portion is sent to Hassi R'mel and to supply the power plant with fuel gas.Finally, the molecular sieve towers are regenerated using gas taken from the third compression stage at 35 kg/cm², regulated by PCV-108. This gas passes through a cooling tower and is then heated in furnaces H-301 A/B for tower regeneration. After heating, the regeneration gas is directed to the first-stage discharge.

### 3.3.2 Equipment description

Separator D-401 sits at the heart of the condensate-recovery section (Section 400) of the AGTU. After high-pressure dehydration, the dry gas (80 bar, 20 °C) enters D-401, where gravitational and centrifugal effects strip out the bulk of condensate from the vapor stream. The recovered liquid is then fed to the heat-integrated gas-liquid exchanger (E-403) and onward to stripper C-401 for stabilization, while the vapor undergoes a controlled pressure let-down (80 → 20 bar via PCV-401) and a final liquid separation stage. A dedicated temperature-control valve (TCV-401) meters cool gas back through E-403 to maintain D-401's internal temperature around 20 °C, ensuring efficient phase separation. Pumps G-401 circulate stabilized condensate, and the treated dry gas is recompressed and dispatched for lift-gas injection or export.

*Table 3.2Key Parameters for Gas-Condensate Separator Operating Conditions*

| Parameter | Value |
|---|---|
| **Substance** | Gas-condensate mixture |
| **Operating Pressure** | 80 bar |
| **Operating Temperature** | 20 °C |
| **Vessel Volume** | 15 $m^3$ |
| **Molar Density (Gas phase)** | 16.62 $mol/m^3$ |
| **Molar Density (Liquid phase)** | 11.08 $mol/m^3$ |

### 3.3.3 Description of potential undesired events:

The HAZOP study was conducted on the entire downstream train, beginning at the outlets of the molecular-sieve filters (V-301 A/B) and continuing through the interstage exchangers (E-403 A/B and E-402), the high-pressure separator (D-401), the stripper (D-402), the pumps (G-401 A/B), and the final heat-up stages (air heater E-401 and furnace H-401), up to the 4″ CSO valve on the stripper. The level-control valve LCV-403A, the temperature-control valve TCV-401, and its bypass were also included to ensure comprehensive coverage.

To evaluate a range of potential release scenarios, simulations were performed for a series of leak sizes—6 mm, 25 mm, and 100 mm orifices—representing increasing severities of vapor release. These scenarios allowed the behaviors of jet fires, flash fires, dispersion patterns, and blast overpressure to be characterized under varying atmospheric and process conditions.

Based on these initial scenarios, the worst-case event—a full rupture of D-401—was assessed. This rupture was assumed to result from downstream valve failure, external mechanical impact, or overheating due to loss of cooling. Such a failure would lead to the sudden release of hundreds of kilograms of gas-condensate mixture at 80 bar, generating a large flammable cloud, significant overpressure, and thermal-radiation hazards in the form of jet and flash fires.

The failure modes considered for pressure equipment such as D-401 included leakage or rupture through vessel walls, welds, or fittings, as well as hot rupture under fire exposure. The primary causes were identified as mechanical overload, fatigue, corrosion, failure of accessories, and material or construction defects—often exacerbated by chemical interactions with the process fluid. According to Smith, 93% of pressure-equipment failures originate from crack formation, with 71% of those beginning at welds or thermally stressed areas, particularly at attachment points or geometric discontinuities. It was also found that over 40% of such cracks pre-existed before service, while 32% resulted from fatigue.

Accordingly, the HAZOP entry for the D-401 rupture node addressed not only initiating causes such as valve malfunction, external impact, or thermal stress, but also the broader failure context—especially weld integrity and fatigue. The complete chain of consequences was considered, including high-velocity two-phase discharge, formation of a flammable vapor cloud, and potential ignition resulting in fire or explosion

### 3.3.4 Controls, regulation, and safety measures in place:

The D-401 separator is located within the AGTU facility. It is designed for the separation of gas and condensates under high pressure. This separator is equipped with a set of control systems and instrumentation that allow precise process monitoring. Level control is ensured by the LIC401 controller, acting on the level control valve to maintain stable phase separation, while the level measurement is performed by the LT401 transmitter. High and low-level alarms (LSHH401 and LSL401) are also installed and reported to the control room, allowing interlock activation in case of critical level deviations. The internal pressure of the separator is monitored by the pressure transmitter PT401 and a local pressure gauge PG401. The separator is protected against overpressure by the PSV401 relief valve, set at 91.5 barg, with discharge routed to the flare. Fluid temperature is monitored by temperature transmitters TT401 and TI402, along with high and low-temperature alarms (TSHH401 and TSHL401). The separator feed is regulated by the FIC401 control loop via the corresponding control valve. The separator is connected to the closed drain system through line 2''-C055-B2, equipped with a manual valve. Local supervision is supported by level (LG401) and pressure (PG401) indicators. All critical measurements are reported to the control room for continuous process monitoring.

### 3.3.5 Industry-Standard Failure Frequencies for Pressure Vessels

Industry-standard failure frequencies for pressure vessels (including high-pressure separators) are available from European sources such as the Flemish *Handbook on Failure Frequencies*. Table 1 below summarizes the generic annual failure frequencies per vessel for small, medium, and large leaks and for full rupture. The "Per 17 vessels" column is simply 17 times the per-vessel frequency, giving the cumulative rate for a system of 17 identical pressure vessels (assuming independent failures).

*Table 3.3 Flemish Handbook on Failure Frequencies for pressure vessels*

| **Failure Mode** | **Frequency per Vessel (yr^-1)** | **Frequency for 17 Vessels (yr^-1)** |
|---|---|---|
| **Small leak** (≤10 mm hole) | 1.2×10^-5 | 2.04×10^-4 |
| **Medium leak** (10–50 mm) | 1.1×10^-6 | 1.87×10^-5 |
| **Large leak** (>50 mm) | 1.1×10^-6 | 1.87×10^-5 |
| **Full rupture** (instantaneous) | 3.2×10^-6 | 5.44×10^-5 |

The AGTU facility comprises 17 pressure vessels. Based on the above data, each vessel has an annual failure frequency of 1.2×10^-5 for a small leak, 1.1×10^-6 for a medium leak, 1.1×10^-6 for a large leak, and 3.2×10^-6 for a full rupture. Thus, the combined annual frequencies for all 17 vessels are:

- **Small leak:** 17 × (1.2×10^-5) = 2.04×10^-4
- **Medium leak:** 17 × (1.1×10^-6) = 1.87×10^-5
- **Large leak:** 17 × (1.1×10^-6) = 1.87×10^-5
- **Full rupture :** 17 × (3.2×10^-6) = 5.44×10^-5

### 3.3.6 Frequency of Occurrence of Consequences:

The consequence tree shown below provides the occurrence frequencies of each consequence related to the event. These occurrence frequencies are based on the leak frequencies and on the probabilities of ignition and explosion. These probabilities of ignition and explosion depend on the nature of the released product and the mass of product involved in the failure.

For the treatment of scenarios derived from the consequence tree, several steps are followed:

First, the various scenarios leading to similar events, such as Jet Fires, are identified within the tree. When multiple branches result in the same type of final event, it is technically justified to sum their occurrence frequencies, provided that they are mutually exclusive.

Next, even if the events belong to the same category (e.g., Jet Fire), their physical effects may differ depending on the size of the leak, the released flow rate, or operating conditions. This is why the evaluation of effects (for example using PHAST) must be carried out separately for each scenario, in order to accurately model thermal or overpressure impacts.

*Table 3.4 Estimated Frequencies of Final Event Outcomes*

| Final event | Frequency |
|---|---|
| Jet fire | 7.32E -06 |
| Flash fire | 1.33E -06 |
| UVCE | 3.2E -07 |
| Pool fire | 2.24E-06 |
| Fire ball | 2.24E-06 |

| | |
|---|---|
| Dispersion | 5.43E -06 |

All the phenomena have an occurrence frequency above the acceptability threshold **(10^-6)** expect UVCE; it is therefore necessary to analyze their consequences.

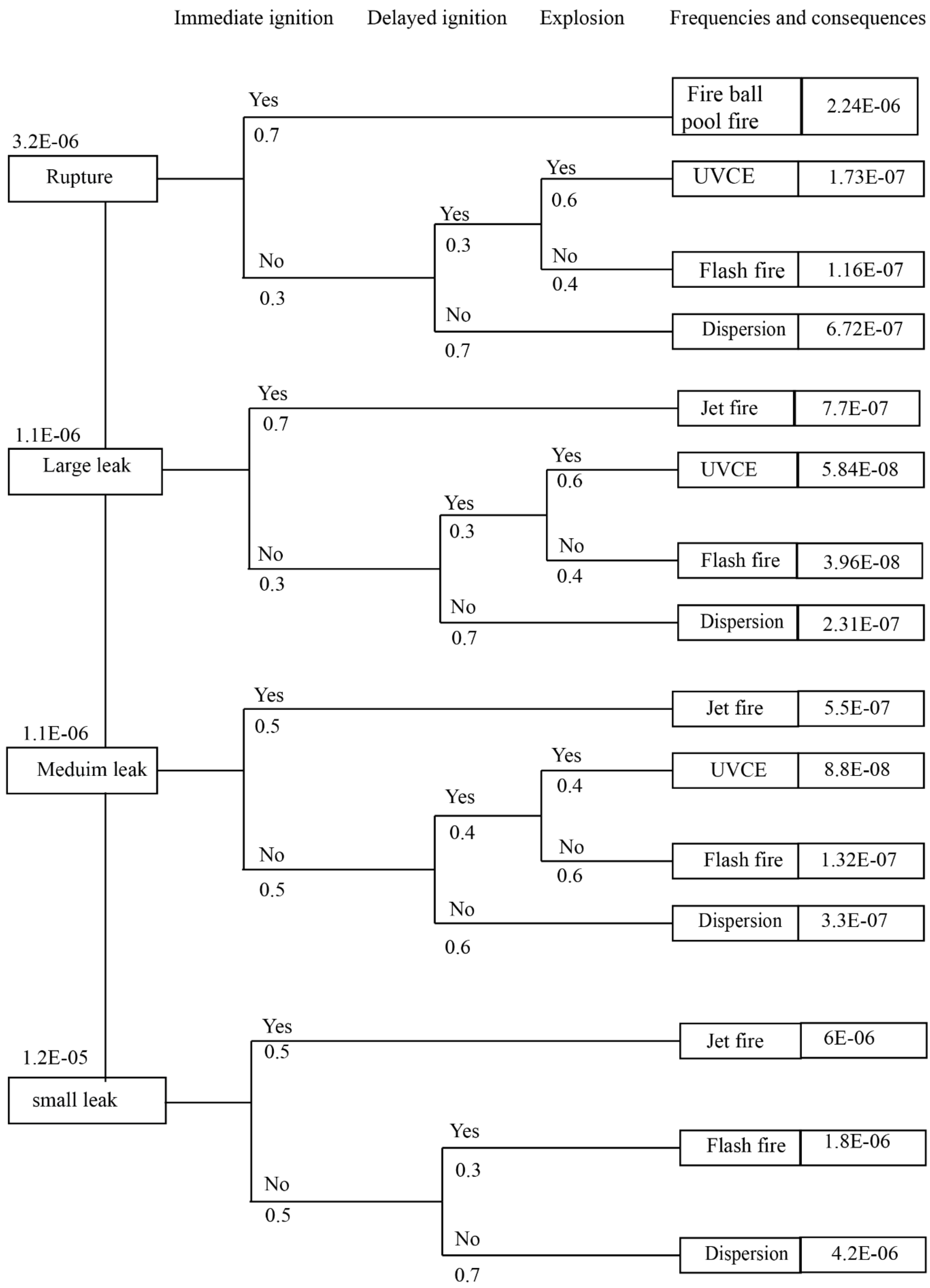


*Figure 3-1 1consequence tree analysis and  frequencies calculation*

## 3.5 Simulation and modulization:

### 3.5.1 PHAST Simulation Overview

As part of the quantitative risk assessment (QRA) for the **Associated Gas Treatment Unit** at the **Tin Fouyé Tabenkort (TFT)** facility, a detailed consequence modeling study was conducted using **PHAST (Process Hazard Analysis Software Tool)** developed by DNV. The focus of the analysis was the **D-401 gas condensate separator**, a pressure vessel operating under the following conditions:

- **Operating Pressure :** 80 bar
- **Operating Temperature:** 20 °C
- **Total Volume :** 15 $m^3$

The objective was to assess the potential consequences associated with accidental loss of containment scenarios involving condensate and gas, to aid in the identification and prioritization of risk mitigation measures.

## 3.6 Release Scenarios

Four release scenarios were considered, reflecting varying degrees of mechanical failure:

- **Full Rupture**
- **Small Leak** (Orifice diameter: 6 mm)
- **Medium Leak** (Orifice diameter: 25 mm)
- **Large Leak** (Orifice diameter: 100 mm)

Each scenario was modeled to evaluate the effects of:

- Gas dispersion and flammable cloud extent.
- Thermal radiation from jet fires and pool fires.
- Flash fire hazards.
- Fireball events (where applicable).

### 3.6.1 Meteorological Conditions

To account for seasonal variability in atmospheric dispersion and fire behavior, simulations were conducted under four distinct weather cases, summarized in the table below:

*Table 3.5Meteorological Conditions Used for Scenario-Based Dispersion and Fire Modeling*

| Scenario | Wind Speed (m/s) | Stability Class | Ambient Temperature (°C) | Humidity | Solar Radiation (kW/m²) |
|---|---|---|---|---|---|
| Summer Day | 6 | C | 45 | 0.2 | 1.2 |
| Summer Night | 3 | D | 20 | 0.3 | – |
| Winter Day | 4 | B | 20 | 0.3 | 1.2 |
| Winter Night | 2 | F | 0 | 0.6 | – |

These atmospheric parameters were selected to represent conservative and realistic conditions for **dispersion, ignition probability, and thermal radiation propagation**.

**Purpose and Outcomes**

The main objective of the simulation campaign is to:

- Quantify the **consequence distances** for flammable effects under varying meteorological conditions;
- Evaluate the **influence of atmospheric stability and temperature** on dispersion and thermal radiation zones;
- Provide a scientific basis for recommending **risk reduction measures**, optimizing **protective barriers**, and refining the **emergency response plan**.

### 3.6.2 Result and interpretation

**6mm continuous small leak**

In the case of a 6 mm leak, the release behaves as a high-velocity jet of cold gas rather than forming a liquid pool. With a mass flow rate of approximately 0.58 kg/s and an exit temperature of −130 °C, the gas reaches its sonic speed limit (~432 m/s), indicating a choked flow regime where downstream conditions do not affect the flow rate. Under typical daytime atmospheric conditions (Pasquill Stability Class D with wind speeds of 3–5 m/s), turbulent mixing is strong enough to dilute the flammable gas quickly, bringing concentrations below the Lower Flammable Limit (LFL) within about 8 meters of the release point. At night, or under stable meteorological conditions (Class E–F, winds <1 m/s), vertical mixing is suppressed, and the LFL cloud can extend up to 12 meters.

If ignition is delayed until the vapor cloud has spread, a **flash fire** may result. During the day, ignition could occur within about 2 seconds; at night, the ignition delay could stretch to as long as 7.5 seconds. The flash fire would be confined to the flammable envelope—up to 12 meters from the source—posing localized but serious risks to personnel and equipment.

In the case of immediate ignition, the gas burns as a **jet fire**, forming a narrow, high-velocity flame approximately 10 meters long. Thermal radiation levels reach 4 kW/m² up to 12 meters and 12.5 kW/m² up to 8 meters, representing the thresholds for light injury and fatality, respectively. These distances define a compact but hazardous thermal exclusion zone around the leak.

Due to the vapor-only nature of the release and the low flammable mass involved, the formation of a **pool fire** or **fireball** is not possible. Additionally, the small quantity of gas and its rapid dilution under open-air conditions make a **vapor cloud explosion (UVCE)** highly unlikely.

*Table 3.6 Consequence Modeling Results for a 6 mm Small Leak Scenario*

| Phenomenon | Key Metrics & Effect Distances |
|---|---|
| **Dispersion** | LFL cloud radius: 8 m (day), 12 m (night) |
| **Jet Fire** | Length: 10 m 4 kW/m²: 12 m 12.5 kW/m²: 8 m |
| **Flash Fire** | Max LFL radius: 9 mIgnition delay: 2 s (day) / 7.5 s (night) |
| **Pool Fire** | Not applicable |

| | |
|---|---|
| **Fireball** | Not generated |
| **VCE Overpressure** | No sustained vapor cloud |


*Figure 3-2Jet fire radiation intensity contours for a small leak scenario*

**25mm continuous medium leak**

When the orifice diameter increases to 25 mm, the transient gas release becomes significantly more intense, reaching a flow rate of approximately 7.3 kg/s and lasting around 16 seconds. This results in the formation of a substantial vapor cloud. Under well-mixed daytime conditions (Pasquill Stability Class B–D with wind speeds of 4–6 m/s), the lower flammable limit (LFL) boundary extends to about 110 meters. However, during nighttime or winter-morning conditions—characterized by atmospheric stability (Class E–F) and low wind speeds (<1–2 m/s)—dispersion is severely limited. Under these calm and stable conditions, the flammable cloud can persist and expand to nearly 130 meters, increasing the likelihood of ignition from hot surfaces, static discharge, or electrical arcing.

If ignition is delayed, the resulting **flash fire** could propagate across the entire flammable cloud, up to 130 meters in radius. This presents a widespread hazard to personnel and equipment, especially under low-dispersion conditions. The threat zone extends far beyond the immediate release area, and ignition timing becomes a critical factor in determining the severity of the event.

In the case of immediate ignition at the release point, the escaping gas burns as a **jet fire**. Thermal radiation modeling indicates peak intensities of 90–110 kW/m² within 11–13 meters of the jet. These levels are capable of causing fatal injuries to unprotected personnel in seconds and would rapidly damage exposed equipment. Even the lower radiation threshold of 4 kW/m², considered hazardous for prolonged exposure, extends up to 60 meters under calm conditions—requiring clearly defined exclusion zones around the leak.

Due to the vapor-only nature of the release, **pool fire** and **fireball** formation are not applicable. There is no liquid inventory or rapid-phase transition to support either phenomenon. However, the size and mass of the vapor cloud create a credible risk of a **vapor cloud explosion (UVCE)**. Should ignition occur while the cloud is unconfined, overpressures can exceed 18 bars at a distance of 50 meters—enough to buckle steel structures and destroy process equipment. Even at 100 meters, overpressures of 0.02 bar are possible, sufficient to break glass and damage lightly built structures.

*Table 3.7 Consequence Modeling Summary for a 25 mm Medium Leak Scenario*

| **Phenomenon** | **Key Metrics & Effect Distances** |
|---|---|
| **Dispersion** | LFL: 110 m (day) / 130 m (night) |
| **Jet Fire** | Peak radiation: 90–110 kW/m² at 11–13 m4 kW/m²: 60 m |
| **Flash Fire** | Max LFL radius: 130 m |
| **Pool Fire** | Not applicable |
| **Fireball** | Not generated |
| **VCE Overpressure** | >18 bar at 50 m 0.02 bar at 100 m |


*Figure 3-3 Flammable gas cloud footprint for a medium leak scenario*


*Figure 3-4 Jet fire radiation intensity contours for a meduim leak scenario*


*Figure 3-5 Flash fire envelope for a medium leak scenario*


*Figure 3-6 Explosion overpressure contours for the worst-case scenario*

**100mm continuous catastrophic leak**

A breach of 100 mm in diameter results in a **large release of gas**, with a mass flow of approximately 160 kg/s sustained over 13 seconds. This immense release forms a significant flammable vapor cloud, extending to around 120 meters under daytime neutral conditions (Pasquill Class D, wind ~6 m/s), and up to 133 meters under winter-night stable conditions (Class F, wind <1 m/s). Under such stable atmospheric conditions, portions of the flammable cloud may persist and drift up to 330 meters downwind, greatly increasing the time and area in which ignition could occur. If ignition is delayed, a **flash fire** could ignite the entire cloud, potentially affecting areas well beyond 130 meters from the source.

In the event of immediate ignition, a **jet fire** forms with an exceptionally long flame, exceeding 200 meters. This poses a major threat to nearby units. Radiation modeling shows that the 4 kW/m² threshold—hazardous to personnel—can reach between 198 and 210 meters, while the lethal 12.5 kW/m² threshold extends to approximately 150 meters. These zones represent critical areas where human life and equipment integrity are at serious risk.

Although no **pool fire** is possible due to the gas-phase nature of the release, conditions are suitable for the formation of a **fireball** if the cloud ignites nearly instantaneously. The fireball is estimated to reach a diameter of 75 meters, with thermal radiation effects spreading to 309 meters at 4 kW/m² and up to 178 meters at the lethal 12.5 kW/m² level. These distances define a broad thermal exclusion zone.

The scenario also presents a major **UVCE (Unconfined Vapor Cloud Explosion)** risk. Should ignition occur before the cloud disperses, the resulting overpressure could exceed

0.2 bar as far as 292 meters, sufficient to damage steel structures and cause widespread mechanical failure. Even at 504 meters, overpressure could still reach 0.02 bar—enough to shatter windows and damage lightly built structures. These values reinforce the necessity of planning for blast effects across a half-kilometer radius.

*Table 3.8 Consequence Modeling Summary for a 100 mm Large Leak Scenario*

| Phenomenon | Key Metrics & Effect Distances |
|---|---|
| **Dispersion** | LFL: 120 m (day) / 133 m (night)Flammable fraction reach: 330 m |
| **Jet Fire** | Length: >200 m 4 kW/m²: 198–210 m 12.5 kW/m²: 150 m |
| **Flash Fire** | Max LFL radius: 133 m |
| **Pool Fire** | Not applicable |
| **Fireball** | Diameter: 75 m4 kW/m²: 286–309 m12.5 kW/m²: 166–178 m |
| **VCE Overpressure** | 0.02 bar at 504 m0.2 bar at 292 m |

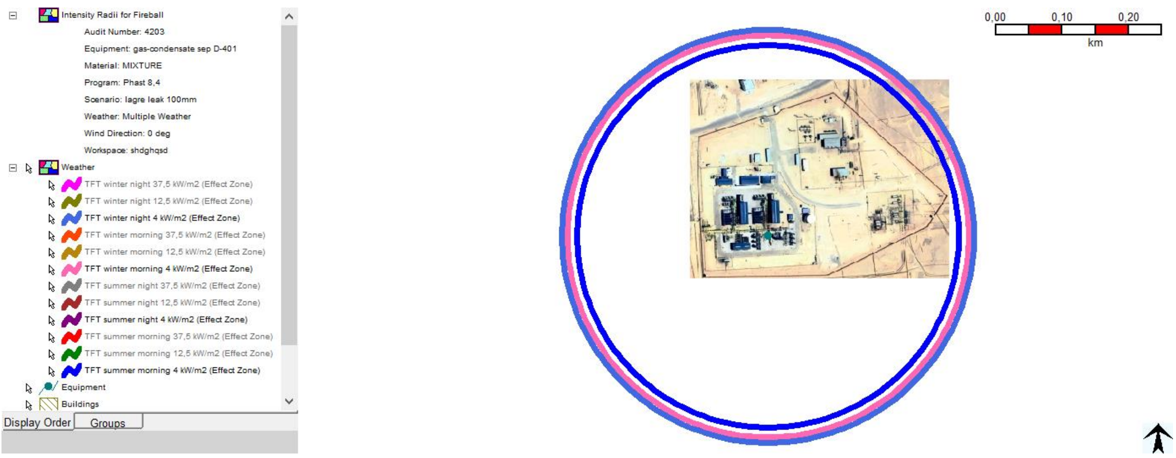


*Figure 3-7 Thermal radiation intensity contours for a fireball resulting from a large leak scenario*

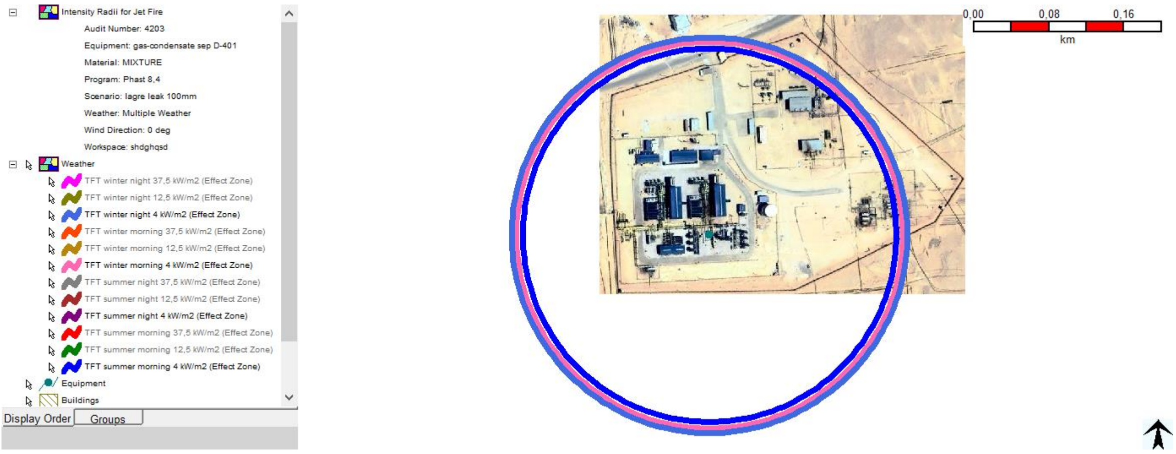


*Figure 3-8 Jet fire radiation intensity contours for a large leak scenario*

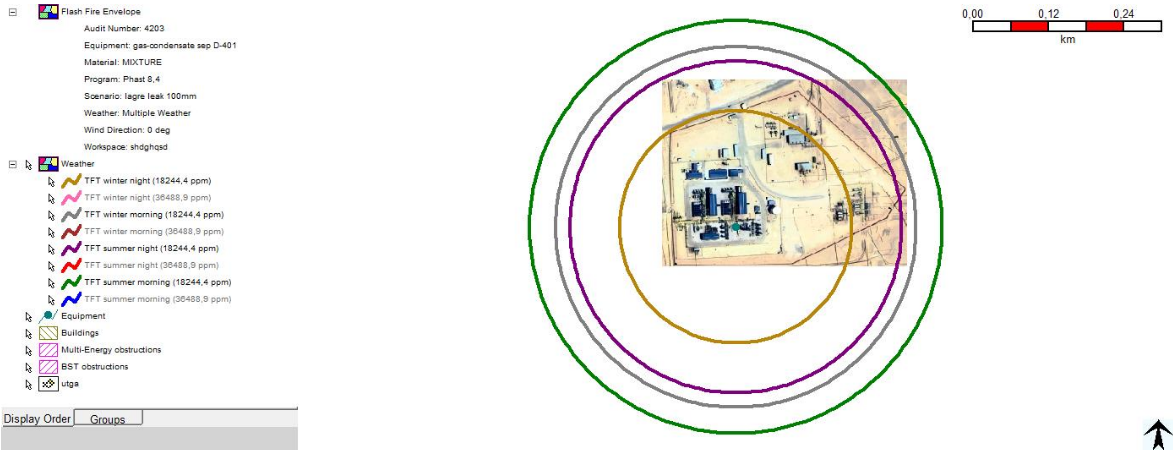


*Figure 3-9 Flash fire envelope for a large leak scenario*

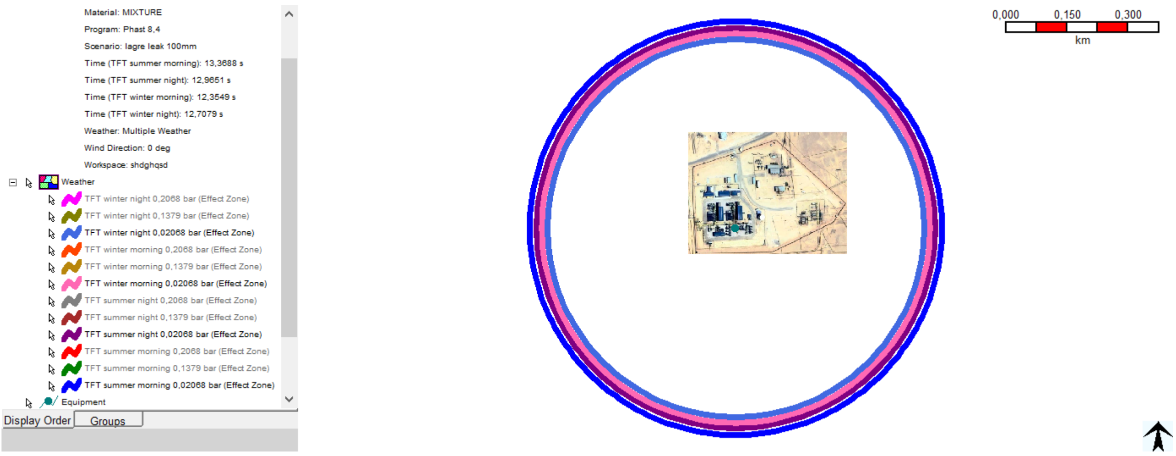


*Figure 3-10 Explosion overpressure contours for the large leak scenario*

**Full rupture catastrophic failure**

A full-bore rupture of the D-401 separator results in the violent discharge of a high-momentum, two-phase mixture at a velocity of approximately 121 m/s. The release contains around 7.8% liquid droplets, making it significantly more hazardous than purely gaseous

leaks. The flammable vapor cloud that forms near the source extends up to 22 meters to the Lower Flammable Limit (LFL) and 10 meters to the Upper Flammable Limit (UFL) under well-mixed daytime conditions (Pasquill Class C, 4 m/s wind). However, in stable atmospheric conditions typical of nighttime (Class E–F, wind <1 m/s), the cloud remains dense and stratified, persisting longer and staying closer to ground level, where ignition sources are more likely to be present.

If ignition is delayed by 5–7 seconds, a **flash fire** could propagate through the vapor cloud, reaching up to 49 meters from the point of release. This delay provides only a short response window and emphasizes the importance of rapid detection and isolation systems. Unlike previous scenarios, a **jet fire** is not the dominant hazard here due to the presence of a liquid phase, which rapidly forms a pool around the release point.

This liquid immediately generates a **pool fire** with a diameter of about 3.5 meters. The thermal radiation impact is severe: 4 kW/m² reaches up to 14 meters, 12.5 kW/m² extends to 9.3 meters, and extreme radiation of 37.5 kW/m² is present within 3.9 meters—intense enough to cause third-degree burns in seconds and compromise structural integrity of nearby steel elements. These thermal zones represent a critical immediate hazard around the release site.

If the flammable cloud ignites nearly instantaneously, a **fireball** similar in scale to that of the 100 mm large leak will form. The estimated diameter is approximately 75 meters, with the 4 kW/m² thermal radiation level extending to 309 meters and the 12.5 kW/m² lethal threshold reaching up to 178 meters. These distances define a large thermal exclusion zone that must be factored into site layout and emergency response planning.

The potential for a **vapor cloud explosion (UVCE)** in this scenario is both plausible and severe. A VCE could produce overpressures exceeding 0.2 bar up to approximately 101 meters and 0.02 bar (enough to break windows and damage light structures) out to around 452 meters. These blast effects extend far beyond the immediate vicinity of the release and demand robust structural protections and wide evacuation perimeters.

*Table 3.9 Consequence Modeling Results for a Full Catastrophic Rupture Scenario*

| **Phenomenon** | **Key Metrics & Effect Distances** |
|---|---|
| **Dispersion** | LFL/UFL: 22 m / 10 m |
| **Jet Fire** | Not applicable (pool fire dominates) |
| **Pool Fire** | 4 kW/m²: 14 m12.5 kW/m²: 9.3 m37.5 kW/m²: 3.9 m |
| **Flash Fire** | Max distance: 49 m Ignition delay: 5–7 s |
| **Fireball** | Diameter: 75 m 4 kW/m²: 309 m 12.5 kW/m²: 178 m |
| **VCE Overpressure** | 0.02 bar at 452 m0.2 bar at 101.5 m |

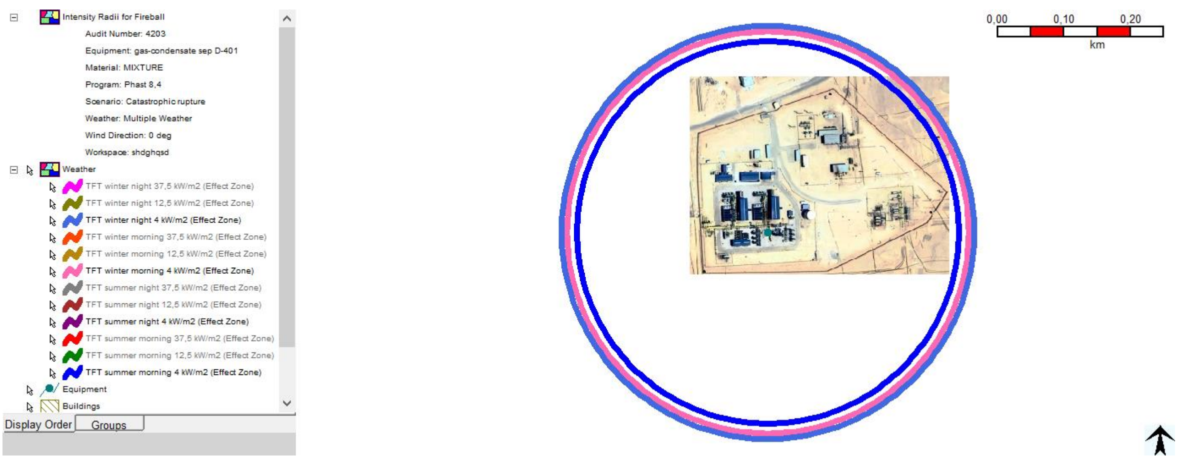


*Figure 3-11 Thermal radiation intensity contours of the fireball resulting from a full rupture scenario*


*Figure 3-12 Flash fire envelope for the full rupture scenario*

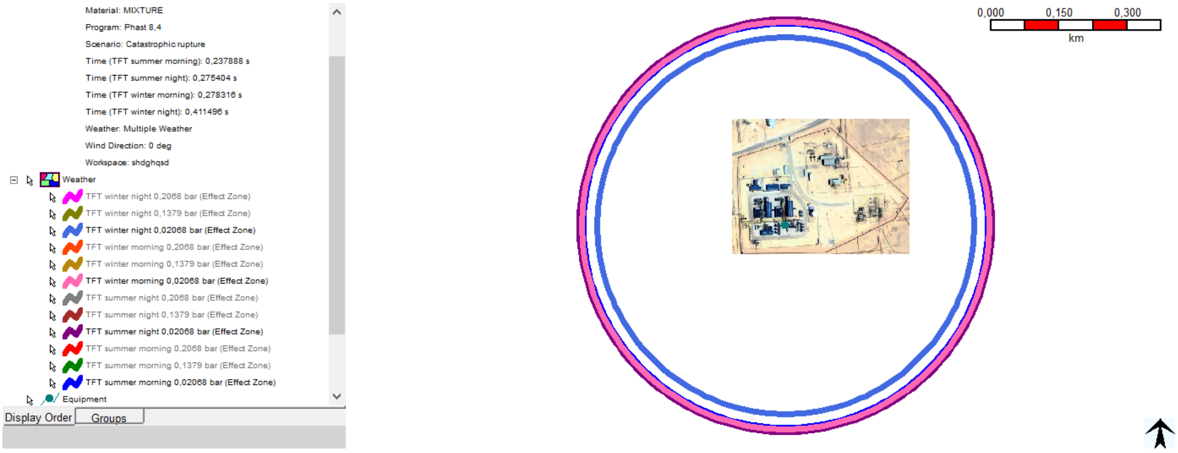


*Figure 3-13 Explosion overpressure contours for the worst-case scenario*

### 3.6.3 The influence of the atmospheric parameters:

Atmospheric conditions play a fundamental role in determining the dispersion behavior and hazard potential of flammable or toxic releases. Among the most influential factors are atmospheric stability and wind speed, which directly govern how vapors spread, dilute, and persist in the environment following a leak or rupture scenario.

Atmospheric stability, classified by the Pasquill-Gifford system into categories A through F, defines the level of vertical turbulence and mixing in the atmosphere.

Under stable atmospheric conditions—typically Classes E and F, which are common at night or under overcast skies with little solar heating—the vertical temperature gradient suppresses turbulence. This limits vertical mixing and causes hazardous gases or vapors to remain concentrated near ground level. In such scenarios, the vertical dispersion of the release is minimal, and the hazardous plume becomes elongated, dense, and slow to dissipate. These effects result in vapor clouds that are larger in horizontal extent and longer in duration, posing a greater risk of prolonged exposure and delayed ignition.

In contrast, unstable or neutral atmospheric conditions—typically Classes B through D—occur during daytime or when solar heating generates thermal convection. These conditions promote vertical mixing and increase atmospheric turbulence, which in turn enhances the dilution of released gases. As a result, vapor plumes disperse more rapidly, concentrations fall more quickly below hazardous thresholds, and the footprint of the hazard is significantly reduced. This increased dispersion limits the formation of extensive flammable clouds and thereby reduces the probability and severity of ignition-related events such as flash fires or vapor cloud explosions. Wind speed is another key factor that modulates dispersion. Low wind speeds, especially below 1–2 meters per second, result in slower horizontal advection of the plume, allowing vapor concentrations to build up near the release point. In the presence of stable conditions, this combination is particularly hazardous because it traps flammable vapors close to the ground and permits the accumulation of large cloud volumes at concentrations above the lower flammability limit (LFL). These dense, ground-hugging clouds significantly increase the potential for delayed ignition, particularly if they travel into congested areas where ignition sources are likely to be present. Once ignited, such clouds can result in widespread flash fires with a broad thermal impact or, if confined or semi-confined, transition into unconfined vapor cloud explosions (UVCEs) that produce destructive overpressure waves.

Moderate wind speeds, generally in the range of 3–6 meters per second, contribute to more effective dispersion by increasing mechanical turbulence and transporting the plume away from the release point more rapidly. This results in lower downwind concentrations and shorter residence times for hazardous vapors. The likelihood of forming large, flammable clouds is greatly diminished, and the overall consequence severity of any subsequent ignition event is correspondingly reduced.

From a modeling perspective, the behavior of hazardous plumes can be described using Gaussian dispersion theory. In this framework, dispersion parameters such as the lateral and vertical spread of the plume (σy\sigma_yσy and σz\sigma_zσz, respectively) are strongly dependent on both atmospheric stability and distance from the source. Under stable conditions, $\sigma z$ —the vertical dispersion term—increases very slowly with distance, indicating minimal vertical spread. Consequently, the ground-level concentration $Cmax$ , which is inversely proportional to wind speed and affected by turbulence, remains elevated over a longer distance and time.

These physical behaviors have direct implications for consequence modeling in risk assessments. Under worst-case meteorological conditions—defined by low wind speeds and stable atmospheres—the dispersion of flammable or toxic releases is significantly limited, while the persistence and ground-level concentration of the vapor cloud are greatly increased. This leads to larger hazard footprints for thermal radiation, explosion overpressure, and toxic exposure. Accordingly, emergency planning and safety design must adopt conservative assumptions based on these worst-case conditions to ensure that mitigation strategies and response measures are robust and effective under the most challenging circumstances.

### 3.6.4 Risk level and recommendations

#### Risk level

*Table 3.10Probability Classification and Frequency Definitions for Risk Assessment*

| **Probability** | **Description** | **Frequency** | **Definition** |
|---|---|---|---|
| **P1** | Improbable | $< 10^{-4}$ / year | Never encountered or heard of, but physically possible (extremely rare). |
| **P2** | Unlikely | $10^{-4}$ to $10^{-2}$ / year | Has occurred (or could occur) in an organization similar to Sonatrach. |
| **P3** | Likely | $10^{-2}$ to $10^{-1}$ / year | Has occurred (or could occur) within Sonatrach, possibly during the facility's lifetime. |
| **P4** | Very Likely | 1 / year | Has occurred frequently within Sonatrach. |

*Table 3.11 Severity Classification Criteria for Risk Assessment*

| **Severity** | **Personnel** | **Environment** | **Public** | **Assets** |
|---|---|---|---|---|
| **S1** | Minor injuries (First aid – ASA) | Minor | No impact | No damage, no production shutdown |
| **S2** | Significant injuries (Medical treatment – AAA) | Controlled internal pollution | Minor injuries | Minor damage and brief production halt |
| **S3** | Permanent disability or 1 | Uncontrolled internal pollution or managed | Significant injuries | Localized damage and |

| | | | | |
|---|---|---|---|---|
| | fatality | off-site pollution | | partial unit shutdown |
| **S4** | Multiple fatalities | Long-term off-site pollution exceeding limits | Fatalities | Major damage and total shutdown |

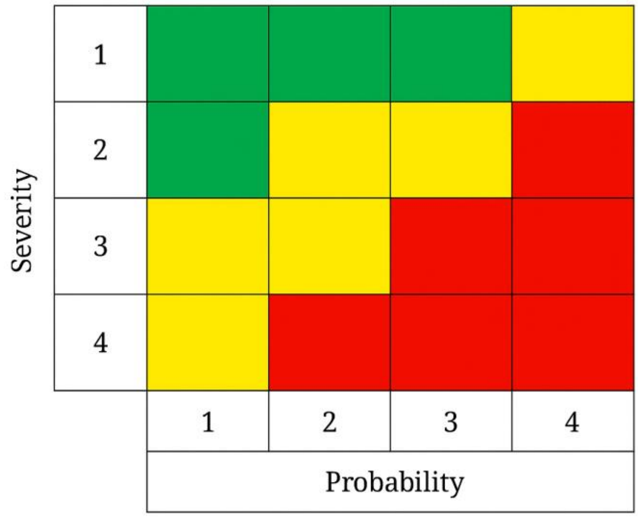


*Figure 3-14 Risk matrix*


*Figure 3-15 risk classification*

*Table 3.12 Risk Assessment Summary of Final Events Based on Probability and Severity*

| N° | Event | Probability | Severity | Risk level |
|---|---|---|---|---|
| **1** | Pool fire | P1 | S3 | ALARP |
| **2** | Fire ball | P1 | S4 | ALARP |
| **3** | UVCE | P1 | S4 | ALARP |
| **4** | Jet fire | P1 | S3 | ALARP |
| **5** | Flash fire | P1 | S3 | ALARP |

## 3.7 Recommendations:

Gas condensate separators handling high‑pressure hydrocarbons pose several fire/explosion hazards (jet fires, pool/flash fires, UVCE, fireballs) even if personnel exposure is nominally low. Further mitigation can be achieved via layered controls. Below we categorize practical measures under **Engineering Controls**, **Detection & Suppression**, **Operational/Procedural**, and **Design Modifications**. Each recommendation is drawn from industry best practices and guidance.

### 3.7.1 Engineering Controls

**Passive Fire Protection (PFP):** Apply fireproof coatings or cladding (e.g. intumescent paint, cementitious wraps) to the separator vessel and connected piping. PFP delays steel heat-up under radiant jet/pool fire, buying time for depressurization. For example, Lloyd's Register‑approved PFP can provide 30‑120 min of protection for vessel surfaces in a hydrocarbon jet fire.

**Pressure Relief & Blowdown:** Ensure high-capacity relief (PSVs or rupture disks) are sized for worst-case fire exposure. Route flare and blowdown lines so that a fire on the separator vents harmlessly into a flare or safe area. Fire-case depressurization (e.g. to flare) removes flammable inventory and reduces vessel pressure. Scandpower guidelines explicitly recommend increasing flare capacity and providing fast depressurization to limit fire escalation. Automatic high-rate blowdown (triggered by fire detection) is preferred over manual to meet these targets.

**Robust Equipment Design:** Use higher-grade materials and extra-thick-walled vessels/piping incritical sections. Where feasible, upsizing wall thickness or using higher

Alloys will lengthen time to-failure in fire. Provide double block ‐ and ‐ bleed or redundant isolation valves on inlet/outletlines to limit any release size.

**Bunds and Drainage:** Install containment (bund walls, berms, or trenches) under the separator to catch any liquid condensate spills. Proper drainage piping should carry leaked condensate to a remote safe drain area. IoMosaic notes that *"if [condensate] material is drained to a safe location, a pool fire is not possible"*. Bunds should be far from ignition sources and equipped for firewater drainage. Cover or sump drains to minimize vapor release. In short, dedicated impoundment (bunds/pits) under condensate drains and leaks prevents accumulation of combustible pools.

**Shielding/Barriers:** Where geometry allows, install non-combustible deflectors or steel panels to shield surrounding equipment from a likely fire impingement zone. Local **blast walls** or offsets can help protect critical nearby items. (For example, if the separator is on an open rack, a steel guard wall on the jet-fire side can reduce thermal exposure to adjacent vessels.)

**Insulation & Lagging:** Insulate hot surfaces (when applicable) and ensure electrical conduit, junction boxes, and cable trays are rated for Class I, Division 2 environments (or similar), to prevent electrical ignition sources near the separator.

### 3.7.2 Detection & Suppression Systems

**Hydrocarbon Gas Detectors:** Install fixed gas sensors (e.g. catalytic or infrared detectors) around the separator. These can alert operators to a leak before ignition. As noted by industry, *«gas detectors detect gas clouds of sufficient size that, if ignited, could produce a flash fire or explosion overpressure"*. Placement should cover likely leak points (flanges, valves, vents) and high points (for lighter-than-air vapors). Integrate detectors into the Emergency Shutdown (ESD) logic. On high alarm, detectors should automatically trip isolation valves or trigger blowdown to rapidly reduce inventory. Automated gas detection plus shutdown has been shown to be a highly effective barrier for all fire scenarios.

**Flame and Smoke Detectors:** Use IR/UV flame detectors aimed at the separator to catch any combustion onset (jet or pool fire). Heat detectors or aspirating smoke detectors inside any shelters can catch concealed fires. Early flame detection enables prompt activation of firewater systems.

**Emergency Shutdown (ESD) System:** A reliable ESD should isolate feed lines and initiate blowdown to flare on alarm. Ensure interlocks so that fire or gas alarms immediately close Emergency Shutdown Valves (ESDVs) and open blowdown vents. A fully integrated ESD + blowdown design removes flammable contents and minimizes fire size. Regular testing of these systems (e.g. functional tests of ESDVs) is critical.

**Water Deluge / Monitors:** Provide a fixed water deluge or monitor system capable of wetting the separator and nearby equipment. Deluge sprays (general area or dedicated vessel deluge) will **cool equipment and dramatically reduce thermal radiation** to surroundings. Although high-velocity jet flames are difficult to extinguish with water (they entrain air), area deluge can cut off 60–80% of heat radiation when properly designed. Position monitors to cover the most exposed face of the separator. Use fine-water spray nozzles (generating small droplets) and design flows for ≥18–24 L/min· $m^2$ over the target area (per Oil&Gas UK guidance). Regularly inspect and flow-test the deluge headers and nozzles.

**Foam Systems:** For pool-fire risk on condensing liquids, consider an AFFF/AR-AFFF foam station. If spilled condensate can accumulate (e.g. in a bund), a foam discharge can form a vapor suppressing blanket. Foam is a well-established defense against oil/condensate pool fires; it will not halt a gaseous jet fire but can quickly contain and extinguish a liquid fire.

Foam hardware (mixers, skimmers, monitors) should be maintained and tested per NFPA/IP standards.

**Firewater and Monitors:** Ensure ample firewater supply and hydrants within range of the separator. Portable monitors (like high-capacity nozzles) give responders flexibility to aim water or foam at leak sources. Maintain a mapped hydrant layout and pressure (per NFPA 20/25or local codes).

### 3.7.3 Operational & Procedural Controls

**Hot Work / Permit-to-Work (PTW):** Enforce strict PTW controls for any welding, burning, or spark-producing activity near the separator. Before and during hot work, ensure gas-free conditions (e.g. via gas tests) and have fire watch personnel with extinguishers at the ready. Hot work permits are a low-cost but critical measure to prevent ignition of leaks.

**Routine Inspection & Maintenance:** Schedule frequent inspections of all separator components (vessels, piping, valves). Critical items include PSV relief devices (bench-test at intervals recommended by API 576), level controls, and flare line integrity. Inspect and maintain seals, gaskets, and valves to prevent leakage. Non-destructive testing (e.g. UT for corrosion) on the vessel body can catch thinning before leak-out. Also inspect firewater systems, detectors, and controls on a regular basis (monthly or per company standard). Record-keeping of maintenance assures all barriers remain functional.

**Leak Detection Walkdowns:** Conduct regular plant walkdowns (or use sniffers/CCTV) to visually check for signs of hydrocarbon accumulation, corrosion, or minor leaks. Early minor-leak fixes are far cheaper than managing a full-blown fire event.

**Emergency Response Drills:** Even if no one normally occupies the separator area, training nearby personnel on responding to gas/alarm events is prudent. Test the control-room alarm logic and emergency actions (e.g. ESD actuation, remote deluge firing). Simulated leak/fire drills help verify procedures and equipment readiness.

**Procedure for Blowdown:** Define clear procedures for depressurizing the separator (e.g. to the flare or recovery system) during maintenance or abnormal situations. Keeping the vessel as empty as practicable (and at low pressure) whenever possible minimizes available fuel. Automatic blowdown interlocks (on high temperature or fire alarm) can further reduce manual error.

**Housekeeping and Ignition Control:** Maintain a clean area free of extraneous hydrocarbons (e.g. no pooled oil, no trash). Prohibit smoking and unapproved electrical equipment in the vicinity. Use intrinsically safe or explosion-proof lighting/fans near the separator. Ground (bond) the vessel and transfer hoses to prevent static sparks during drainage or sampling.

### 3.7.4 Design Modifications

**Layout & Spacing:** Optimize the plant layout to minimize hazard overlap. Maintain safe separation between the separator and occupied buildings or critical equipment. Use API/IP guidelines for flare, gas/vapor dispersion, and thermal-radiation distances. For example, locate the separator downwind of the control room or instrument air compressors. Ensure undisturbed air flow around the vessel so any leaked gas disperses upward. In general, layout should allow clear firewater coverage and safe egress routes.

**Segregation of Streams:** Where possible, separate condensate and gas streams so that a leak in one line does not involve all inventory. For example, a separate pipeline to the condensate pump suction (with its own PSV) can localize leaks. Double block valves in series with a bleed point add margin.

**Vent/Flare Routing:** Pipe relief and drain lines away from personnel areas. If not flaring, use stacks with flame arrestors or burners at a safe height. Route all vents to sheltered or fenced flares. A well-designed flare knockout system upstream of the stack prevents liquid carryover (which can cause fireballs). Ensuring that vented gas goes to high-discharge flares reduces the risk of on-site ignition of a large cloud.

**Drain and Bund Design:** Engineer the separator support structure so that any floor drains slope into a single, contained sump or pit (with firewater drainage). The drainage area should be sized to hold the maximum credible liquid spill, with valving to control release. The IoMosaic reference recommends that drains be *"far enough away from thermal radiation fire sources"* and that containment is provided for fire water. Essentially, design the pad so that a leak flows away from the separator into a safe catchment, preventing pool formation.

**Passive Inerting:** If technically and economically feasible, maintain a slight blanket of inert gas (e.g. $N_2$) on top of the liquid condensate in the separator. This is more common on large storage tanks, but small separators can use low-flow nitrogen purges to keep vapor space nonflammable. Inerting removes the oxidizer (oxygen), effectively breaking the fire triangle.

**Advance Instrumentation:** Upgrade to smart instrumentation (with more frequent self-testing or redundancy) for critical safety devices (PSVs, level alarms, pressure transmitters). Use HART or fieldbus diagnostics to monitor any drift/failure, and alarm if a safety instrument is out of service. Though not a "quick fix," increasing instrumentation reliability is an industry-recognized practice.

**Justification & Cost‑Effectiveness:** Many of these measures are proven, low- to moderate-cost improvements. For example, adding gas detectors and tying them into automatic shutdown is relatively inexpensive yet cuts off flammable inventory immediately. Likewise, routine inspection and minor PFP coating are routine engineering tasks. Water deluge systems require capital but can use existing water supplies and cover many vessels at once.

# General conclusion

This study aimed to conduct a comprehensive risk assessment of the gas-condensate separator D-401 within the Associated Gas Treatment Unit (AGTU) at the Tin Fouyé Tabenkort (TFT) oil field. Given the highly hazardous nature of hydrocarbon processing under high pressure and temperature conditions, the separator poses significant risks related to fire, explosion, and toxic releases. To ensure safe operation and protect human life, the environment, and equipment, an integrated approach combining qualitative and quantitative techniques was adopted.

The Hazard and Operability Study (HAZOP) enabled the systematic identification of potential process deviations, their root causes, and associated consequences. It also facilitated the evaluation of existing safeguards and led to the proposal of targeted recommendations for risk reduction. Through this structured methodology, several critical deviations—such as high pressure, overfilling, and valve failures—were identified as major threats requiring improved control and monitoring systems.

Complementing the HAZOP, consequence modeling using PHAST software allowed the quantification of the severity and spatial extent of accidental scenarios. Simulated cases of vessel rupture and varying leak sizes under different meteorological conditions provided detailed insight into potential fire and explosion zones, thermal radiation intensities, and safe separation distances. These simulations confirmed the importance of weather parameters in determining dispersion behavior and hazard reach, emphasizing the need for context-specific emergency planning.

Overall, the integration of HAZOP and PHAST offered a robust and holistic risk assessment framework. The findings underline the critical importance of maintaining equipment integrity, implementing effective detection and shutdown systems, and ensuring periodic risk reviews. This study not only strengthens the safety case for the D-401 separator but also serves as a model for assessing other high-risk process equipment within AGTU and similar facilities.

Through this work, a deeper understanding of process safety management has been achieved, demonstrating the value of combining engineering analysis, simulation tools, and safety culture to prevent major industrial accidents.

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# Appendices

## Hazop tables and ATGU P&ID